

**PRELIMINARY DESIGN OF METHANOL PLANT FROM
CO₂ AND H₂ WITH CAPACITY OF
150,000 TON/YR**

CHEMICAL PLANT DESIGN

Filed as One of the Requirements
To Obtain a Bachelor's Degree in Chemical Engineering



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YOGYAKARTA
2026**

STATEMENT OF ORIGINALITY

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PRELIMINARY DESIGN OF METHANOL PLANT FROM CO₂ AND H₂ WITH CAPACITY OF 150,000 TON/YEAR

We, the undersigned

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Hereby declare that all results of this Plant Design Project are our own original work. Should it be proven in the future that any part of this work is not the result of our own efforts, we are willing to accept all risks and consequences that may arise.

This declaration is made truthfully, with the hope that it will be used appropriately.

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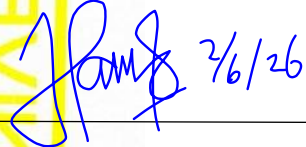
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PREFACE

Assalamu'alaikum Warahmatullahi Wabarakatuh.

"يَرْفَعُ اللَّهُ الَّذِينَ آمَنُوا مِنْكُمْ وَالَّذِينَ أُوتُوا الْعِلْمَ دَرَجَاتٍ ۚ وَاللَّهُ بِمَا تَعْمَلُونَ خَبِيرٌ"

All praise is due to Allah SWT for His mercy, guidance, and blessings, through which this Final Project has been successfully completed. May peace and blessings always be upon our Prophet Muhammad SAW, his companions, and his followers.

This Final Project entitled "Preliminary Design of Manufacturing of Methanol From CO₂ and H₂ with a Capacity of 150,000 Tons/Year" has been prepared as an application of the chemical engineering knowledge acquired during the course of study, and as one of the requirements for obtaining the Bachelor of Chemical Engineering degree at the Faculty of Industrial Technology, Universitas Islam Indonesia, Yogyakarta.

The successful completion of this project would not have been possible without the support and assistance of many individuals. Therefore, the authors would like to express their deepest gratitude to:

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2. Our beloved parents, whose encouragement and unwavering support have continuously motivated us to strive for excellence throughout our studies in Chemical Engineering at UII.
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7. Our fellow friends, for their encouragement, support, and prayers.
8. All individuals whose names may not be mentioned here but have contributed to the completion of this final project.

We acknowledge that this report may still have shortcomings. Therefore, we welcome any constructive criticism and suggestions for its improvement. Finally, we hope that this Final Project report can be beneficial to all readers. Aamiin.

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LIST OF SYMBOLS

ASTM	American Society for Testing and Materials
BEP	Break-Even Point
BFW	Boiler Feed Water
CCUS	Carbon Capture, Utilization, and Storage
CEPCI	Chemical Engineering Plant Cost Index
DCFR	Discounted Cash Flow Rate of Return
DCS	Distributed Control System
DMC	Direct Manufacturing Cost
DPC	Direct Plant Cost
FCI	Fixed Capital Investment
FMC	Fixed Manufacturing Cost
GE	General Expenses
IDLH	Immediately Dangerous to Life and Health
IEA	International Energy Agency
IMC	Indirect Manufacturing Cost
LHHW	Langmuir-Hinshelwood-Hougen-Watson (kinetic model)
MC	Manufacturing Cost
MTBE	Methyl Tert-Butyl Ether
NFPA	National Fire Protection Association
NPV	Net Present Value
PAT	Profit After Tax
PBT	Profit Before Tax
PEC	Purchased Equipment Cost
PEFD	Process Engineering Flow Diagram
PLN	Perusahaan Listrik Negara (State Electricity Company)
POT	Pay Out Time
PPC	Physical Plant Cost
RO	Reverse Osmosis
ROI	Return on Investment
RWGS	Reverse Water-Gas Shift
SDP	Shut-Down Point
SEZ	Special Economic Zone
SIS	Safety Instrumented System
STEL	Short-Term Exposure Limit
TCI	Total Capital Investment
TDS	Total Dissolved Solids
TLV-TWA	Threshold Limit Value - Time-Weighted Average
TPC	Total Production Cost
UEL	Upper Explosive Limit
UPS	Uninterruptible Power Supply
WCI	Working Capital Investment

ABSTRACT

This methanol plant design is motivated by the growing domestic demand for methanol in Indonesia and the country's high import dependency. At the same time, Indonesia faces significant CO₂ emissions from its industrial sector. By utilizing CO₂ as a raw material, this plant simultaneously contributes to reducing greenhouse gas emissions, to achieve Indonesia's net-zero emissions target by 2060. The plant is planned to be located in the Arun Lhokseumawe Special Economic Zone (SEZ), Aceh Province, Indonesia, with a production capacity of 150,000 ton/yr and operating 330 days/yr. The main raw materials of this plant are CO₂ sourced from industrial flue gas and H₂ produced from water electrolysis. The selected process is catalytic CO₂ hydrogenation using a multi tubular fixed bed reactor (Lurgi type) with a Cu/ZnO/Al₂O₃ catalyst. The reactor is operated at 220°C and 50 bar. The methanol product is purified through a distillation column to achieve a purity of 99.7 wt%, meeting AA-grade commercial specifications. The economic evaluation is based on the design year 2030. The results show that the plant requires a Fixed Capital Investment (FCI) of \$88,261,848 and a Total Capital Investment (TCI) of \$106,906,091. The economic feasibility indicators obtained are as follows: Return on Investment (ROI) of 14.70% before tax and 11.47% after tax; Pay Out Time (POT) of 4.1 years before tax and 4.66 years after tax; Break Even Point (BEP) of 40.03% of design capacity; Shut Down Point (SDP) of 13.29%; and Discounted Cash Flow Rate (DCFR)/IRR of 28.12%. All feasibility indicators exceed the minimum required values; therefore, the proposed methanol plant with a capacity of 150,000 tons/year is declared economically feasible and financially attractive for establishment.

Keywords: methanol, CO₂ hydrogenation, Cu/ZnO/Al₂O₃ catalyst, Lurgi reactor, economic feasibility.

ABSTRAK

Desain pabrik metanol ini didorong oleh meningkatnya permintaan domestik metanol di Indonesia dan tingginya ketergantungan impor negara tersebut. Pada saat yang sama, Indonesia menghadapi emisi CO₂ yang signifikan dari sektor industrinya. Dengan memanfaatkan CO₂ sebagai bahan baku, pabrik ini secara bersamaan berkontribusi untuk mengurangi emisi gas rumah kaca, guna mencapai target emisi nol bersih Indonesia pada tahun 2060. Pabrik ini direncanakan akan berlokasi di Kawasan Ekonomi Khusus (KEK) Arun Lhokseumawe, Provinsi Aceh, Indonesia, dengan kapasitas produksi 150,000 ton/tahun dan beroperasi selama 330 hari/tahun. Bahan baku utama pabrik ini adalah CO₂ yang bersumber dari gas buang industri dan H₂ yang dihasilkan dari elektrolisis air. Proses yang dipilih adalah hidrogenasi CO₂ katalitik menggunakan reaktor unggun tetap multi tabung (tipe Lurgi) dengan katalis Cu/ZnO/Al₂O₃. Reaktor dioperasikan pada suhu 220°C dan tekanan 50 bar. Produk metanol dimurnikan melalui kolom distilasi untuk mencapai kemurnian 99,7% berat, memenuhi spesifikasi komersial kelas AA. Evaluasi ekonomi didasarkan pada tahun desain 2030. Hasilnya menunjukkan bahwa pabrik membutuhkan Investasi Modal Tetap (FCI) sebesar \$88.261.848 dan Total Investasi Modal (TCI) sebesar \$106.906.091. Indikator kelayakan ekonomi yang diperoleh adalah sebagai berikut: Pengembalian Investasi (ROI) sebesar 14,70% sebelum pajak dan 11,47% setelah pajak; Waktu Pengembalian Modal (POT) sebesar 4,1 tahun sebelum pajak dan 4,66 tahun setelah pajak; Titik Impas (BEP) sebesar 40,03% dari kapasitas desain; Titik Penutupan (SDP) sebesar 13,29%; dan Tingkat Arus Kas yang Didiskontokan (DCFR)/IRR sebesar 28,12%. Semua indikator kelayakan melebihi nilai minimum yang dibutuhkan; Oleh karena itu, pabrik metanol yang diusulkan dengan kapasitas 150,000 ton/tahun dinyatakan layak secara ekonomi dan menarik secara finansial untuk didirikan.

Kata kunci: metanol, hidrogenasi CO₂, katalis Cu/ZnO/Al₂O₃, reaktor Lurgi, kelayakan ekonomi

CHAPTER I

INTRODUCTION

1.1 Background

The main environmental issue of the 21st century has been the rapid increase in global greenhouse gas emissions caused by transportation, industry, and electricity production. Approximately 76% of the world's greenhouse gas emissions are carbon dioxide (CO₂), making it the most significant contributor to climate change among all greenhouse gas emissions. Carbon capture, utilization, and storage (CCUS) technology has been developed in response to the urgent need to reduce CO₂ emissions and combat global warming (IEA, 2023). Among these approaches, catalytic hydrogenation of CO₂ to methanol — a versatile chemical and energy carrier for industrial applications is one of the most promising processes.

This method is particularly important in Indonesia, which is one of the largest CO₂ emitters in Southeast Asia, with annual emissions reaching 600 million tons, primarily from industrial activities and fuel combustion (IEA, 2023). As a result, there are two advantages to increasing domestic methanol production from CO₂ and green hydrogen. This reduces greenhouse gas emissions, improves energy security, and reduces reliance on imported materials. Furthermore, Indonesia has many renewable energy resources that can be used to create green hydrogen using water electrolysis, especially solar energy, wind, hydroelectric power plants and biomass energy. The conventional Power-to-Methanol pathway can be realized by combining green hydrogen with CO₂ captured from industrial flue gas and biogas sources (IRENA, 2022). This strategy corresponds to the objectives of Indonesia's national energy policy. This includes an increase in

renewable energy share to 23% by 2025 and achieving net-zero greenhouse gas emissions by 2060 (MEMR, 2023).

Methanol (CH_3OH) is considered an essential feedstock for the production of formaldehyde, acetic acid, methyl tert-butyl ether (MTBE), and various polymers and solvents. Methanol is an important component in the transition to a stable energy system. This is because it can be used directly as a pure fuel or mixed with gasoline to reduce emissions (Olah et al., 2018).

The chemical reaction for the production of methanol from CO_2 and H_2 is presented below:



This reaction is catalyzed by a $\text{Cu}/\text{ZnO}/\text{Al}_2\text{O}_3$ catalyst at temperatures of 200–300 °C and pressures of 50–100 bar (Chen et al., 2019).

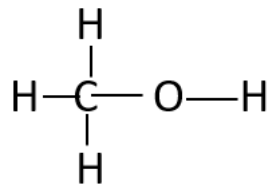


Figure 1.1 Chemical structure of methanol

Given these prospects, a critical answer for Indonesia's energy transition is the Preliminary Design of a Plant for Methanol from CO_2 and Hydrogen. This concept offers a sustainable solution for reducing CO_2 emissions while creating a useful chemical product by fusing environmental advantages with industrial usefulness. With an emphasis on process integration, material and energy balances, and feasibility assessment in Indonesian conditions, this study focuses on the conceptual design and technical evaluation of a CO_2 hydrogenation-based methanol synthesis facility.

1.2 Plant Capacity

Table 1.1 Methanol production company in Indonesia

COMPANY	CAPACITY (TON/YR)
PT. Kaltim Methanol Industri (KMI)	800,000

The list of names of overseas companies that produce methanol along with their production capacity can be seen from *www.ICIS.com* in Table 1.2.

Table 1.2 Methanol production company abroad

COUNTRY	COMPANY	CAPACITY (TON/YR)
USA	Coastal Chemical, Cheyenne	75,000
USA	Air Products, Pensacola	180,000
USA	Motiva Enterprises, Delaware City	300,000
Germany	BASF	480,000
Canada	Methanex, Kitimat	510,000
Germany	Milder-Helm Methanol	660,000
USA	Lyondell, Channelview	750,000
Norway	Statoil	830,000
USA	Beaumont Methanol, Beaumont	840,000
Netherlands	Bio MCN	1,000,000
Russia	Siberian Methanol Chemical	1,000,000

Source: *www.icis.com* (2025)

a. The production capacity of this plant is determined based on a supply–demand analysis of the domestic methanol market. Supply is derived from imports and existing domestic production, while demand reflects total consumption and export volumes.

b. *Supply*

Supply of a product is obtained from imports and domestic production. According to data from the Badan Pusat Statistic, methanol import data in Indonesia shows fluctuating figures but tends to increase every year. Methanol import data in Indonesia from 2021 to 2025 can be seen in Table 1.3.

Table 1.3 Methanol import data 2021 to 2025

NO	YEAR	AMOUNT (TON)
1	2021	979,974
2	2022	959,237
3	2023	984,164
4	2024	1,186,274
5	2025	844,525

(Badan Pusat Statistics,2025)

Based on the information above, Indonesian industry depends on foreign methanol to meet its needs for building materials. The results of the regression analysis of Indonesia's methanol import data may be seen in Figure 1.3, which is a graph showing the relationship between the number of methanol imports and the number of import years.

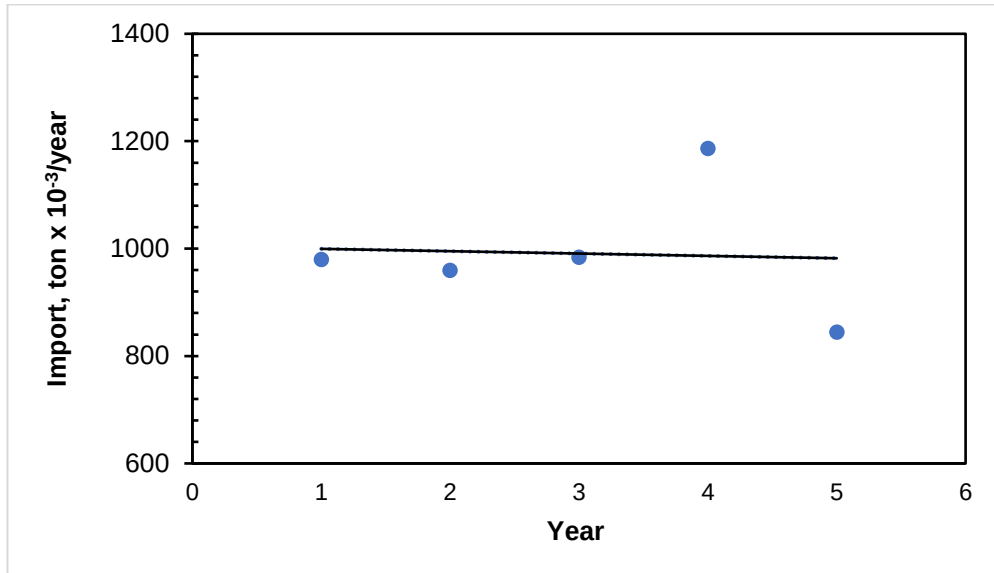


Figure 1.2 Methanol Import Chart 2021 to 2025

The estimated import of methanol in Indonesia in the coming year during the construction of the factory can be calculated using the following equation:

$$y = -4386,3x + 1.10^6 \quad (1.2)$$

Where x is the year and y is the amount of methanol imports. The factory is projected to be built in 2030, so the projected methanol imports in 2030 (year 10) are:

$$y = -4386,3x + 1.10^6$$

$$y = -4386,3 (10) + 1.10^6$$

$$y = 956,137 \text{ ton/yr}$$

To determine *supply* is as follows:

$$\text{Supply} = \text{Import} + \text{Production}$$

$$= (956,137 + 800,000) \text{ ton/yr}$$

$$= 1,756,137 \text{ ton/yr}$$

c. *Demand*

The *demand* value is obtained from the export value and consumption of methanol. According to data from the Badan Pusat Statistik, methanol export data in Indonesia from 2021 to 2025 can be seen in Table 1.3

Table 1.4 Methanol export data 2021 to 2025

NO	YEAR	AMOUNT (TON)
1	2021	140,142
2	2022	57,155
3	2023	47,070
4	2024	132,285
5	2025	30,343

(Badan Pusat Statistics, 2025)

Based on the information above, Indonesia's methanol exports have a very small number. The results of the regression of the methanol export data can be seen in Figure 1.3, which shows the relationship between annual methanol export volumes and year.

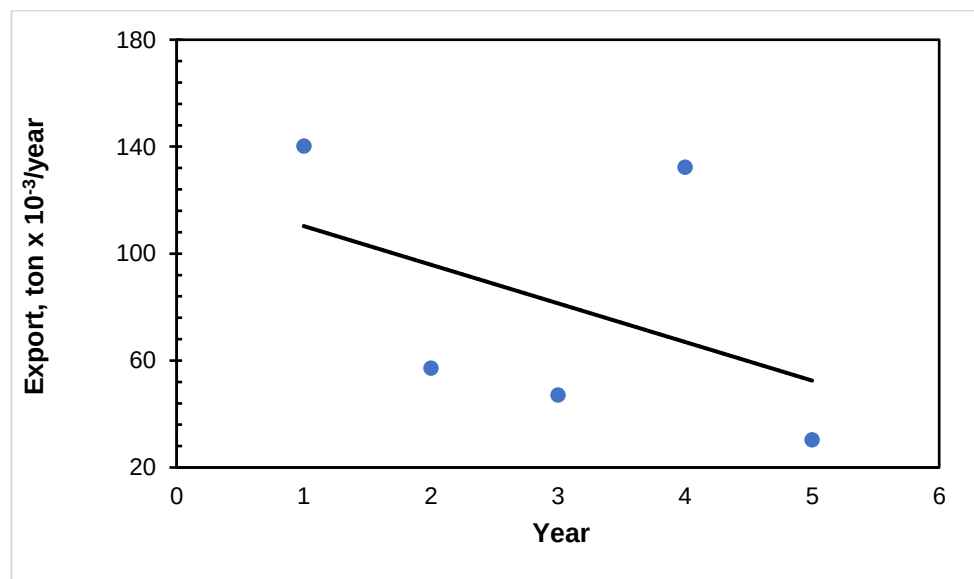


Figure 1.3 Methanol Export Chart 2021 to 2025

The estimated export of methanol in Indonesia in the coming year can be calculated using the following equation:

$$y = -14447x + 124739 \quad (1.3)$$

Where x is the year and y is the amount of methanol exports. The factory is projected to be built in 2030, so the projected methanol exports in 2030 are:

$$\begin{aligned} y &= -14447x + 124739 \\ y &= -14447(10) + 124739 \\ y &= -19,731 \text{ ton/yr} \end{aligned}$$

According to statistical data of manufacturing industry published by Badan Pusat Statistic regarding the need of methanol in Indonesia from each year has increased and decreased. The data can be seen in the following table 1.5

Table 1.5 Methanol consumption data in Indonesia

NO	YEAR	AMOUNT (TON)
1	2020	1,639,833
2	2021	1,702,082
3	2022	1,737,094
4	2023	1,853,988
5	2024	1,614,182

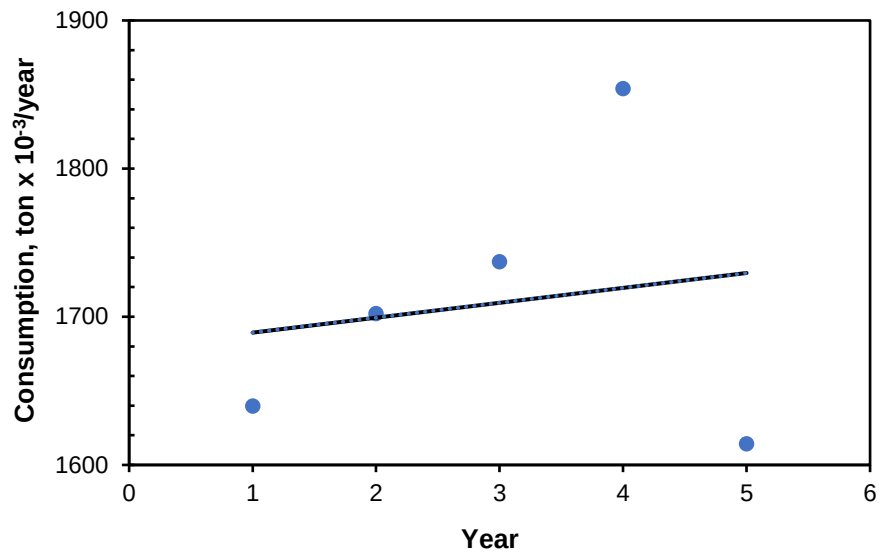


Figure 1.4 Methanol Consumption Chart in Indonesia

The estimated consumption of methanol in Indonesia in the coming year can be calculated using the following equation:

$$y = 10060x + 2.10^6 \quad (1.4)$$

Where x is the year, and y is the amount of methanol consumption. The factory is projected to be built in 2030, then the projection of methanol consumption in 2030 is:

$$y = 10060x + 2.10^6$$

$$y = 10060 (10) + 2.10^6$$

$$y = 2,100,600 \text{ ton/yr}$$

Therefore, to determine *demand* and capacity is as follows:

$$\begin{aligned} \text{Demand} &= \text{Export} + \text{Consumption} \\ &= (-19,731 + 2,100,600) \\ &= 2,080,869 \text{ ton/yr} \end{aligned}$$

$$\text{Capacity} = \text{Demand} - \text{Supply}$$

$$= (2,080,869 - 1,756,137)$$

$$= 324,732 \text{ ton/yr}$$

Note:

Identify Consumption Trends

$$r = \frac{\text{Consumption 2024} - \text{Consumption 2020}}{2020 - 2024}$$

$$r = \frac{288,139.877 - 329,432,789}{2020 - 2024}$$

$$r = \frac{-41,292,912}{4}$$

$$r = -10.323,228 \text{ ton/yr}$$

Margin Factor

$$\text{Plant Capacity} = \text{Consumption}_{2030} \times (1 + \text{margin})$$

$$\text{Plant Capacity} = 226,200,509 \times 1,10 = 248.820,6 \quad (10\%)$$

$$\text{Plant Capacity} = 226,200,509 \times 1,15 = 260.130,6 \quad (15\%)$$

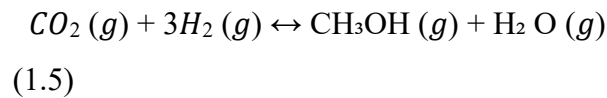
$$\text{Plant Capacity} = 226,200,509 \times 1,20 = 271.440,6 \quad (20\%)$$

$$\text{Final Decided Plant Capacity} = 150,000 \text{ ton/yr}$$

1.3 Literature Review

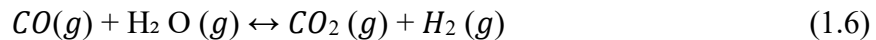
1.3.1 Raw Material and Products

In general, methanol can be produced from the hydrogenation of carbon dioxide using a catalyst. In general, the methanol synthesis reaction in the gas phase on a Cu-based catalyst can be presented as follows:



The above reaction is exothermic and results in a decrease in the number of moles or volume. To achieve high equilibrium conversion based on the principle of equilibrium, a process with high pressure and low temperature is desirable. However, on the other hand, this reaction occurs with the help of a solid catalyst, so it requires a high temperature to achieve a high reaction rate. Thus, a temperature optimization process is required to obtain optimal conversion.

In addition to the above reaction, the water-gas shift reaction can occur:

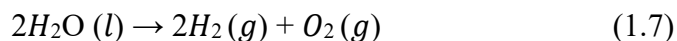


In methanol synthesis, the selection of catalyst type plays an important role in influencing the operating conditions of methanol synthesis. Each catalyst exhibits optimal catalytic activity under specific conditions. For example, the Cu/Zn/Al₂O₃ catalyst works well under operating conditions of 180°C – 280°C and 15 bar - 51 bar pressure, where the hydrogenation of carbon dioxide using the Cu/Zn/ Al₂O₃ catalyst produces the main products of methanol, CO, and water. Methane,

dimethyl ether, and methyl formate are also produced from the reaction, but the selectivity for these products is less than 0.1%. (Saito *et al.*,1998)

1.3.2 *Electrochemical Process*

The reaction for production methanol from CO₂ and H₂ begins with the production of H₂ gas via water electrolysis. Electrolysis of water is the process of breaking down water compounds into O₂ gas and H₂ gas using an electric current. At the cathode, two water molecules react by capturing two electrons, reducing to H₂ gas and hydroxide ions (OH⁻). Meanwhile, at the anode, two other water molecules decompose into oxygen gas (O₂), releasing 4 H⁺ ions and transferring electrons to the cathode. The H⁺ and OH⁻ ions then neutralize each other, reforming several water molecules. The overall equivalent reaction for water electrolysis can be written as follows.



Hydrogen gas produced from electrolysis can be used as a raw material for methanol production.

Carbon dioxide and other raw ingredients are bought straight from PT. Pupuk Iskandar Muda (PIM) Aceh.

Figure 1.5 Methanol Production with Electrochemical Process

Figure 1.5 Methanol Production with Electrochemical Process

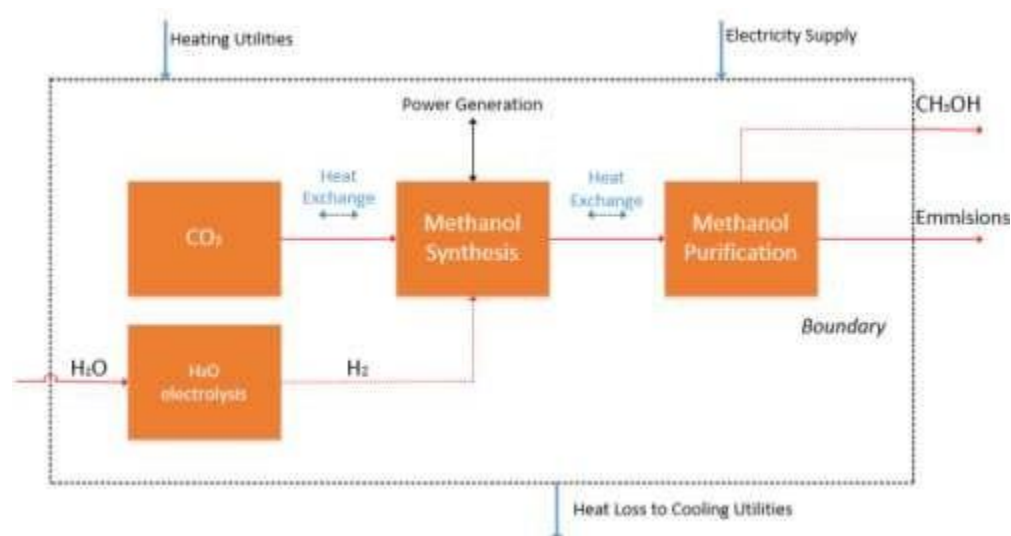


Figure 1.5 Methanol Production with Electrochemical Process

Methanol is still infrequently produced using electrochemical technologies that use water electrolysis to produce H_2 due to the high electrical energy required. In addition to producing O_2 gas as a by-product, the electrolysis unit's output does not emit any emissions. O_2 gas sales might thereby pay for the factory's electricity expenses, which are often greater in electrolysis unit

1.3.3 Process Technologies Developed by Several Companies

a. Low Pressure Methanol Synthesis Process – ICI

This process was first developed in the 1960s by the process development company Imperial Industries, Ltd. This synthesis process uses low pressure with a Cu-based catalyst. The use of Cu catalysts was developed in the 1920s, but at that time, these catalysts were not yet used in the methanol synthesis process. This was because Cu-based catalysts can be poisoned if sulfur compounds are present in the reactor feed, so low-pressure methanol synthesis processes using Cu-based catalysts could only be developed once technology for separating sulfur from syngas became available.

This process uses syngas feed containing carbon monoxide, carbon dioxide, hydrogen, and methane. A shift converter is used to regulate the

CO/H₂ ratio. The feed is then pressurized to 50 atm in a centrifugal compressor and fed into a quench reactor at an operating temperature of 270°C. The quench converter is a single bed containing an inert support catalyst. The reaction product is crude methanol containing water, dimethyl ether, esters, carbonyl iron, and other alcohols. The reaction product is then cooled and the crude methanol is purified by distillation. In its development, because it was considered less profitable, ICI replaced the type of reactor used from a quench reactor to a cooled tube, which is essentially the same as that used by Lurgi (Lee, 1990).

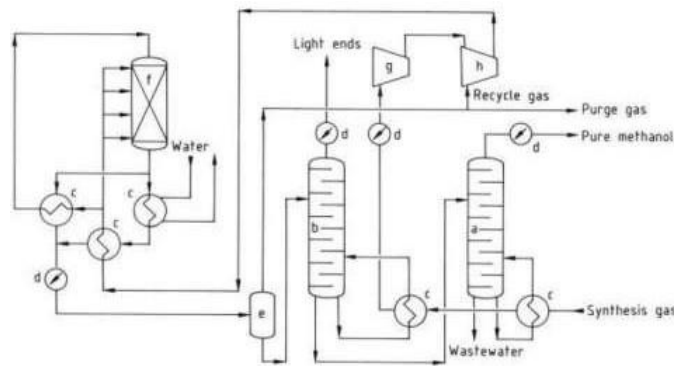


Figure 1.6 Low Pressure Methanol Synthesis Process Diagram – ICI

b. Low Pressure Methanol Synthesis Process – Lurgi

In the methanol synthesis process using Lurgi technology, a reactor operating at a temperature range of 220–260°C and a pressure range of 40–100 bar is used. The reactor design differs from its predecessor, ICI technology. Lurgi technology uses a quasi-isothermal shell and tube reactor, where the methanol reaction occurs on the tube side containing the catalyst and on the shell side through which the coolant flows. In addition, in this technology, the reactor also

functions as a steam generator with a pressure of 40–50 bar (Lee, 1990).

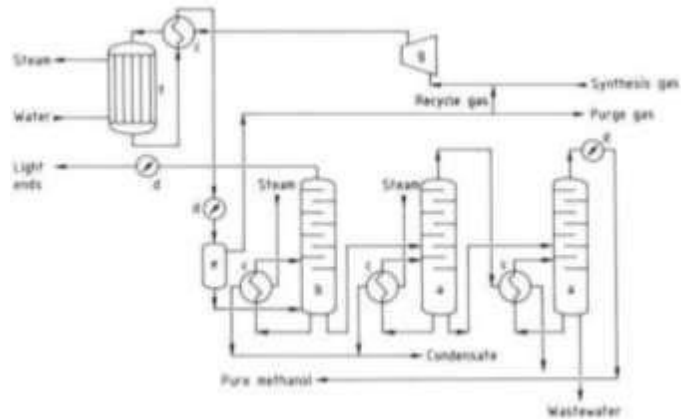


Figure 1.7 Low Pressure Methanol Synthesis Process Diagram – Lurgi

c. Low and Medium Pressure Methanol Synthesis Process – Mitsubishi Gas Chemical (MGC)

In the methanol synthesis process using MGC technology, methanol synthesis still uses copper (Cu)-based catalysts with reactor operating conditions in the temperature range of 200 – 280°C and pressure range of 50 – 150 atm. Initially, this Japanese company used a pressure of 150 atm, but later developed it to a pressure of less than 100 atm. The MGC process uses a reactor with double-walled tubes where the annulus is filled with catalyst. Syngas flows through the inner pipe while the outer pipe is filled with cooling water (Ullmann, 2005). The MGC process uses hydrocarbon's as feedstocks. The feedstocks is desulfurized before entering the steam reformer, which operates at 500°C. The outlet stream from the steam reformer has a temperature of 800–850°C and contains carbon monoxide, carbon dioxide, and hydrogen. Next, the syngas produced is pressurized with a centrifugal compressor and mixed with the recycle stream before being fed into the reactor (Lee, 1990).

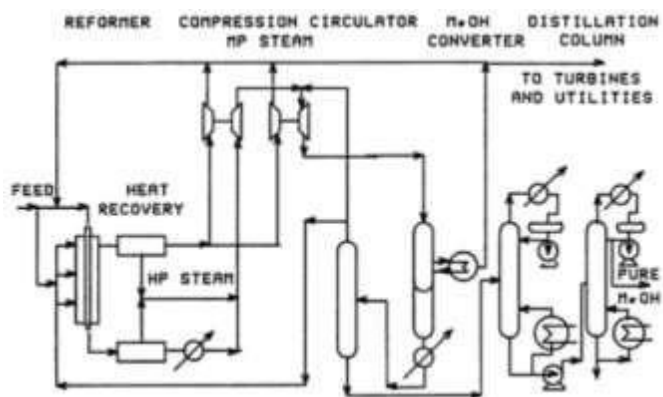


Figure 1.8 Diagram of Low/Medium Pressure Methanol Synthesis Process–MGC

d. Medium Pressure Methanol Synthesis Process – Kellogg M.W.

Kellogg Co. introduced a very different synthesis reaction, but it was essentially an adiabatic reactor. The reactor was spherical and contained a bed of catalyst. Similar to the ICI process, the heat generated by the reaction is controlled by intermediate coolers. This process uses copper catalysts and operates at a temperature range of 200-280°C and a pressure of 100-150 atm (Ullmann, 2005).

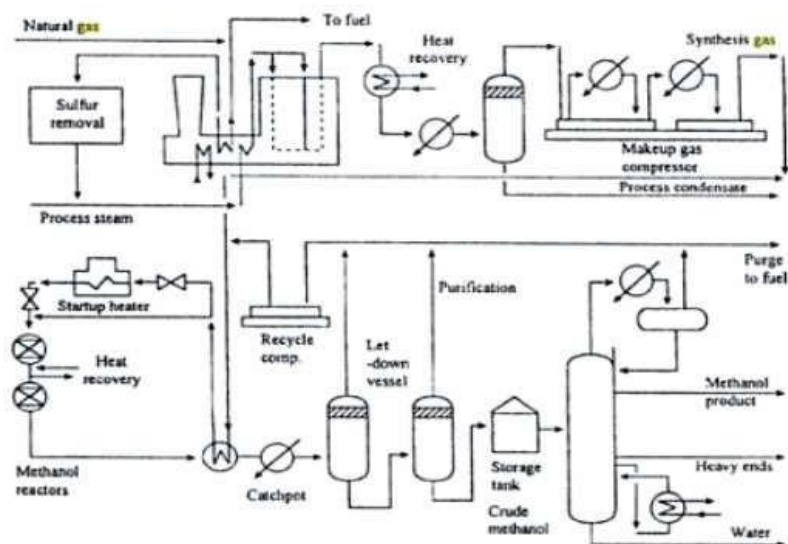


Figure 1.9 Diagram of the Medium Pressure Methanol Synthesis Process – Kellogg

e. Medium Pressure Methanol Synthesis Process – Nissui Topsoe

The reactor scheme for the Nissui Topsoe process from Denmark was designed by Nihon Suiso Kogyo of Japan. The reactor used is an adiabatic type with three radial flows, each of which has a catalyst stack and an internal heat exchanger. The synthesis gas flows radially through the catalyst bed. The operating pressure of this process is above 150 bar and the operating temperature is 200-310°C. The first product needs to be cooled before the second reactor. The cooling results in low-pressure steam. The catalyst used is Cu-Zn-Cr, which is active at 230-280°C and 100-200 atm (Lee, 1990).

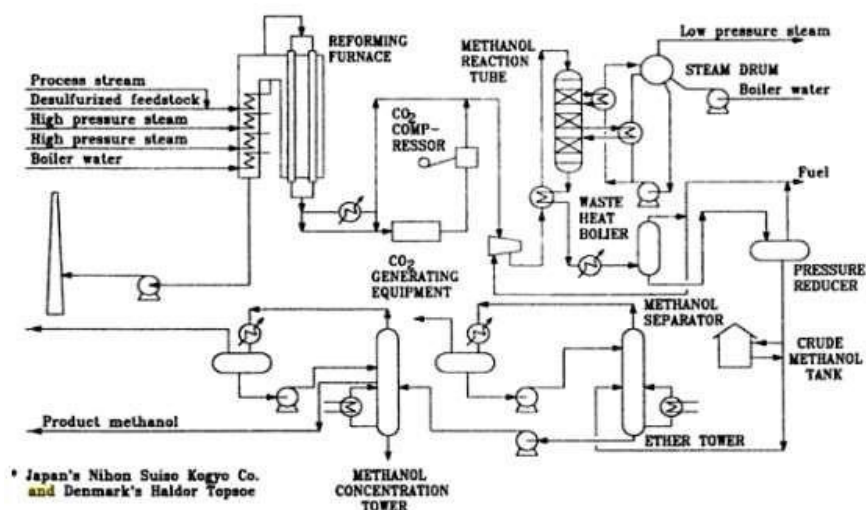


Figure 1.10 Medium Pressure Methanol Synthesis Process Diagram

Source: Lee, S. (1990). *Handbook of alternative fuel technologies*. CRC Press.

1.3.4 Industrial Applications of CO₂ and H₂ as Raw Materials

Several industrial initiatives have demonstrated the technical and commercial viability of utilizing CO₂ and H₂ as primary feedstocks for the synthesis of green methanol through Power-to-X (PtX) technology. Carbon Recycling International (CRI),

headquartered in Iceland, is among the earliest pioneers of this approach. Since 2012, CRI has operated the George Olah Plant in Svartsengi, Iceland, where geothermal CO₂ emissions are combined with green hydrogen — produced via water electrolysis powered by renewable electricity — to synthesize methanol through its proprietary Emissions-to-Liquids (ETL) process (Carbon Recycling International, 2023). The ETL process employs a one-step catalytic CO₂ hydrogenation route, which is reported to achieve higher methanol yields compared to conventional multi-step syngas-based processes (Carbon Recycling International, 2023).

Table 1.6 Comparison of methanol synthesis Processes from Various Licensors

No	Specification	ICI	Lurgi	MGC	Kellog	Nissui Topsoe
1	Operation condition: -Pressure (bar) -Temperature (°C)	50-100 220-280	40-100 220-260	50-150 200-280	100-150 200-280	100-200 200-310
2	Reactor: -Characteristics -Number of Reactors -Cooler	<i>Quench</i> 1 <i>Cold quench</i>	<i>Shell & Tube</i> 1 Cooling water (on shell)	<i>Annular</i> 1 Cooling water (outertube)	Adiabatic (axial) se3-4 Intermediate coolers	Adiabatic (radial) 3-4 Intermediate coolers
3	Advantages	has been proven to be the most widely used	high thermal efficiency and selectivity, more stable temperature	ideal temperature profile, minimal catalyst required	high production speed and capacity	high production speed and capacity
4	Disadvantages	low thermal efficiency, catalyst degradation	production capacity is not very large	complex, expensive reactor costs	high operating conditions, reduced selectivity	high operating conditions, reduced selectivity
5	Typical Per- Pass Conversion	15-25%	20-30%	20-30%	20-30%	20-30%

From the various processes above, the methanol synthesis process using a Lurgi reactor (multitube fixed bed reactor) was selected based on the following considerations:

a. Good Heat Transfer

With a multitube reactor, the presence of tubes expands the contact area between the reactants and the cooling medium. The larger the heat transfer area, the faster the heat transfer rate. The methanol synthesis reaction is an exothermic reaction. Therefore, rapid heat transfer is necessary to properly control the temperature of the methanol synthesis reaction.

b. High Product Selectivity

This process uses a copper-based catalyst supported by zinc and aluminum compounds ($\text{Cu/ZnO/Al}_2\text{O}_3$). The use of copper-based catalysts has been proven to have high selectivity for methanol products, which is why the majority of methanol plants currently use these catalysts.

c. Relatively Simple Reactor Fabrication

The reactor used is similar to a multitube heat exchanger (HE), which is a device commonly used in chemical plants. With the advancement of HE manufacturing technology, it is relatively easy to manufacture this type of reactor.

d. Operating Conditions Fall Under the Low Category

Given the type of reactor and catalyst used in the Lurgi process, the operating conditions required are not particularly extreme ($P < 100 \text{ atm}$, $T < 300^\circ\text{C}$). In comparison, the first industrial-scale methanol synthesis process (the BASF process) had an operating pressure of 300 atm and extreme temperatures. With the operating conditions of the Lurgi process, the

thickness of the equipment used is reasonable, so the price of reactor materials is relatively inexpensive. In addition, lower operating conditions result in lower inherent hazards from the process.

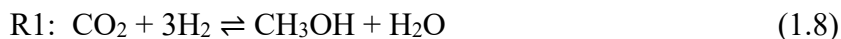
With the above considerations, the selected production process employs water electrolysis to generate H_2 gas, combined with CO_2 sourced from PT. Roda Manunggal Inti (RMI). Thus, the electricity supply for the electrolysis process can be obtained from the coal-fired power plant with the following considerations:

1. Creating an environmentally friendly factory.
2. Offering new options for industrial-scale methanol production technology based on the concept of electrolysis.
3. With methanol priced at \$0.9525 per kg, oxygen by-product priced at \$0.8931 per kg, and CO_2 raw material priced at \$0.4490, establishing this methanol factory is a very promising venture.

1.4 Overview of Thermodynamics and Kinetics

1.4.1 Thermodynamics Review:

The synthesis of methanol from CO₂ and H₂ does not proceed through a single reaction. Three reactions occur simultaneously over the catalyst: the direct hydrogenation of carbon dioxide to methanol (R1), the reverse water–gas shift reaction (R2), and the hydrogenation of carbon monoxide to methanol (R3). These reactions are coupled, since R1 is the algebraic sum of R3 and R2 (Bansode & Urakawa, 2014; Bowker, 2019). The three reactions evaluated in this design are presented below:



The thermodynamic assessment is based on the standard enthalpy of formation (ΔH_f°) and the standard Gibbs energy of formation (ΔG_f°) of each species, evaluated at the reference temperature of 298.15 K. All species participating in the reactions are in the gas phase. The property data used in the calculation are summarized in Table below.

Table 1.7 Standard thermodynamic properties of the components at 298.15 K

Component	MW (g/mol)	ΔH_f° (kJ/mol)	ΔG_f° (kJ/mol)
CO ₂ (g)	44.01	−393.51	−394.39
H ₂ (g)	2.016	0.00	0.00
CO (g)	28.01	−110.53	−137.16
CH ₃ OH (g)	32.04	−200.94	−162.24
H ₂ O (g)	18.02	−241.82	−228.59

Source: Smith, Van Ness, & Abbott (2018); NIST Chemistry WebBook (Linstrom & Mallard, 2023)

a. Standard Enthalpy of Reaction (ΔH°)

The standard enthalpy of reaction is calculated from the enthalpies of formation of the products and reactants according to the following equation (Smith, Van Ness, & Abbott, 2018):

$$\Delta H^\circ_{\text{rxn}} = \sum v_i \Delta H_f^\circ(\text{products}) - \sum v_i \Delta H_f^\circ(\text{reactants}) \quad (1.11)$$

where v_i is the stoichiometric coefficient of component i . Applying Equation 1.4 to each reaction at 298.15 K gives:

$$\Delta H^\circ_{\text{R1}} = [(-200.94) + (-241.82)] - [(-393.51) + 3(0)] = -49.25 \text{ kJ/mol}$$

$$\Delta H^\circ_{\text{R2}} = [(-110.53) + (-241.82)] - [(-393.51) + 0] = +41.16 \text{ kJ/mol}$$

$$\Delta H^\circ_{\text{R3}} = [(-200.94)] - [(-110.53) + 2(0)] = -90.41 \text{ kJ/mol}$$

The negative ΔH° values for R1 and R3 indicate that the methanol-forming reactions are exothermic, releasing heat to the surroundings, whereas the positive ΔH° of R2 shows that the reverse water–gas shift reaction is endothermic. Because the dominant methanol-forming reactions are exothermic, the reactor must be provided with an adequate cooling system to maintain a stable operating temperature, and a moderate operating temperature is favorable for the equilibrium yield of methanol (Bowker, 2019).

b. Standard Gibbs Energy of Reaction (ΔG°) and Spontaneity

The spontaneity of a reaction is evaluated from the standard Gibbs energy of reaction. A reaction is thermodynamically spontaneous when ΔG° is negative. The standard Gibbs energy of reaction is obtained from the Gibbs energies of formation (Smith, Van Ness, & Abbott, 2018):

$$\Delta G^\circ_{\text{rxn}} = \sum v_i \Delta G_f^\circ(\text{products}) - \sum v_i \Delta G_f^\circ(\text{reactants}) \quad (1.12)$$

Evaluating Equation 1.5 at 298.15 K yields:

$$\Delta G^\circ_{\text{R1}} = [(-162.24) + (-228.59)] - [(-394.39) + 3(0)] = +3.56 \text{ kJ/mol}$$

$$\Delta G^\circ_{\text{R2}} = [(-137.16) + (-228.59)] - [(-394.39) + 0] = +28.64 \text{ kJ/mol}$$

$$\Delta G^{\circ}R3 = [(-162.24)] - [(-137.16) + 2(0)] = -25.08 \text{ kJ/mol}$$

At standard conditions (298.15 K), R3 has a negative ΔG° and is therefore spontaneous, while R1 and R2 have small positive ΔG° values and are non-spontaneous. A summary of the standard enthalpy, Gibb's energy, and thermal character of each reaction is presented in Table below.

Table 1.8 Standard enthalpy and Gibb's energy of reaction at 298.15 K

Reaction	$\Delta H^{\circ}298$ (kJ/mol)	$\Delta G^{\circ}298$ (kJ/mol)	Character
R1	-49.25	+3.56	Exothermic
R2	+41.16	+28.64	Endothermic
R3	-90.41	-25.08	Exothermic

The slightly positive ΔG° of R1 at standard conditions does not mean that methanol cannot be produced. It indicates that the reaction is equilibrium-limited at low temperature and ambient pressure. Because R1 proceeds with a net decrease in the number of moles (four moles of reactant form two moles of product), an increase in operating pressure shifts the equilibrium toward methanol according to Le Chatelier's principle. For this reason, the process is operated at an elevated pressure of 50 atm, which is consistent with industrial CO₂-to-methanol practice (Bansode & Urakawa, 2014).

c. Equilibrium Constant (K)

The relationship between the standard Gibbs energy of reaction and the equilibrium constant at the reference temperature is expressed as (Smith, Van Ness, & Abbott, 2018):

$$\ln K_{298} = -\Delta G^{\circ}298 / (R T) \quad (1.13)$$

where R is the universal gas constant ($8.314 \times 10^{-3} \text{ kJ/mol}\cdot\text{K}$) and T is the absolute temperature (K). Substituting the $\Delta G^{\circ}298$ values into Equation 1.6 gives the equilibrium constants at 298.15 K listed in Table 1.3.

The reactor is operated at $T = 220\text{ }^{\circ}\text{C}$ (493.15 K), which is far above the reference temperature. Over a temperature interval of nearly 200 K the standard enthalpy of reaction cannot be assumed constant, because the heat capacities of the reactants and products differ. The equilibrium constant at the reaction temperature is therefore evaluated using a temperature-dependent enthalpy of reaction, in which the heat-capacity change of the reaction (ΔC_p) is expressed through the ideal-gas polynomial (Smith, Van Ness, & Abbott, 2018):

$$C_p / R = A + B T + C T^2 + D T^{-2} \quad (1.14)$$

The heat-capacity coefficients of each component are given in Table 1.3, and the heat-capacity change of reaction is obtained from $\Delta A = \sum v_i A_i$, and similarly for ΔB , ΔC and ΔD .

Table 1.9 Ideal-gas heat-capacity coefficients of the components ($C_p/R = A + BT + CT^2 + DT^{-2}$)

Component	A	$10^3 B \text{ (K}^{-1}\text{)}$	$10^6 C \text{ (K}^{-2}\text{)}$	$10^{-5} D \text{ (K}^2\text{)}$
CO ₂	5.457	1.045	0.000	-1.157
H ₂	3.249	0.422	0.000	0.083
CO	3.376	0.557	0.000	-0.031
CH ₃ OH	2.211	12.216	-3.450	0.000
H ₂ O	3.470	1.450	0.000	0.121

Source: Smith, Van Ness, & Abbott (2018), Table C.1

The variation of the standard enthalpy of reaction with temperature is then:

$$\Delta H^{\circ}(T) = \Delta H^{\circ}_{298} + R \int_{298}^T (\Delta C_p/R) dT \quad (1.15)$$

and the equilibrium constant at the reaction temperature is obtained by integrating the van 't Hoff relation with this temperature-dependent enthalpy:

$$\ln K_T = \ln K_{298} + \int_{298}^T [\Delta H^{\circ}(T) / (R T^2)] dT \quad (1.16)$$

For comparison, the equilibrium constant was also estimated using the simplified van 't Hoff equation, which assumes that ΔH° remains constant and equal to its value at 298 K:

$$\ln (K_T / K_{298}) = - (\Delta H^\circ_{298} / R) [(1/T) - (1/298.15)] \quad (1.17)$$

Both results are presented in Table 1.4 so that the effect of the heat-capacity correction can be seen. The values obtained from the rigorous method (with ΔC_p) are adopted in this design, since they are more representative of the actual operating temperature.

Table 1.10 Equilibrium constants at 298.15 K and at the reaction temperature (493.15 K)

Reaction	K (298 K)	K (493 K)
R1	2.38×10^{-1}	1.1×10^{-2}
R2	9.60×10^{-6}	1.7
R3	2.48×10^4	6.4×10^{-3}

All equilibrium constants are positive ($K > 0$) at both temperatures, confirming that each reaction is thermodynamically possible. All equilibrium constants are positive ($K > 0$) at both temperatures, confirming that each reaction is thermodynamically possible. It should be noted that a value expressed with a negative exponent, such as $K = 2.38 \times 10^{-1}$ for R1 at 298 K, represents a small positive number ($K < 1$); it does not indicate a negative equilibrium constant. A value of K below unity means only that the reaction is equilibrium-limited and that, at chemical equilibrium, an appreciable fraction of the reactants remains unconverted, which is overcome in practice through elevated pressure, an active catalyst, and continuous removal of the product.

The heat-capacity correction reduces the equilibrium constants of the exothermic methanol-forming reactions (R1 and R3) by several orders of magnitude relative to the simplified van 't Hoff estimate, confirming that the simplification is not negligible over this temperature range. The temperature dependence remains consistent with the

thermal character of the reactions: for the exothermic reactions R1 and R3 the equilibrium constant decreases as temperature increases (R1 from 2.38×10^{-1} to 1.1×10^{-2} , and R3 from 2.48×10^4 to 6.4×10^{-3}), whereas for the endothermic reverse water–gas shift reaction R2 it increases with temperature (from 9.60×10^{-6} to 1.7). This behavior explains the operating-condition selection for the reactor: a moderate temperature of 220 °C provides a sufficient reaction rate while limiting the loss of the methanol-forming equilibrium, and the elevated pressure of 50 atm compensates for the equilibrium limitation by driving the mole-reducing methanol-synthesis reaction toward the product side (Bowker, 2019; Bansode & Urakawa, 2014).

1.4.2 Kinetics Review:

Thermodynamics determines the theoretical maximum conversion and spontaneity. While kinetics addresses the rate at which the methanol synthesis reaction proceeds. kinetics determines whether this conversion can be achieved in practical time scales. The concentration can influence reaction rate (partial pressures in gas phase), temperature, catalyst activity, and reactor configuration (Skrzypek et al., 1994; Graaf et al., 1988).

For methanol production from CO₂ and H₂, reactor design and catalyst volume sizing are both governed by the kinetics of the methanol synthesis reaction. The model created by Vanden Bussche and Froment (1996) is one of the most popular and trustworthy kinetic models for industrial methanol synthesis over a Cu/ZnO/Al₂O₃ catalyst. The reverse water-gas shift (RWGS) reaction and methanol synthesis are both described by this model, which may be used at temperatures between 180 and 280 °C and pressures of up to 51 bar .

According to this model, carbon dioxide is directly hydrogenated to produce methanol, while CO is produced as an intermediate through the reverse water-gas shift reaction. The process is based on a Langmuir–Hinshelwood–Hougen–Watson (LHHW) approach, where the rate expression incorporates both surface response phenomena and adsorption.

$$r_{MeOH} = \frac{k \left(P_{CO_2} P_{H_2}^3 - \frac{P_{MeOH} P_{H_2O}}{K_{eq}} \right)}{\left(1 + K_{CO_2} P_{CO_2} + K_{CO} P_{CO} + K_{H_2O} P_{H_2O} \right)^2} \quad (1.31)$$

The Arrhenius equation determines the relationship between temperature and the rate constant k through the activation energy, which represents the minimum energy required for reactant molecules to form products (Smith et al., 2005).

Arrhenius Equation:

$$k = A \exp\left(\frac{-E_a}{RT}\right) \quad (1.32)$$

k = Rate constant (mol/(kg catalyst·s·barⁿ))

A = Pre-exponential frequency factor

E_a = Activation energy (J/mol)

R = Universal gas constant (8.314 J/(mol·K))

T = Temperature (K)

Parameter	Value	Unit
A1	4.897×10^7	mol/(kg catalyst·s·bar ^{1.5})
E_a 1	113000	J/mol
A 2	9.64×10^{11}	mol/(kg catalyst·s·bar ^{1.5})
E 2	152900	J/mol
A3	1.09×10^5	mol/(kg catalyst·s·bar ^{1.5})
E 3	87500	J/mol

Table 1.10 Kinetic parameters for methanol synthesis over Cu/ZnO Catalyst

Based on Vanden Bussche and Froment's kinetic model is mechanistically grounded and accurately forecasts reactor performance under a variety of operating situations, it is especially well suited for industrial reactor design. The design of fixed-bed methanol reactors, notably multitube Lurgi-type reactors, have made substantial use of it. The model is used for this initial design using a commercial Cu/ZnO/Al₂O₃ catalyst at an operating temperature of 220 °C and a pressure of 50 atm. Methanol synthesis proceeds quickly under these circumstances while retaining a high degree of selectivity for methanol.

CHAPTER II

PRODUCT DESIGN

2.1 Product Specification

Table 2.1 Main product specifications (CH₃OH)

Parameter	Methanol
Molecular Formula	CH ₃ OH
Molecular Weight (g/mol)	32.04
Boiling Point (°C)	64.7
Freezing Point (°C)	-97
Density g/cm ³ (20°C)	0.7918
Vapor Pressure, 20°C (kPa)	12.3
Water Solubility	Completely Miscible
Viscosity, 20°C (mPa.s)	0.59
Concentration (%)	99.7
Phase	Liquid

Table 2.2 By-Product specifications (O₂)

Parameter	Oxygen
Molecular Formula	O ₂
Molecular Weight (g/mol)	32
Boiling Point (°C)	-182.962
Freezing Point (°C)	-218.79
Density g/L (gas)	1.429
Vapor Pressure, 90K (kPa)	100
Triple Point, (0.1463 kPa)	54.361 K
Molar Heat Capacity, (J/mol K)	29.378
phase	Gas

2.2 Raw Material Specification

Table 2.3 Raw Material Specification (CO₂)

Parameter	Carbon Dioxide
Molecular Formula (gas)	CO ₂
Molecular Weight (g/mol)	44.0341
Boiling Point (°C)	-78
Freezing Point (°C)	-57
Density kg/m ³ (40atm, 40°C)	49.87
Viscosity cp (-78°C)	0.07
Solubility in water (g/L)	1.45

Table 2.4 Raw Material Specification (H₂ O)

Parameter	Water
Molecular Formula (Liquid)	H ₂ O
Molecular Weight (g/mol)	18.01528
Boiling Point (°C)	99.974
Freezing Point (°C)	0
Density g/cm ³ (Liquid)	0.995
Vapor Pressure, 25°C (atm)	0.03127
Heat of Vaporization (kcal/mol)	9.7171
Thermal Conductivity W/mK	0.6065
Physical Appearance	Liquid colorless, odorless, tasteless

Table 2.5 Supporting Materials Specification (Cu/ZnO/Al₂O₃ Catalyst)

Parameter	Cu/ZnO/Al ₂ O ₃ Catalyst
Molecular Formula (Solid pellets)	Cu/ZnO/Al ₂ O ₃ (composition: 60 wt% Cu : 30 wt% ZnO : 10 wt% Al ₂ O ₃)
Molecular Weight (g/mol)	-
Operating Temperature (°C)	200 - 300
Density g/cm ³	1 - 1.3
Pressure, 25°C (atm)	50 - 100
Thermal Conductivity W/mK	0.2 - 0.5
Physical Appearance	Blue- Green cylindrical pellets

Table 2.6 Supporting Materials Specification (Nitrogen N₂)

Parameter	Nitrogen
Molecular Formula (Liquid)	N ₂
Molecular Weight (g/mol)	28.014
Boiling Point (°C)	-195.8
Freezing Point (°C)	-210
Density g/cm ³ (Liquid)	1.251
Vapor Pressure, 25°C (atm)	-
Heat of Vaporization (kJ/mol)	5.57
Thermal Conductivity W/mK	0.0260
Physical Appearance	colorless, odorless, tasteless

2.3 Quality Control

2.3.1 Quality Control of Raw Materials

The raw materials must be tested before the production process starts. This is to guarantee that the raw materials utilized fulfill the requirements established in the production design. ASTM 1972 is the evaluation standard for raw material quality.

2.3.2 Control of Product Quality

From the raw materials stage to the finished product stage, production monitoring and control are required to preserve the end product's quality. Laboratory analysis or the use of control tools can be used to monitor the quality of raw materials, auxiliary materials, semi-finished products, and items that support process quality.

2.3.3 Process Control

Control devices in the control room are used to supervise and manage operations utilizing automatic control techniques that make use of indicators. Signals or indications like lights, warning noises, and so forth can be used to identify any deviations from the predetermined indicators, such as the flow rate of raw materials, products, level control, or temperature management. If there is a deviation, it needs to be automatically or manually restored to its initial state.

Temperature and pressure controls are two examples of the control instruments employed. Under some circumstances, the following control tools must be installed:

a. Level control

This apparatus is mounted at the top of the tank. A light will flash and an alert will sound if the conditions are not met.

b. Temperature control

Every processing unit has this gadget attached. Like the level control, an alarm will ring and a light will flash if the conditions are not met.

c. Flow rate control

This apparatus serves to control the process's inflow and outflow.

d. Pressure control

This apparatus controls the process's intake and outflow pressures. An audio and visual warning signal will be activated if the requirements are not met.

2.3.4 Quantity Control

Operator mistakes, machine damage, raw material delays, unnecessary tool repairs, and other factors can all lead to quantity variations. To guarantee that the production process runs well, these deviations must be confirmed and assessed.

2.3.5 Process Material Control

In order to prevent shortages and achieve the intended production capacity, process material control helps to regulate the availability of raw materials.

CHAPTER III

PROCESS DESIGN

3.1 Process and Material Flow Diagram

3.1.1 Qualitative Flow Diagram

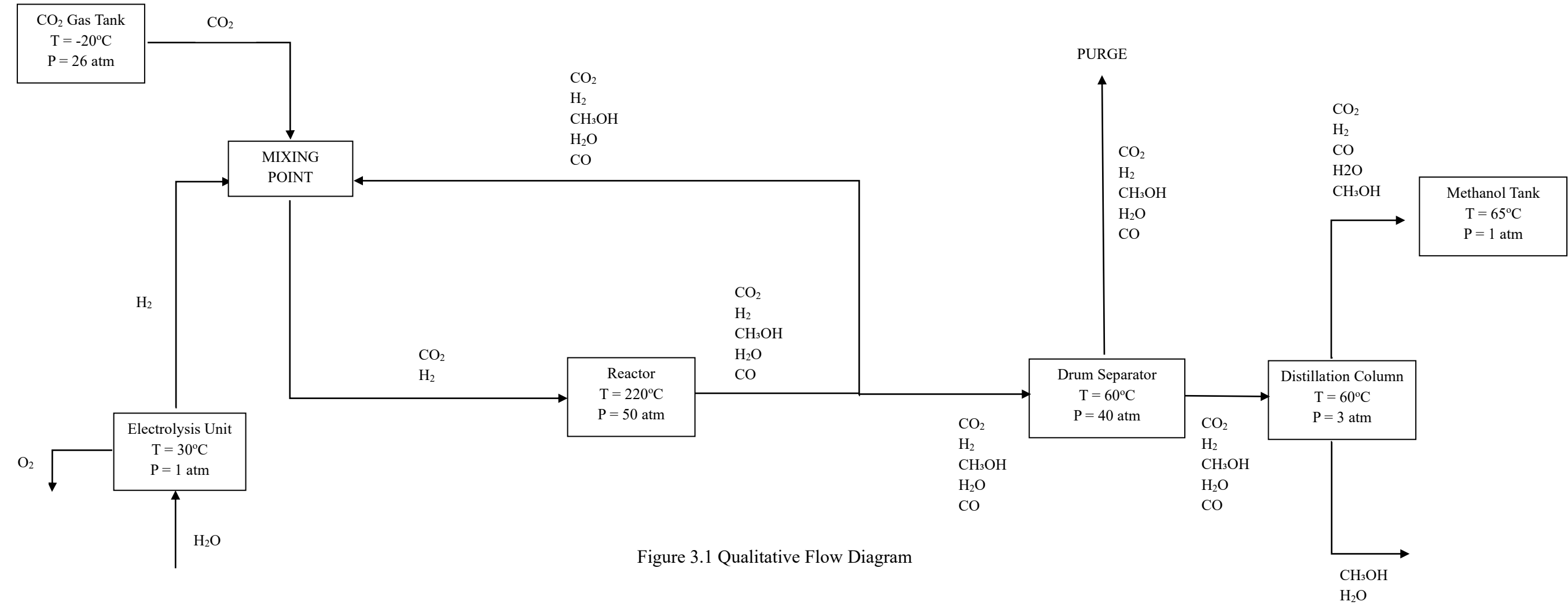


Figure 3.1 Qualitative Flow Diagram

3.1.2 Quantitative Flow Diagram

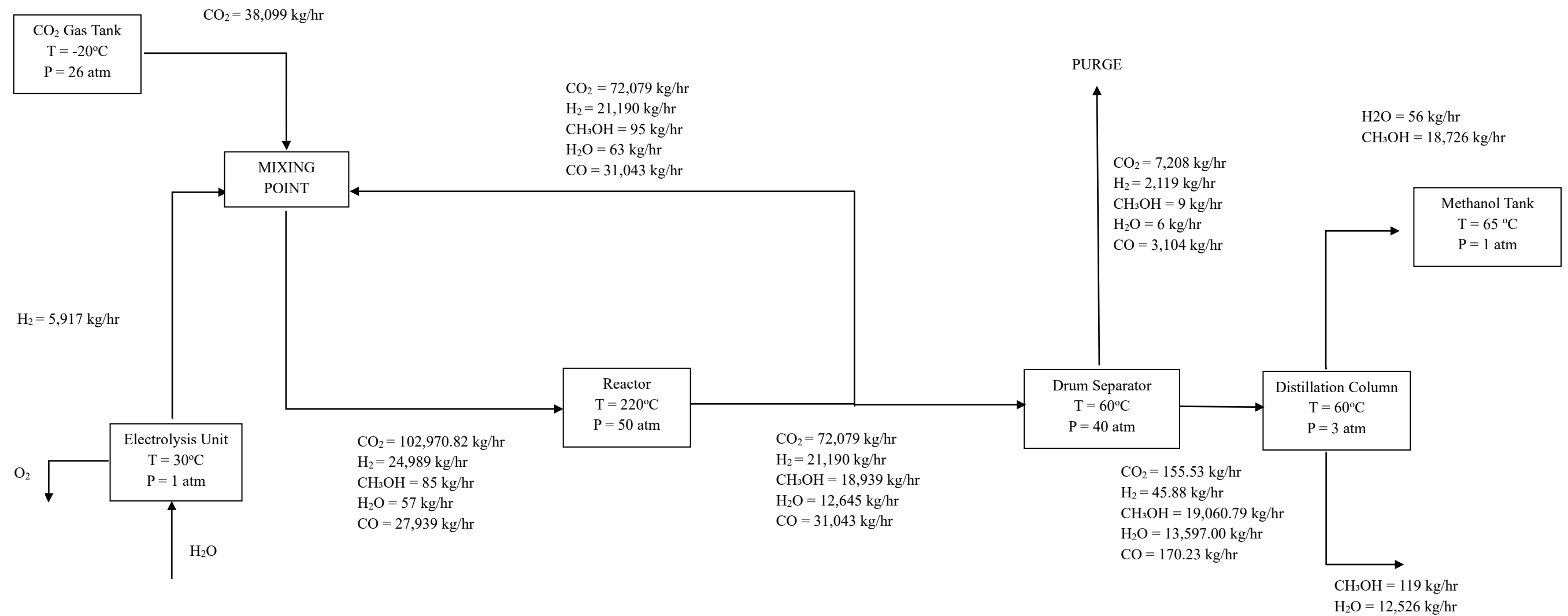


Figure 3.2 Quantitative Flow Diagram

3.2 Process Description

This section describes the production of methanol from carbon dioxide (CO_2) and hydrogen (H_2) using catalytic hydrogenation in a recycle process. The plant is designed to produce 150,000 tonnes of methanol per year with a product purity of 99.7 wt%. The overall process consists of feed storage and handling, gas compression, feed preheating, methanol synthesis in a fixed-bed reactor, cooling and gas-liquid separation, recycle and purge, methanol purification by distillation, and product storage.

3.2.1 Feed preparation and Raw Material Storage

The main raw materials used in this process are carbon dioxide and hydrogen. CO_2 is stored in tank TK-101 as a liquid at approximately 30 °C and 26 bar. Storing CO_2 in liquid form significantly reduces the required storage volume compared to gas-phase storage, owing to its high liquid density. Hydrogen is stored in TK-102 at 30 °C and 200 atm as compressed gas. In industrial practice, hydrogen is commonly supplied from an on-site water electrolysis unit or by pipeline from a nearby hydrogen production facility. Fresh CO_2 (Stream S1) and fresh H_2 (Stream S2) are withdrawn from storage and sent to feed compressors, K-102 and K-103, respectively. Both streams are compressed from atmospheric pressure to the reactor operating pressure of 50 bar (Perry and Green, 2019; Towler and Sinnott, 2022).

3.2.2 Mixing Feed Preheating

The compressed fresh CO_2 and H_2 are combined with the recycle gas stream (S5) in the mixing point MP-01. The recycle stream contains unreacted CO_2 and H_2 , along with smaller amounts of carbon monoxide, methanol, and water vapor.

The mixed reactor feed (S3) enters the feed preheater HE-101, where its temperature is increased from approximately 30 °C to the reactor inlet temperature of 220 °C. The preheater uses high-pressure steam or hot process gas as the heating medium. This heating step is required to achieve sufficient reaction rates over the $\text{Cu/ZnO/Al}_2\text{O}_3$ catalyst. The selected inlet temperature of 220 °C represents a compromise between reaction kinetics, equilibrium conversion, and catalyst stability (Vanden Bussche and Froment, 1996; Froment and Bischoff, 1990).

3.2.3 Methanol Synthesis Reactor (R-101)

The preheated feed enters a multi-tubular fixed-bed reactor (R-101) packed with a commercial copper/zinc oxide/alumina ($\text{Cu/ZnO/Al}_2\text{O}_3$) catalyst. The reactor operates isothermally at 220 °C and 50 bar. Methanol is produced primarily by the catalytic hydrogenation.

In the selected multi-tubular fixed-bed reactor, the $\text{Cu/ZnO/Al}_2\text{O}_3$ catalyst is packed inside the tubes, while cooling water flows through the shell side. This configuration is preferred because methanol synthesis is highly exothermic and requires efficient heat removal. The small tube diameter provides a large heat-transfer area, allowing rapid removal of reaction heat and maintaining near-isothermal conditions throughout the catalyst bed. This reduces hot-spot formation, prevents catalyst deactivation, and improves methanol selectivity. Therefore, industrial Lurgi reactors place the catalyst in the tubes and use the shell side for cooling to achieve better temperature control and longer catalyst life (Lee, 1990; Perry and Green, 2019).

The $\text{Cu/ZnO/Al}_2\text{O}_3$ catalyst is the industry standard for low-pressure methanol synthesis due to its high selectivity, activity, and long operating life. The reactor is designed for a single-pass CO_2 conversion of approximately 30%, which is consistent with industrial practice where unreacted gases are recovered and recycled to increase overall conversion. Heat generated by the exothermic reactions is removed through the shell side of the reactor to maintain near-isothermal conditions and prevent catalyst deactivation by overheating (Vanden Bussche and Froment, 1996; Graaf et al., 1988; Behrens et al., 2012).

3.2.4 Reactor Effluent Cooling (E-102)

The reactor effluent stream (S4) exits the reactor at approximately 220 °C and contains methanol, water, carbon monoxide, unreacted carbon dioxide, and hydrogen. This stream is cooled to 60 °C in the reactor outlet cooler (E-102) using cooling water. Cooling condenses the majority of methanol and water while maintaining most of the permanent gases in the vapor phase. Efficient cooling is critical to maximize liquid recovery in the downstream separator and reduce the load on the distillation column (Towler and Sinnott, 2022; Coulson and Richardson, Vol. 6).

3.2.5 Gas-Liquid Separation (DS-101)

The cooled reactor effluent enters a vertical flash drum separator (DS-101), where the liquid and vapor phases are separated based on differences in density. The separator operates at approximately 60 °C and high pressure. The liquid stream (S7) consists primarily of crude methanol and water with trace dissolved gases, while the vapor stream contains hydrogen, carbon dioxide, carbon monoxide, and small quantities of methanol and water.

A high-pressure separator is preferred in industrial methanol plants because it minimizes recompression duty and reduces energy consumption in the recycle loop. Proper sizing of the separator is based on the Souders-Brown method to ensure adequate vapor disengagement and liquid residence time (Perry and Green, 2019; Coulson and Richardson, Vol. 6).

3.2.6 Recycle Gas Compression and Purge (K-101)

Gas recycle is essential because methanol synthesis is equilibrium-limited and single-pass conversion is relatively low. Recycling unreacted reactants significantly increases overall carbon utilization and improves process economics. The purge ratio is selected to balance reactant recovery against the accumulation of inerts in the system (Luyben, 2010; Olah et al., 2018).

3.2.7 Methanol Purification by Distillation (DC-101)

The crude liquid stream from the separator (S7) contains methanol, water, and trace dissolved gases. This stream is sent to a sieve-tray distillation column (DC-101) operating at approximately 3 atm. The purpose of the column is to produce high-purity methanol as the overhead product and remove water as the bottom product.

Methanol is more volatile than water, allowing separation by conventional distillation. The column is designed to produce methanol with a purity of 99.7 wt%, meeting commercial specifications such as AA grade methanol. The operating pressure of 3 atm provides a suitable balance between column temperature, condenser duty, and cooling water availability (Perry and Green, 2019; Coulson and Richardson, Vol. 2).

3.2.8 Distillation Condenser and Reflux System (E-103, ACC-01, P-102)

The overhead vapor from the distillation column is condensed in a total condenser (E-103) using cooling water. The condensed liquid is collected in an accumulator drum (ACC-01), where it is split into two streams: reflux and product. The reflux stream is pumped back to the top of the column by the reflux pump (P-102) to provide internal liquid flow and improve separation efficiency. Higher reflux ratios reduce the number of required trays but increase condenser and reboiler duties. The selected reflux ratio in this design is approximately 1.5 times the minimum reflux ratio, which is a common design practice (Seader et al., 2016; Coulson and Richardson, Vol. 2).

3.2.9 Distillation Reboiler (E-104)

A kettle-type reboiler (E-104) supplies the heat required to vaporize part of the column bottoms and generate the upward vapor flow necessary for separation. The reboiler duty is determined from the material and energy balances around the column. Adequate reboiler design is essential to maintain stable column operation and achieve the specified methanol purity (Perry and Green, 2019; Towler and Sinnott, 2022).

3.2.10 Methanol Product Storage (TK-103A/B)

The purified methanol product stream (S8) is transferred by the product pump (P-103) to atmospheric storage tanks (TK-103A and TK-103B). Two tanks are provided to allow simultaneous filling and product dispatch. The tanks are designed according to API 650 and operate near ambient temperature and pressure. Methanol is stored in fixed-roof carbon steel tanks equipped with nitrogen blanketing to reduce fire risk and minimize moisture absorption. The storage capacity is based on approximately 30 days of production to provide operational flexibility and accommodate shipping schedules (API 650; Methanol Institute, 2023).

3.2.11 Wastewater Handling (TK-104)

The bottom stream from the distillation column (S9), which consists mainly of water with trace methanol, is sent to the wastewater storage tank (TK-104). Depending on environmental regulations residual methanol may be recovered in a secondary treatment step or the stream may be sent directly to a biological wastewater treatment unit. Proper handling of this stream is necessary to comply with environmental discharge standards (Towler and Sinnott, 2022; Metcalf & Eddy, 2014).

3.3 Equipment Specifications

Table 3.1 Reactor Design Specification

Parameter	Value	Unit
1. GENERAL IDENTIFICATION		
Equipment Type	Packed Bed Reactor (PBR)	—
Process	CO ₂ Hydrogenation to Methanol	—
Plant Capacity	150,000	ton/yr
Key Reaction	CO ₂ + 3H ₂ → CH ₃ OH + H ₂ O	—
Side Reaction (RWGS)	CO ₂ + H ₂ → CO + H ₂ O	—
Reactor Type	Multi-tube fixed bed	—
2. OPERATING CONDITIONS		
Operating Temperature	220	°C
Operating Temperature (K)	493.15	K
Operating Pressure	50	atm
Operating Pressure (bar)	50.6625	bar
Operating Pressure (kPa)	5,066.25	kPa
3. FEED STREAM — REACTOR INLET (STREAM S3)		
Total Molar Feed Rate	19,027.01	kmol/hr
CO ₂ Molar Flow Rate (F _{a0})	701.17	mol/s
CO ₂ Mole Fraction (y _{a0})	0.1327	mol/mol
H ₂ Molar Flow	13,347.19	kmol/hr
H ₂ Mole Fraction	0.7015	mol/mol
4. REACTION STOICHIOMETRY & EXPANSION PARAMETERS		
Limiting Reactant	CO ₂	—
Per-Pass Conversion (X ^p)	0.30	fraction
Expansion Factor (ε _a)	-0.266	dimensionless
Initial Total Conc. (C ₀)	1,235.66	mol/m ³

Initial CO ₂ Conc. (C _{a0})	163.93	<i>mol/m³</i>
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5. KINETIC MODEL — VANDEN BUSSCHE & FROMENT (1996)

Rate Constant k ₁ (at 220°C)	1.388×10^{-4}	<i>mol/(kg·s·bar^{2.5})</i>
Rate Constant k ₂ (RWGS, at 220°C)	1.118	<i>mol/(kg·s·bar)</i>
Adsorption Const. K ₁ (K_CO ₂)	0.296	<i>dimensionless</i>
Adsorption Const. K ₂ (K_CO)	0.398	<i>bar^{-0.5}</i>
Adsorption Const. K ₃ (K_H ₂ O)	0.288	<i>bar⁻¹</i>
Equilibrium Const. K _{eq1} (MeOH)	4.219×10^{-5}	<i>bar⁻²</i>
Equilibrium Const. K_RWGS	141.57	<i>dimensionless</i>
Activation Energy Ea ₁	36,696	<i>J/mol</i>
Activation Energy Ea ₂	-94,765	<i>J/mol</i>

6. CATALYST & REACTOR SIZING

Catalyst Type	Cu/ZnO/Al ₂ O ₃ (commercial grade)	—
Catalyst Bulk Density (ρ _{bulk})	1,190	<i>kg/m³</i>
Design Catalyst Weight (W _{design})	435.38	<i>kg</i>
Safety Factor (SF)	1.20	—
Actual Catalyst Weight (W _{actual})	522.46	<i>kg</i>
Catalyst Volume (V _{cat})	0.4390	<i>m³</i>

7. TUBE & SHELL GEOMETRY

Tube Length (L)	12	<i>m</i>
Tube Inner Diameter (d)	0.046	<i>m</i>
Volume per Tube (V _{tube})	0.01994	<i>m³</i>
Number of Tubes (N _{tubes})	23	<i>tubes</i>
Total Tube Bundle Area (A _{bundle})	0.03822	<i>m²</i>
Equivalent Shell Diameter (D _{shell})	0.2206	<i>m</i>
Shell Material	SS 316L (stainless steel)	—

8. COST ESTIMATION

Sizing Parameter ($A_{\text{cost}} = V_{\text{cat}}$)	0.4390	m^3
Total Installed Cost ($\times N$ reactors)	1,459,482	USD

Table 3.2 Vertical Drum Separator

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	DS-101	—
Equipment Name	Drum Separator	—
Type	Vertical Drum	—
Function	Gas–liquid separation of reactor outlet S4	—
Material of Construction	Stainless Steel 316L	—
Design Code	ASME Sec. VIII Division 1	—
Number of Units	1 operating, 0 spare	—
B. Operating Conditions		
Operating Temperature	60	°C
Operating Pressure	40	atm
Design Pressure ($\times 1.1$)	44	atm
Design Pressure (MPa)	0.044	MPa
C. Inlet / Outlet Stream Data		
S4 Total Feed — Molar Flow	17,815.4	kmol/hr
D. Souders–Brown Vessel Sizing (C&R Vol. 6, Eq. 11.x)		
Vapour Density at DS Conditions	1.294	kg/m ³
Liquid Density (weighted avg.)	870.94	kg/m ³
Souders–Brown Coefficient K_{SB}	0.107	m/s
Maximum Vapour Velocity u_{max}	2.774	m/s
Design Velocity u_{design} (75%)	2.081	m/s

Vapour Volumetric Flow Q_V	41.59	m^3/s
Required Cross-sectional Area A	19.99	m^2
Calculated Vessel Diameter D_{calc}	5.05	m
Selected Vessel Diameter D_{sel}	5.25	m

E. Vessel Height and Liquid Residence Time

L/D Ratio (vertical drum)	4.0	—
Total Vessel Height H_{vessel}	21	m
Liquid Sump Height H_{liquid} (25%)	5.25	m
Liquid Holdup Volume V_{hold}	113.65	m^3
Liquid Volumetric Flow Q_{liq}	0.01053	m^3/s
Liquid Residence Time τ	10,788	s

F. Wall Thickness (ASME Sec. VIII Div. 1, UG-27)

Vessel Radius r	2.625	m
Allowable Stress SS316L S_{allow}	137	MPa
Joint Efficiency E_{joint}	1.0	—
Calculated Shell Thickness t_{calc}	6.42	mm
Corrosion Allowance CA	3.0	mm
Total Design Thickness t_{total}	9.42	mm

G. Equipment Cost (Turton et al., 4th ed.)

Vessel Volume for Cost A_{vessel}	454.60	m^3
Estimated Cost C_{BM} (2024 USD)	11,289,013	USD

Table 3.3 Distillation Column

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	T-101	—
Equipment Name	Distillation Column	—
Type	Sieve tray column	—
Function	Purify methanol from water: S7 feed $\rightarrow$ S8 distillate ($\geq$ 99.85 wt% MeOH) + S9 bottoms	—
Material of Construction	Stainless Steel 316L	—
Design Code	ASME Sec. VIII / BS PD 5500	—
Primary Reference	C&R Vol. 2 Ch. 11 / Perry's	—
Number of Units	1 operating, 0 spare	—
B. Feed and Product Stream Data		
Feed Stream	S7	—
Feed Total Molar Flow	1,381,996.59	mol/hr
Feed MeOH Mole Fraction z_F	0.4408	mol/mol
Feed Temperature T_{feed}	60	$^{\circ}\text{C}$
Feed Quality q	1.031	—
Distillate Flow D (S8)	592,694.75	mol/hr
Distillate MeOH Mole Fraction x_D	0.9973	mol/mol
Bottoms Flow B (S9)	756,933.90	mol/hr
Bottoms MeOH Mole Fraction x_B	0.0050	mol/mol
MeOH Recovery	99.37	%
C. Operating Conditions		
Operating Pressure	3	atm
Top Temperature (bubble pt x_D)	64.55	$^{\circ}\text{C}$

Bottom Temperature (bubble pt x_B)	100.0	°C
Design Pressure ($\times 1.1$)	0.3344	MPa

D. McCabe–Thiele Design (C&R Vol. 2, Ch. 11)

Minimum Reflux Ratio $R_{\min}$	1.699	—
Actual Reflux Ratio R ($1.5 \times R_{\min}$)	2.549	—

E. Tray Efficiency (O'Connell Correlation, C&R Eq. 11.68)

Average Relative Volatility α_{avg}	3.802	—
Average Liquid Viscosity μ_L	0.450	mPa·s
Overall Tray Efficiency E_O	43.43	%
Theoretical Stages N_{th}	20	—
Actual Number of Trays N_{actual}	47	—
Feed Tray (actual)	14	—

F. Column Diameter (Fair's Correlation, C&R Sec. 11.13)

Tray Spacing t_s	0.50	m
Vapour Density ρ_V	1.097	kg/m ³
Liquid Density ρ_L	853.76	kg/m ³
Liquid–Vapour Flow Factor F_{LV}	0.02015	—
Capacity Parameter C_{SB} (corrected)	0.0799	m/s
Flooding Velocity u_f	2.228	m/s
Design Velocity u_{design} ($80\% \times u_f$)	1.782	m/s
Net Column Area A_{net}	9.561	m ²
Total Column Area A_{total}	10.865	m ²
Calculated Diameter D_{calc}	3.719	m
Selected Diameter D_{sel}	3.75	m
% Flooding at Design	78.70	% ($\leq 85\%$ ✓)

G. Column Height (C&R Sec. 11.13)

Tray Section Height H_{trays}	23	m
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Top Disengagement	1.5	m
Bottom Sump	2.0	m
Feed Nozzle Allowance	0.5	m
Total Height without Skirt H	27	m
Skirt Height	3	m
Total Height with Skirt H _{total}	30	m
H / D Ratio	7.2	(acceptable; < 30)

H. Shell Wall Thickness (ASME Sec. VIII Div. 1)

Design Pressure P _{design}	0.3344	MPa
Shell Radius r	1.875	m
Allowable Stress SS316L	137	MPa
Weld Joint Efficiency E	1.0	—
Calculated Shell Thickness t	7.58	mm
Corrosion Allowance (CA)	3.0	mm
Shell Mass W _{shell}	3,812	kg

I. Antoine Constants ($\log_{10}(P_{\text{sat}} / \text{mmHg}) = A - B / (C + T \text{ } ^\circ\text{C})$)

MeOH Latent Heat λ 64.7 °C	35,300	J/mol
H ₂ O Latent Heat λ 100 °C	40,700	J/mol

J. Equipment Cost (Turton et al., 4th ed.)

Column Vessel Volume (cost basis)	331.34	m ³
Total Capital Investment (TCI)	37,274.200	USD

Table 3.4 Feed Preheater

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	HE-101	—
Equipment Name	Feed Preheater	—
Type	Shell & tube, 1–2 pass	—
Function	Heat S3 mixed feed from 30 °C to 220 °C	—
Material of Construction	Carbon Steel	—
Design Code	TEMA / ASME Sec. VIII	—
Number of Units	1 operating, 0 spare	—
B. Process Side (Shell Side — Cold)		
Stream	S3 — Mixed reactor feed	—
Mass Flow Rate	226,763.9	kg/hr
Molar Flow Rate	19,027	kmol/hr
Inlet Temperature T_cold_in	40	°C
Outlet Temperature T_cold_out	220	°C
Operating Pressure	50	atm
C. Utility Side (Tube Side — Hot)		
Utility Medium	HP steam / hot process gas	—
Inlet Temperature T_hot_in	350	°C
Outlet Temperature T_hot_out	320	°C
D. Thermal Design [LMTD Method]		
Heat Duty Q	31,326	kW
ΔT_1 (hot in – cold out)	50	K

ΔT_2 (hot out – cold in)	290	K
LMTD (counter-current)	199.4	K
Overall HTC U (gas–gas service)	50	W/m ² ·K
Required Heat Transfer Area	3,142	m ²
Design Area	7,877	m ²

E. Mechanical Sizing [TEMA Standards]

Tube OD	25.4	mm
Tube Wall Thickness	2.11	mm
Tube Length	4.88	m
Area per Tube	0.389	m ²
Number of Tubes N _{tubes}	20,229	—
Tube Pitch (1.25 × OD)	31.75	mm
Estimated Shell ID	3,100	mm
Estimated Cost C _{BM} (2024 USD)	1,142,278	USD

Table 3.5 Reactor Outlet Cooler

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	E-102	—
Equipment Name	Reactor Outlet Cooler	—
Type	Shell & tube, 1–2 pass	—
Function	Cool S4 from 220 °C to 60 °C for DS-101 inlet	—
Material of Construction	Carbon Steel	—

Design Code	TEMA / ASME Sec. VIII	—
Number of Units	1 operating, 0 spare	—
B. Process Side (Shell Side — Hot)		
Stream	S4 — Reactor outlet	—
Total Mass Flow	226,763.1	kg/hr
Total Molar Flow	17,815.4	kmol/hr
Inlet Temperature T_hot_in	220	°C
Outlet Temperature T_hot_out	60	°C
Operating Pressure	50	atm
Average Cp (mixed gas at ~180 °C)	33.39	J/mol·K
C. Utility Side (Tube Side — Cold)		
Utility Medium	Cooling water	—
Inlet Temperature T_cw_in	30	°C
Outlet Temperature T_cw_out	45	°C
Cooling Water Flow Rate	2,277,207	kg/hr
D. Thermal Design [LMTD Method]		
Heat Duty Q	39,661	kW
ΔT_1 (hot in – cold out)	255	K
ΔT_2 (hot out – cold in)	30	K
LMTD (counter-current)	105.1	K
Overall HTC U (gas–liquid service)	150	W/m ² ·K
Required Heat Transfer Area	2,515	m ²
Design Area (+15% fouling margin)	2,959	m ²
Tube OD	25.4	mm

Tube Wall Thickness	2.11	mm (BWG 14)
Tube Length	4.88	m (TEMA 16 ft)
Area per Tube	0.389	m ²
Number of Tubes N _{tubes}	7,598	—
Tube Pitch (1.25 × OD)	31.75	mm (triangular)
Estimated Shell ID	1,900	mm
Estimated Cost C _{BM} (2024 USD)	1,142,278	USD

Table 3.6 Distillation Condenser

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	E-103	—
Equipment Name	Distillation Condenser	—
Type	Shell & tube, 1–2 pass	—
Function	Total condenser — T-101 overhead vapour → liquid distillate S8	—
Material of Construction	Carbon Steel	—
Design Code	TEMA / ASME Sec. VIII	—
Number of Units	1 operating, 0 spare	—
B. Process Side (Shell Side — Condensing)		
Stream	T-101 overhead vapour	—
Condensing Temperature	64.55	°C
Operating Pressure	3	atm
Latent Heat at x _D λ _{top}	35,314	J/mol

C. Utility Side (Tube Side — Cold)

Utility Medium	Cooling water	—
Inlet Temperature T _{cw_in}	30	°C
Outlet Temperature T _{cw_out}	45	°C
Cooling Water Flow Rate	1,184,727	kg/hr

D. Thermal Design [LMTD Method]

Condenser Duty Q _C	20,634	kW
LMTD	30.0	K
Overall HTC U (condensing service)	800	W/m ² ·K
Required Heat Transfer Area	859.8	m ²
Design Area (+15% fouling margin)	1,011.5	m ²
Tube-side ΔP	33.36	kPa
Shell-side ΔP (design)	12.41	kPa
Design dirt factor, R _d	0.000628	m ² ·°C/W

E. Mechanical Sizing [TEMA Standards]

Tube OD	25.4	mm
Tube Wall Thickness	2.11	mm
Tube Length	4.88	m
Area per Tube	0.389	m ²
Number of Tubes N _{tubes}	2,598	—
Tube Pitch (1.25 × OD)	31.75	mm
Estimated Shell ID	1,100	mm
Estimated Cost C _{BM} (2024 USD)	507,858	USD

Table 3.7 Distillation Reboiler

A. Equipment Identification

Equipment Tag	E-104	—
Equipment Name	Distillation Reboiler	—
Type	Kettle reboiler (U-tube)	—
Function	Provide heat to T-101 bottoms; vaporize liquid for stripping section	—
Material of Construction	Carbon Steel	—
Design Code	TEMA / ASME Sec. VIII	—
Number of Units	1 operating, 0 spare	—

B. Process Conditions

Bottoms Temperature	100	°C
Steam Condensing Temperature	180	°C

C. Thermal Design (LMTD Method)

Reboiler Duty Q_R	21,287	kW
Overall HTC U (kettle reboiler)	600	W/m ² ·K
LMTD	25	K
Required Heat Transfer Area A	1,419.1	m ²
Fouling Margin (15% — TEMA)	1.15	—
Design Area A_{design}	1,632.0	m ²
Steam Flow Required	38,316	kg/hr

D. Tube-Side Geometry (TEMA 1-inch OD, BWG 14, 16 ft)

Tube OD / ID	25.4 / 21.18	mm
Tube Length	4.88	m (16 ft)
Area per Tube	0.389	m ² /tube
Number of Tubes N_t	4,195	—
Estimated Bare Module Cost (2024)	1,172,727	USD

Table 3.8 CO₂ storage tank

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	TK-101	—
Equipment Name	CO ₂ Feed Storage	—
Type	spherical vessel	—
Material	Carbon Steel (cryogenic grade)	—
Design Code	ASME Sec. VIII Div. 1 / NFPA 55	—
Number of Units	1 operating, 0 spare	—
B. Process Conditions		
Fluid	Gas CO ₂	—
Mass Flow Rate	37,202.98	kg/hr
Fluid Density	1,032	kg/m ³
Operating Temperature	30	°C
Operating Pressure	26	bar
Design Pressure (×1.1)	28.5	bar
C. Sizing		
Storage Duration	7	days
Mass to Store	6,250,101	kg
Required Volume	6,056.30	m ³
Design Volume (×1.265)	7,661.22	m ³
L/D Ratio	1.0	—
Tank Diameter D (selected)	21.5	m
Tank Height H (corrected)	21.10	m
D. Wall Thickness (ASME Sec. VIII Div. 1)		
Shell Thickness t _{ASME}	177.3	mm
Estimated Cost (2024)	1,969,617.827	USD

Table 3.9 H₂ storage tank

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	TK-102	—
Equipment Name	H ₂ Feed Storage	—
Type	Multiple bullet pressure vessels	—
Material	Carbon Steel (ASME certified)	—
Design Code	ASME Sec. VIII Div. 1 / NFPA 55	—
Number of Units	~292 bullet vessels in practice	—
B. Process Conditions		
Fluid	Compressed H ₂	—
Mass Flow Rate	5,112.56	kg/hr
Fluid Density	16	kg/m ³ (200 bar, 25 °C)
Operating Temperature	30	°C
Operating Pressure	200	bar
Design Pressure (×1.1)	220	bar
C. Storage Sizing		
Storage Duration	3	days
Mass to Store	368,104	kg
Total Design Volume	29,103	m ³
Individual Bullet Volume	≤ 100	m ³ per vessel
Approximate No. of Bullets	292	—
Bullet Diameter (each)	2.0	m (typical industrial)
Bullet Length (each)	15	m (typical industrial)
D. Wall Thickness [ASME Sec. VIII — per bullet]		
Shell Thickness t (each bullet)	≈ 80	mm
Estimated Cost C _{BM} (2024 USD)	13,131,382.879	USD

Table 3.10 Methanol Storage TK-103A / TK-103B

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tags	TK-103A & TK-103B	—
Equipment Name	Methanol Product Storage (two-tank system)	—
Type	Vertical fixed-roof atmospheric (API 650)	—
Material	Carbon Steel	—
Design Code	API 650 / API 653	—
Number of Units	2 operating (one filling, one dispatching)	—
B. Process Conditions (each tank)		
Fluid	Methanol product (≥ 99.7 mol%)	—
Mass Flow Rate per Tank	9,670.30	kg/hr
Fluid Density	791	kg/m ³
Operating Temperature	25	°C
Operating Pressure	1	bar
C. Storage Sizing (each tank)		
Storage Duration	30	days
Volume Required	8,802.3	m ³
Design Volume ($\times 1.265$)	11,134.9	m ³
L/D Ratio	1.0	—
Tank Diameter D (rounded)	24.5	m
Tank Height H (corrected)	23.62	m
D. Wall Thickness [API 650]		
API 650 Design Stress SD	160	MPa
Shell Thickness tape	14.03	mm
Cost C _{BM} per Tank (2024 USD)	3,283,566.865	USD
Combined Cost TK-103A + TK-103B	6,567,133.730	USD

Table 3.11 Wastewater Storage Tank

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	TK-104	—
Equipment Name	Wastewater Storage	—
Type	Vertical open-top atmospheric (API 650)	—
Material	Carbon Steel	—
Design Code	API 650	—
Number of Units	1 operating, 0 spare	—
B. Process Conditions		
Fluid	Water-rich bottoms S9 (0.5 mol% residual MeOH)	—
Mass Flow Rate	13,688.82	kg/hr.
Fluid Density	983	kg/m ³
Operating Temperature	25	°C
Operating Pressure	1	bar
C. Storage Sizing		
Storage Duration	7	days
Mass to Store	2,299,722	kg
Design Volume (×1.265)	2,959.5	m ³
L/D Ratio	1.0	—
Tank Diameter D (rounded)	16.0	m
Tank Height H (corrected)	14.72	m
D. Wall Thickness [API 650]		
API 650 Design Stress S _d	160	MPa
Shell Thickness t _{API}	7.10	mm
Estimated Cost	576,552.800	USD

Table 3.12 Recycle Compressor

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	K-101	—
Equipment Name	Recycle Compressor	—
Type	Multi-stage centrifugal compressor	—
Inlet Stream	S5 — Vapour Recycle	—
Material	Carbon Steel	—
Design Standard	API 617 (8th Ed.)	—
Number of Units	1 duty + 1 spare (installed, auto-changeover)	—
B. Process Conditions		
Molar Flow Rate	15,645.6	kmol/hr
Molar Flow Rate	4,346	mol/s
Inlet Temperature T _{in}	60	°C
Inlet Pressure P _{in}	3	atm
Outlet Pressure P _{out}	50	atm
Heat Capacity Ratio γ (C _p /C _v)	1.35	—
C. Compression Design		
Overall Compression Ratio r	16.67	—
Number of Compression Stages N	3	—
Stage Compression Ratio r _{stage}	2.554	—
Isentropic Work per Stage W _{is}	2,940.5	J/mol
Actual Work W _{actual} ($\eta_{is} = 0.75$)	3,920.7	J/mol
Shaft Power P _{shaft}	51,118	kW
Motor Power P _{motor}	53,809	kW
Outlet Temperature per Stage T _{out}	151.7	°C
Estimated Cost	637,658	USD

Table 3.13 CO₂ Feed Compressor

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	K-102	—
Equipment Name	CO ₂ Feed Compressor	—
Type	Multi-stage centrifugal compressor	—
Inlet Stream	S1 — CO ₂ Fresh Feed	—
Material	Carbon Steel	—
Design Standard	API 617 (8th Ed.)	—
Number of Units	1 duty + 1 spare (installed, auto-changeover)	—
B. Process Conditions		
Molar Flow Rate	845.3	kmol/hr
Molar Flow Rate	234.8	mol/s
Inlet Temperature T _{in}	30	°C
Inlet Pressure P _{in}	1	atm
Outlet Pressure P _{out}	50	atm
Heat Capacity Ratio γ (C _p /C _v)	1.29	—
C. Compression Design		
Overall Compression Ratio r	50	—
Number of Compression Stages N	3	—
Stage Compression Ratio r _{stage}	3.684	—

Isentropic Work per Stage W_{is}	3,819.1	J/mol
W_{actual} ($\eta_{is} = 0.75$)	5,092.1	J/mol
Shaft Power P_{shaft}	3,587	kW
P_{motor} ($\eta_{motor} = 0.95$)	3,776	kW
Outlet T per Stage T_{out}	133.3	°C
Estimated Cost	3,087,362	USD

Table 3.14 H₂ Feed Compressor

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	K-103	—
Equipment Name	H ₂ Feed Compressor	—
Type	Multi-stage centrifugal compressor	—
Inlet Stream	S2 — H ₂ Fresh Feed	—
Material	Carbon Steel	—
Design Standard	API 617 (8th Ed.)	—
Number of Units	1 duty + 1 spare (installed, auto-changeover)	—
B. Process Conditions		
Molar Flow Rate	2,536.0	kmol/hr
Molar Flow Rate	704.4	mol/s
Inlet Temperature T_{in}	30	°C

Inlet Pressure P_{in}	1	atm
Outlet Pressure P_{out}	50	atm
Heat Capacity Ratio γ (C_p/C_v)	1.41	—
C. Compression Design		
Overall Compression Ratio r	50	—
Number of Compression Stages N	3	—
Stage Compression Ratio r_{stage}	3.684	—
Isentropic Work per Stage W_{is}	3,996.5	J/mol
Actual Work W_{actual} (η_{is} $= 0.75$)	5,328.7	J/mol
Shaft Power P_{shaft}	11,261	kW
Motor Power P_{motor} ($\eta_{motor} = 0.95$)	11,854	kW
Outlet Temperature per Stage T_{out}	169.8	°C
Estimated Cost	1,869,880	USD

Table 3.15 Liquid Feed Pump

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	P-101	—
Equipment Name	DS-101 Liquid Feed Pump	—
Type	Centrifugal pump	—
Fluid Handled	S7 — MeOH-H ₂ O liquid (DS-101 → T-101 feed)	—
Material	Stainless Steel 316L	—
Design Standard	API 610 (12th Ed.)	—
Number of Units	1 duty + 1 standby (permanently installed, auto-changeover)	—
B. Hydraulic Design		
Mass Flow Rate	33,029.42	kg/hr
Liquid Density ρ	861.14	kg/m ³
Volumetric Flow Q	38.36	m ³ /hr
Volumetric Flow Q	0.01065	m ³ /s
Suction Pressure P _{suc}	3.04	bar
Discharge Pressure P _{dis}	3.50	bar
Pressure Head H _p	5.45	m
Static Head H _{static}	5.0	m
Friction Losses H _{friction} (10% of H _p)	0.54	m
Total Head H _{total}	10.99	m
Hydraulic Power P _{hyd}	0.989	kW

Shaft Power P_{shaft} ($\eta_{\text{pump}} = 0.70$)	1.41	kW
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Motor Power P_{motor} ($\eta_{\text{motor}} = 0.95$)	1.49	kW
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C. NPSH Verification

NPSH Available	36.84	m
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NPSH Required	3.0	m
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Estimated Cost	269,995	USD
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Table 3.16 Reflux Pump

Parameter	Value	Unit
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A. Equipment Identification

Equipment Tag	P-102	—
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Equipment Name	Reflux Pump	—
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Type	Centrifugal pump	—
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Fluid Handled	MeOH-H ₂ O condensate (T-101 overhead → reflux)	—
---------------	--	---

Material	Stainless Steel 316L	—
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Design Standard	API 610 (12th Ed.)	—
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Number of Units	1 duty + 1 standby (permanently installed, auto-changeover)	—
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B. Hydraulic Design

Mass Flow Rate	9,670.30	kg/hr
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Liquid Density ρ	791	kg/m ³
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Volumetric Flow Q	12.23	m ³ /hr
Volumetric Flow Q	0.003396	m ³ /s
Suction Pressure P _{suc}	3.04	bar
Discharge Pressure P _{dis}	3.50	bar
Pressure Head H _p	5.93	m
Static Head H _{static}	3.0	m
Friction Losses H _{friction} (10% of H _p)	0.59	m
Total Head H _{total}	9.52	m
Hydraulic Power P _{hyd}	0.251	kW
Shaft Power P _{shaft} (η_{pump} = 0.70)	0.358	kW
Motor Power P _{motor} (η_{motor} = 0.95)	0.377	kW

C. NPSH Verification

NPSH Available	38.31	m
NPSH Required	3.0	m
Estimated Cost	289,575	USD

Table 3.17 Product Pump

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	P-103	—
Equipment Name	Product Pump	—
Type	Centrifugal pump	—
Fluid Handled	S8 — Methanol product (T-101 distillate → TK-103)	—
Material	Stainless Steel 316L	—
Design Standard	API 610 (12th Ed.)	—
Number of Units	1 duty + 1 standby (permanently installed, auto-changeover)	—
B. Hydraulic Design		
Mass Flow Rate	19,340.60	kg/hr
Liquid Density ρ	791	kg/m ³
Volumetric Flow Q	24.45	m ³ /hr
Volumetric Flow Q	0.006792	m ³ /s
Suction Pressure P _{suc}	1.01	bar
Discharge Pressure P _{dis}	3.0	bar
Pressure Head H _p	25.65	m
Static Head H _{static}	8.0	m
Friction Losses H _{friction} (10% of H _p)	2.56	m
Total Head H _{total}	36.21	m

Hydraulic Power P_{hyd}	1.908	kW
Shaft Power P_{shaft} ($\eta_{pump} = 0.70$)	2.73	kW
Motor Power P_{motor} ($\eta_{motor} = 0.95$)	2.87	kW

C. NPSH Verification

NPSH Available	18.70	m
NPSH Required	3.0	m
Estimated Cost	320,520	USD

Table 3.18 Bottoms Pump

Parameter	Value	Unit
A. Equipment Identification		
Equipment Tag	P-104	—
Equipment Name	Bottoms Pump	—
Type	Centrifugal pump	—
Fluid Handled	S9 — Water-rich bottoms (T-101 → TK-104)	—
Material	Stainless Steel 316L	—
Design Standard	API 610 (12th Ed.)	—
Number of Units	1 duty + 1 standby (permanently installed, auto-changeover)	—
B. Hydraulic Design		
Mass Flow Rate	13,688.82	kg/hr
Liquid Density ρ	983	kg/m ³
Volumetric Flow Q	13.93	m ³ /hr

Volumetric Flow Q	0.003868	m^3/s
Suction Pressure P_{suc}	1.01	bar
Discharge Pressure P_{dis}	2.0	bar
Pressure Head H_p	10.27	m
Static Head H_{static}	8.0	m
Friction Losses H_{friction} (10% of H_p)	1.03	m
Total Head H_{total}	19.29	m
Hydraulic Power P_{hyd}	0.720	kW
Shaft Power P_{shaft} ($\eta_{\text{pump}} = 0.70$)	1.03	kW
Motor Power P_{motor} ($\eta_{\text{motor}} = 0.95$)	1.08	kW

C. NPSH Verification

NPSH Available	16.40	m
NPSH Required	3.0	m
Estimated Cost	260,627	USD

Table 3.19 Velocity Criteria

Parameter	Value	Unit
Liquid line design velocity	1.5	m/s (C&R Vol. 6, range 0.5–3.0 m/s)
High-pressure gas velocity (P > 10 bar)	15	m/s
Low-pressure gas velocity (P < 10 bar)	20	m/s
Pipe roughness ϵ	0.046	mm (commercial steel)
Assumed pipe length per line	50	m (conservative pressure drop estimate)
Friction factor correlation	Churchill (1977)	— (valid all flow regimes)
Pipe schedule	NPS Schedule 40	— (ASME B36.10M)

Table 3.20 Pipeline Sizing

Stream	Description	Phase / Pressure	Mass Flow (kg/hr)	NPS Selected (mm)	ΔP / 100 m (bar)
S1	CO ₂ Fresh Feed	Gas, 1 atm	37,202.98	500 mm NPS	0.00594
S2	H ₂ Fresh Feed	Gas, 1 atm	5,112.56	500 mm NPS	0.00065
S3	Mixed Reactor Feed	Gas, 50 atm	226,763.9	350 mm NPS	0.249
S4	Reactor Outlet	Gas, 50 atm	226,763.1	350 mm NPS	0.237
S5	Vapour Recycle	Gas, 3 atm	184,446.7	500 mm NPS	0.0173
S6	Purge	Gas, 3 atm	9,287.04	250 mm NPS	0.0368
S7	Dist. Feed Liquid	Liquid, 3 atm	33,029.42	100 mm NPS	0.173
S8	MeOH Product	Liquid, 1 atm	19,340.6	80 mm NPS	0.223
S9	Water Bottoms	Liquid, 1 atm	13,688.82	65 mm NPS	0.347

Table 3.21 Heat Exchanger Design

Parameter (Symbol)	Value	Unit
Equipment tag	HE-101	—
Function	Feed preheater S3: 30→220°C	—
Configuration	10 shells in series, 1-2 STHE each	—
Cold inlet T (T _{c,in})	30.00	°C
Cold outlet T (T _{c,out})	220.00	°C
Hot utility inlet T (T _{h,in})	260.00	°C
Hot utility outlet T (T _{h,out})	240.00	°C
Total heat duty (Q)	30,958.81	kW
LMTD (counterflow) (LMTD)	102.52	°C
F correction factor (F)	0.926	—
Effective LMTD (LMTD _{eff})	94.93	°C
Overall HTC (U)	50.00	W/(m ² ·K)
Total area required (A _{req})	6,522.23	m ²
Total area with fouling (A _{design})	7,500.56	m ²
Area per shell (A _{sh})	750.06	m ²
Tubes per shell (N _{t,sh})	1,957	tubes
Total tubes (all shells) (N _{t,tot})	19,570	tubes
Tube OD / ID (d _o /d _i)	0.025 / 0.0208	m
Tube length (L)	4.88	m
Bundle diameter (D _b)	1.455	m
Shell diameter (per shell) (D _{sh})	1.500	m
Tube-side vel ^o City (u _t)	10.386	m/s
F factor check	0.926	—
Material	SS 316L	—
TOTAL bare module cost — all shells (C_{BM,tot})	16,078,447	USD (2024)

3.4 Mass Balance

3.4.1 Overall Plant Mass Balance

Table 3.22 Mass Balance Overall Plant

Component	Mr	Input		Output		
		S1 CO ₂ Feed	S2 H ₂ Feed	S6 Purge	S13 CH ₃ OH Prod.	S14 Bottoms
		Mass (kg/hr)	Mass (kg/hr)	Mass (kg/hr)	Mass (kg/hr)	Mass (kg/hr)
CO ₂	44.010	38,099	0.000	7,208	0	0
H ₂	2.016	0.000	5,917	2,119	0	0
CH ₃ OH	32.042	0.000	0.000	9	18,726	119
H ₂ O	18.015	0.000	0.000	6	56	12,526
CO	28.010	0.000	0.000	3,104	0	0
Sub Total		38,099	5,917	12,446	18,782	12,645
Total (kg/hr)		44,016		43,873		

Table 3.23 Mass Balance Overall Stream Table

Component (kg/hr)	MW (g/mol)	S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12	S13	S14
Description		Fresh CO ₂ Feed	Fresh H ₂ Feed	Compressed CO ₂	Compressed H ₂	Mixed Reactor Feed	Preheated Feed	Reactor Effluent	Cooled Effluent	Recycle Vapor	Purge	Recycle	Degassed Liquid to DC	CH ₃ OH Product	Bottoms
T (°C)		30	30	40	56	60	220	220	60	60	60	42	60	65	100
P (atm)		1	1	50	50	50	50	50	49	49	49	50	3	1	3
Phase		Gas	Gas	Gas	Gas	Gas	Gas	Gas	2-Phase	Gas	Gas	Gas	Liquid	Liquid	Liquid
CO ₂	44.01	38,099	0	38,099	0	102,970	102,970	72,079	7,208	72,079	7,208	64,871	0	0	0
H ₂	2.016	0	5,917	0	5,917	24,989	24,989	21,190	2,119	21,190	2,119	19,071	0	0	0
CH ₃ OH	32.042	0	0	0	0	85	85	18,939	18,853	95	9	86	18,845	18,726	119
H ₂ O	18.015	0	0	0	0	57	57	12,645	12,588	63	6	57	12,582	56	12,526
CO	28.01	0	0	0	0	27,939	27,939	31,043	3,104	31,043	3,104	27,939	0	0	0
TOTAL		38,099	5,917	38,099	5,917	156,040	156,040	155,896	43,872	124,470	12,446	112,024	31,427	18,782	12,645

3.4.2 Equipment Mass Balances:

Table 3.24 Mass Balance Mixing Point (MP)

Component	Mr	IN S3	IN S4	IN S11	OUT S5
CO ₂	44.010	38,099	0	64,871	102,970
H ₂	2.016	0	5,917	19,071	24,989
CH ₃ OH	32.042	0	0	86	85
H ₂ O	18.015	0	0	57	57
CO	28.010	0	0	27,939	27,939
Sub Total		156,040			156,040
TOTAL		38,099	5,917	112,024	156,040

Table 3.25 Mass Balance preheater (E-101)

Component	Mr	IN S5	OUT S6
CO ₂	44.010	102,970	102,970
H ₂	2.016	24,989	24,989
CH ₃ OH	32.042	85	85
H ₂ O	18.015	57	57
CO	28.010	27,939	27,939
TOTAL		156,040	156,040

Table 3.26 Mass Balance Reactor (R-101)

Component	Mr	Input	Output
		S6 Reactor Feed	S7 Reactor Outlet
CO ₂	44.010	102,970	72,079
H ₂	2.016	24,989	21,190
CH ₃ OH	32.042	85	18,939
H ₂ O	18.015	57	12,645
CO	28.010	27,939	31,043
Sub Total		156,040	155,896
Total (kg/hr)		156,040	155,896

Table 3.27 Mass Balance cooler (E-101)

Component	Mr	IN S7	OUT S8
CO ₂	44.010	72,079	72,079
H ₂	2.016	21,190	21,190
CH ₃ OH	32.042	18,939	18,939
H ₂ O	18.015	12,645	12,645
CO	28.010	31,043	31,043
TOTAL		155,896	155,896

Table 3.28 Mass Balance Drum Separator (DS-01)

Component	Mr	Input	Output	
		S8 Feed	S9 Purge	S12 Liq. Feed
CO ₂	44.010	72,079	72,079	0
H ₂	2.016	21,190	21,190	0
CH ₃ OH	32.042	18,939	95	18,845
H ₂ O	18.015	12,645	63	12,582
CO	28.010	31,043	31,043	0
Sub Total		155,896	124,470	31,427
Total (kg/hr)		155,896	155,896	

Table 3.29 Mass Balance Purge

Component	Mr	In S9	OUT S10	OUT S11
CO ₂	44.010	72,079	7,208	64,871
H ₂	2.016	21,190	2,119	19,071
CH ₃ OH	32.042	95	9	86
H ₂ O	18.015	63	6	57
CO	28.010	31,043	3,104	27,939
			12,446	112,024
TOTAL		124,470	124,470	

Table 3.29 Mass Balance Recycle Compressor (RB-101)

Component	Mr	IN S11	OUT S11
CO ₂	44.010	64,871	64,871
H ₂	2.016	19,071	19,071
CH ₃ OH	32.042	86	86
H ₂ O	18.015	57	57
CO	28.010	27,939	27,939
TOTAL		112,024	112,024

Table 3.30 Mass Balance Distillation Column (DC-01)

Component	Mr	Input	Output	
		S12 Feed	S13 CH ₃ OH Product	S14 Bottoms
		Mass (kg/hr)	Mass (kg/hr)	Mass (kg/hr)
CO ₂	44.010	0	0	0
H ₂	2.016	0	0	0
CH ₃ OH	32.042	18,845	18,726	119
H ₂ O	18.015	12,582	56	12,526
CO	28.010	0	0	0
Sub Total		31,427	18,782	12,645
Total (kg/hr)		31,427	31,427	

3.5 Heat Balance

3.5.1 Heat Balance for Each Equipment:

Table 3.27 Heat Balance at Mixer + HX-01

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Stream S1 (CO ₂ feed, 30 C)	157,578.19	Stream S3 (Mixed feed, 30 C)	2,870,634.37
Stream S2 (H ₂ feed, 30 C)	366,097.37	Q_HX-01 (Heat Removed by CW)	14,131,572.27
Stream S5 (Recycle, 60 C)	16,478,531.07		
Total	17,002,206.64	Total	17,002,206.64

Table 3.28 Heat Balance at HE-101 (Feed Preheater)

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Stream S3 inlet (30 C)	2,870,634.37	Stream S3 outlet (220 C)	115,664,234.57
Q_HE-101 (Steam, 10 bar)	84.7×10 ⁶ .20		
Total	115,664,234.57	Total	115,664,234.57

Table 3.29 Heat Balance at R-101 (Reactor, Isothermal 220 C)

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Stream S3 (Reactor inlet, 220 C)	114,322,366.57	Stream S4 (Reactor outlet, 220 C)	109,175,553.67
Q_rxn released (net exothermic)	28,907,957.08	Q_R-101 (Heat Removed by CW)	34,054,769.98
Total	143,230,323.65	Total	143,230,323.65

Table 3.30 Heat Balance at E-102 (Reactor Outlet Cooler)

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Stream S4 inlet (220 C)	109,175,553.67	Stream S4 outlet (60 C)	19,176,010.42
		Q_E-102 (Heat Removed by CW)	89,999,543.25
Total	109,175,553.67	Total	109,175,553.67

Table 3.31 Heat Balance at DS-01 (Flash Drum Separator, Adiabatic)

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Stream S4 (60 C)	19,176,010.42	Stream S5 (Vapour Recycle)	16,478,531.07
		Stream S6 (Vapour Purge)	829,707.53
		Stream S7 (Liquid to DC-01)	1,867,771.81
Total	19,176,010.42	Total	19,176,010.41

Table 3.32 Heat Balance at E-103 (Distillation Condenser, DC-01)

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Overhead vapour (V = 1481.66 kmol/hr)	55,458,483.27	Q_E-103 (Heat Removed by CW)	55,458,483.27
Total	55,458,483.27	Total	55,458,483.27

Table 3.33 Heat Balance at E-104 (Distillation Reboiler, DC-01)

Component	Input (kJ/hr)	Component	Output (kJ/hr)
Stream S7 feed (60 C)	1,867,771.81	Stream S8 (MeOH distillate, 60 C)	971,185.71
Q_E-104 (Steam, Reboiler duty)	56,492,381.77	Stream S9 (Water bottoms, 100 C)	1,930,484.60
		Q_E-103 (Condenser, heat removed)	55,458,483.27
Total	58,360,153.58	Total	58,360,153.58

3.5.2 Total Heat Balance:

Table 3.34 Overall Heat Balance Summary

Unit Operation	H_in (kJ/hr)	Q_in (kJ/hr)	Total Output (kJ/hr)
MIXER + HX-01	17,002,206.64	-	17,002,206.64
HE-101 (Feed Preheater)	2,870,634.37	84.7×10 ⁶	115,644,234.57
R-101 (Reactor)	114,322,366.57	28,907,957.08	143,230,323.65
E-102 (Reactor Cooler)	109,175,553.67	-	109,175,553.67
DS-01 (Flash Separator)	19,176,010.42	-	19,176,010.42
DC-01 (Distillation Column)	1,867,771.81	56,492,381.77	58,360,153.58
Total	264,414,543.48	196,852,070.85	461,266,614.33

Table 3.35 Utility Summary

Equipment	Q (kJ/hr)	Q (kW)	Utility Type
HX-01 (Mixer Cooler)	14,131,572.27	3,925.44	Cooling Water
HE-101 (Feed Preheater)	84.7×10 ⁶	31,326.81	Steam (10 bar)
R-101 (Reactor Cooling)	34,054,769.98	9,459.66	Cooling Water
E-102 (Outlet Cooler)	89,999,543.25	24,999.87	Cooling Water
E-103 (Condenser)	55,458,483.27	15,405.13	Cooling Water
E-104 (Reboiler)	56,492,381.77	15,692.33	Steam (10 bar)
Total Steam (HE-101 + E-104)	167,944,113.97	46,651.14	Steam
Total Cooling (HX-01+R-101+E-102+E-103)	193,644,368.77	53,790.10	Cooling Water

CHAPTER IV

PLANT DESIGN

4.1 Chemical Plant Location

One of the most important aspects to consider when establishing a plant is the determination of its location, as this directly influences the success and efficiency of production operations. After evaluating several potential locations, the Arun Loksemawe Special Economic Zone (SEZ) in Aceh Province, Sumatra, Indonesia, was selected as the recommended factory site based on a multi-criteria quantitative scoring analysis. These factors other factors that need to be considered in determining the location of the factory can be grouped into 2, namely primary factors and secondary factors.

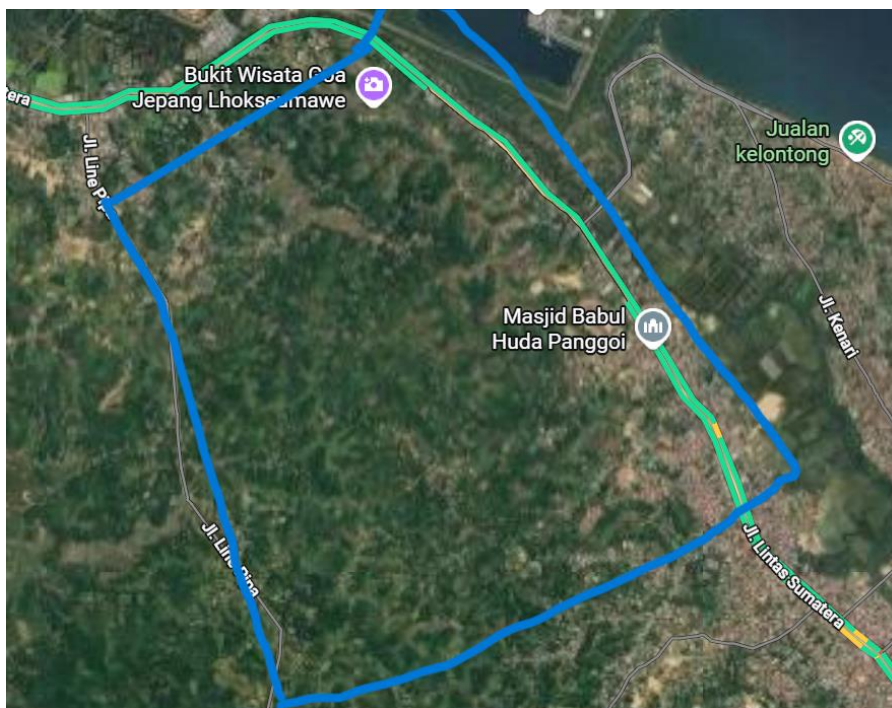


Figure 4.1 Factory Location

Factors that influence the choice of factory location include:

1. Primary Factor

Primary factors are those that directly influence the main objectives of the plant, including production activities and product distribution. These factors include:

- a. Close to Raw Materials

It would be more profitable if the location is close to material. Raw materials because it can save time and costs. The raw materials that used in establishing a Methanol factory is Syngas and water sources as the main ingredients for water electrolysis to become hydrogen. The primary raw materials for this methanol plant are carbon dioxide (CO₂), sourced from industrial flue gas via PT. Roda Manunggal Inti (RMI) Krakatau Karbonindo Nusantara, and hydrogen (H₂), produced on-site by water electrolysis. Both raw materials are available within or near the Arun Lhokseumawe Special Economic Zone, making this location particularly suitable for the proposed production process.

- b. Transportation

Transportation facilities are available in Arun Lhokseumawe Special Economic Zone (SEZ), Aceh from or to the factory has made it possible for the delivery of materials raw materials and products smoothly. Considering the location of the factory in the area Arun Lhokseumawe SEZ, Aceh is close to the deep-sea port suitable for chemical product export. Carried out for the delivery of raw materials and products can by land or sea. This is very useful and facilitate the delivery of raw materials and products and can reduce the burden of transportation costs.

The Krueng Geukueh Port (Port of Lhokseumawe), a deep-sea port that can handle chemical product tankers up to 30,000 DWT, serves the SEZ. The factory produces 150,000 tons of methanol annually, which necessitates frequent bulk exports to both domestic and foreign markets. This port is essential to the export strategy.

The Trans-Sumatera Highway, which connects Lhokseumawe to Banda Aceh (280 km west) and Medan (280 km south), provides road access, allowing for the distribution of domestic products and the transportation of raw materials by land. Although the SEZ does not currently have direct access to rail connectivity, dedicated industrial routes within the SEZ that connect to the highway network may serve as a supplement.

2. Secondary Factors

It is a factor that directly influences the means which improves the performance of factory management, namely including production process and worker welfare. Factors that included in secondary factors are:

a. Utilities

Supporting facilities or utilities needed such as water, electricity and fuel can be obtained quite easily, because it is one of the areas located in area industry. Water requirements for the plant can be fulfilled using seawater, which is abundant due to the plant's proximity to the coastline. Access to basic utilities is made possible by the Arun Lhokseumawe SEZ. The PLN (Perusahaan Listrik Negara) Aceh grid supplies electricity, and the industrial zone has additional power generation capability. Within the available SEZ power allocation, the

plant's anticipated electrical requirement for compressors, pumps, lighting, and instrumentation is between 2- 5 megawatts.

b. Government Policy

Regarding government regulation for industrial area in Lhokseumawe, Aceh, the development of industrial activities in Indonesia is governed by Government Regulation (PP) Number 20 of 2024 concerning Industrial Zoning, which replaced the earlier regulation on industrial estates and expanded industrial development to include comprehensive spatial planning, infrastructure support, and fiscal and non-fiscal facilities. This regulation mandates that industrial activities °Occur within designated industrial zone and aligns them with regional spatial plans. In addition, Prerendering No. 26 Tahun 2025 provides requirements for industrial park standards and accreditation to ensure adequacy of infrastructure and services. These regulations provide the legal framework for siting and operating industrial facilities such as the proposed 150,000 t/y methanol plant in the Arun Lhokseumawe Special Economic Zone.

c. Labor and Expert Staff

The availability of skilled and semi-skilled labor in Aceh, backed by universities and vocational schools like Syrah Kuala University, guarantees sufficient human resources for plant operation. It is possible to hire knowledgeable staff by working with overseas companies. In the meanwhile, the number of employees will be determined by taking into account the abilities and demands that are specific to the organization.

d. Social Conditions

The state of society is an important aspect because the condition of the community in the factory environment will be very influencing the establishment of a factory. To support and get support from the community, then the establishment of this factory at least have a positive impact or provide benefits for the community, as well as providing facilities around the factory which enables people/employees to live decently.

e. Housing and Other Facilities

Employees require housing facilities. The business may build a dorm or other housing, rent it out, or, in the event that it is unable to supply this demand, the business may by offering (renting) accommodation near the factory location, this can collaborate with the local community.

f. Climate

Climate can affect worker productivity, climate that good can increase work productivity, and vice versa. The climate referred to here is air humidity, wind, heat rays sun, etc. The climate variations in Lhokseumawe can be said to be appropriate for industrial areas.

g. Soil Condition

With the establishment of this factory in Lhokseumawe, it shows that the type and structure of the existing soil can indeed be used for establishing a factory or in accordance with the industrial area. And The condition of the land in Lhokseumawe is also included in the soil structure which is stable.

Apart from primary and secondary factors, there are also special factors which need to be considered such as:

a. Factory Waste

It is also necessary to pay attention to waste from factories, especially the impact to the health of the community around the factory location. This could be done by making a waste disposal site in a certain tanks and flows, specifically for the waste disposal process factory waste without polluting the surrounding environment.

b. Control of flood and fire hazards

Building a factory far from residential areas and avoiding the establishment of factories in flood-prone areas are two ways to achieve this. This is meant to prevent unanticipated events from spreading to the local population and hurting numerous people.

4.2 Plant Layout

Factory layout is a component of plant design that governs the arrangement of buildings, process areas, equipment, offices, and utilities, thereby ensuring smooth, safe, and efficient plant operations.

The flow of processes and activities of existing workers becomes basic considerations in arranging buildings in a factory so that the process can run effectively, safely and continuously.

There are several factors that need to be considered in determining *factory layout* includes:

- 1) Ease of process and operations adapted to ease of equipment maintenance and ease of control production result.
- 2) occupational safety
- 3) There is possibility of factory expansion
- 4) Appropriate and economical distribution of utilities

- 5) Fairly free movement between process equipment and other equipment that stores hazardous material
- 6) Effective and economical use of space
- 7) Problems with processing factory waste so that it does not interfere or pollute the environment

Based on the above factors, the layout of the Methanol factory is arranged for the placement of buildings in the factory area, including:

1. Process Area

The area where the Methanol production process takes place taking place, this area is placed in a location that makes it easy *supply* of raw materials from product storage areas as well facilitates monitoring and repair of tools.

2. Storage Area

Raw materials and products produced are stored in the area. In this case, the storage is placed in an area that is easy to reach by transport equipment.

3. Utility Area / Supporting Facilities

The area where production support equipment is located such as water, electricity, heating, and waste treatment facilities.

4. Administration and Office Area

It is the center of factory administration activities for matters with external and internal parties in the factory.

5. Laboratory Area

As a place for research and development as well as a place for product and raw material *quality control*.

6. Public Facilities

Like share facilities in general, which consist of canteen, parking lot, medical clinic, and medical facilities such as prayer room. The placement of this facility aims to provide a sense of comfort for employees to utilize these facilities.

7. Expansion Area

The purpose of this area is for future factory needs future. Such as increased production capacity due to product improvement.

8. Factory Maintenance and Care Area

Area used to carry out maintenance activities and requirement repairs according to factory needs. This area can refer to as the workshop (*maintenance*) area.

The plant is divided into ten functional zones as described in Table 4.2. The overall orientation of the plant is designed with the prevailing north-east monsoon wind direction in mind: process units are located in the central area, administrative buildings and control room are positioned to the north-east (upwind), and the flare stack and storage tanks are positioned to the south-east (downwind).

Table 4.1 Plant Zone Description, Position, and Safety Distances

Zone	Zone Name	Position on Plot	Justification	Minimum Safety Distance	Major Equipment / Facilities
1	Process Area (ISBL)	Centre-West	Core of plant; process flow west-to-east; central access for operations	≥ 15 m from admin; ≥ 30 m from storage tanks (NFPA 30 §22.4)	Methanol Synthesis Reactors R-101A/B; Separators DS-01, DS-02; Distillation Column MD-01; Heat Exchangers HE-101 to E-104;

					Compressors C-101, C-102
2	Raw Material Storage (TK-101, TK-102)	North-West	Downwind of admin; adjacent to feed point; away from fired sources	≥ 30 m from process area; ≥ 60 m from admin building (NFPA 59A)	CO ₂ Storage Tanks TK-101 A/B; H ₂ Cylinder Banks TK-102
3	Product Storage (TK-103, TK-104)	South-East (downwind, leeward)	Prevailing NE wind carries vapors away from process area; near port access road	≥ 30 m from fired heaters; ≥ 15 m from process area (NFPA 30 §22.4)	Methanol Product Tanks TK-103 A/B (API 650); Crude Methanol Buffer Tank TK-104; Loading Pump P-101 A/B
4	Utility Area	South-West	Separate from process; central access to all units; near power supply entry point	≥ 15 m from process units (IP Model Code Part 2)	Cooling Tower CT-101; Boiler/Steam Generator B-101; Power Substation E-SUB; Water Treatment Plant WTP-101; Instrument Air Compressor AC-101
5	Control Room & Administration	North-East (upwind)	Upwind of process area (NE monsoon direction); outside hazardous area Zone 2; blast-resistant construction	≥ 30 m from process units; ≥ 60 m from storage tanks (IP Model Code §5.4)	Main Control Room (DCS); Administration Offices; Meeting Rooms; Safety Office (HSE); Medical Clinic
6	Flare Stack	South-East corner (downwind)	Maximum downwind distance from personnel areas; safe radiation zone maintained	Flare exclusion zone radius ≥ 90 m at ground level; ≥ 15 m from any structure (API 521)	Emergency Relief Flare Stack FL-101; Knockouts Drum KO-101

7	Laboratory	North-East, adjacent to admin	Near admin for non-production staff; at safe distance from hazardous area	≥ 30 m from process area boundary	Quality Control Laboratory; Environmental Monitoring Station
8	Workshop & Maintenance	West (adjacent to process area)	Close to process units for rapid response; separate building to prevent contamination	≥ 15 m from hazardous area (IP Model Code §5.4)	Maintenance Workshop; Tool Store; Spare Parts Warehouse
9	Public Facilities (Canteen, Mosque, Parking)	North (main gate side, upwind)	Non-production staff facilities at plant boundary; upwind of process area	Outside battery limits; ≥ 50 m from process area	Canteen; Prayer Room (Mush olla); Employee Parking; Security Post; Main Gatehouse
10	Future Expansion Area	East (boundary)	Reserved for capacity increase; does not block existing operations	To be designated when needed	Reserved Land (minimum 20% of total plant area)

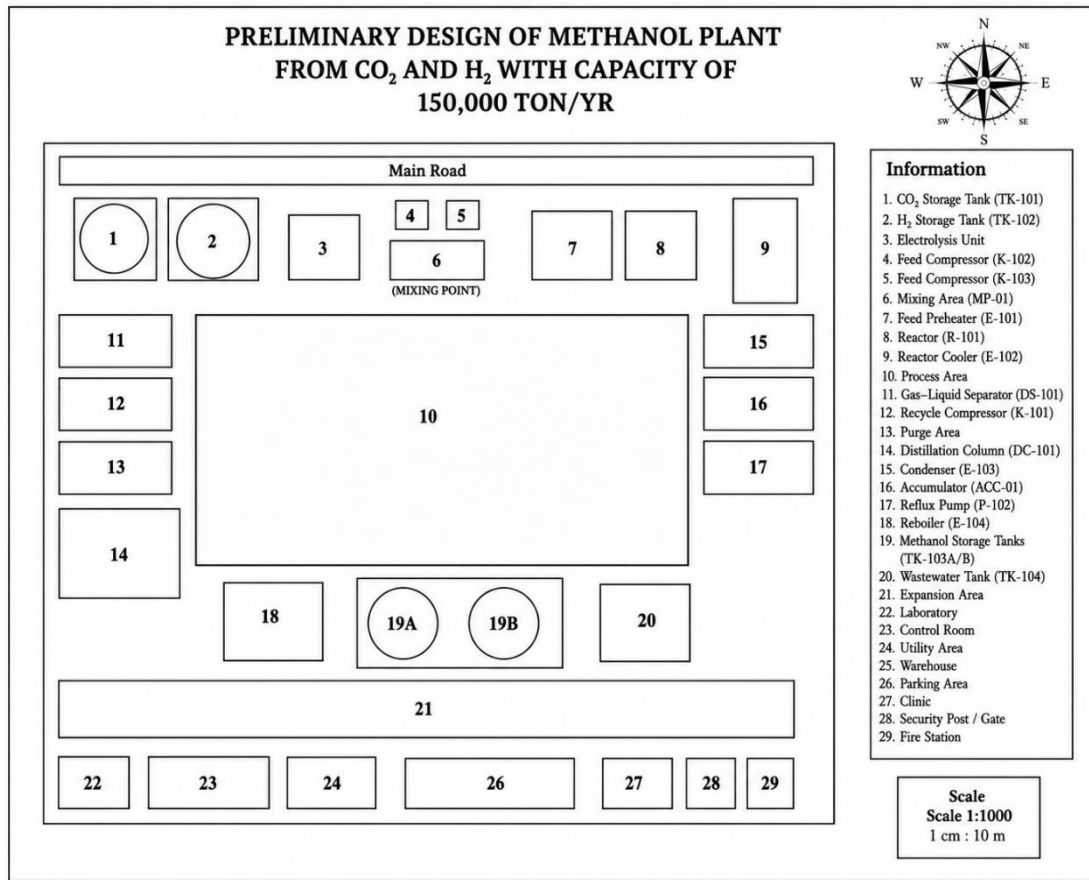


Figure 4.2 plant layout

4.3 Equipment Layout

The equipment layout specifies how each piece of process equipment is physically arranged inside the process area (ISBL). The layout is intended to guarantee the following: safe inter-equipment spacing for maintenance and emergency access; gravity-assisted flow where practical (e.g., distillation column to reflux drum); logical process flow sequence from feed compression through reaction, separation, and purification to product storage; and effective piping routing to reduce capital costs.

4.3.1 Equipment Arrangement Principles:

To ensure efficiency, safety, operational effectiveness, and accessibility for inspection and maintenance. Several crucial elements must be considered while planning the layout of process equipment. These factors include:

- 1) Flow of Raw Materials and Products: Economic efficiency is greatly increased by an ideal flow of raw materials and products. The associated handling costs decrease with the proximity of raw materials and completed goods to transportation routes.
 - 2) Airflow: To maintain sufficient ventilation and avoid the buildup of dangerous gases, proper airflow in and around the process equipment must be guaranteed.
 - 3) Lighting: To guarantee safety and operational correctness, there must be enough illumination throughout the whole facility, paying special attention to high-risk locations.
 - 4) Traffic Flow: To avoid congestion, guarantee efficient manufacturing processes, and preserve worker safety, the flow of workers and goods must be properly organized.
 - 5) Economic Considerations: Production equipment should be positioned to minimize operating costs while maintaining plant efficiency and safety.
- Spacing Between Process Equipment: Equipment operating under high temperature and pressure should be positioned separately from other process units.

When designing equipment layouts, the following guidelines are used:

1. Sequential configuration according to the flow of the process
2. Sufficient distance for upkeep and security
3. When appropriate, vertical location for gravity flow

4. Appropriate access for equipment replacement and lifting

4.3.2 Reactor and Heat Exchange Section

To effectively remove heat from the exothermic reaction, the methanol synthesis reactor is situated close to the heat exchangers. In order to prevent thermal runaway and maintain the ideal reactor temperature, cooling systems are included.

4.3.3 Separation and Purification Section

The reactor exit is located downstream of the gas-liquid separator. With enough room for tray or packing maintenance, the distillation column is positioned vertically. Streams of recycled material are effectively returned to the reactor intake.

4.3.4 Storage and Transfer Section

Methanol storage tanks are located at the edge of the production area and connected to loading facilities for transport. Transfer pumps and piping systems are designed to minimize leakage and fire hazards.

Table 4.2 Major Equipment Items

Tag	Equipment Name	Grid Position (Col–Row)	Elevation	Zone
R-101A/B	Methanol Synthesis Reactor (Fixed-Bed)	D5–E5 (twin units)	Grade + elevated on structure	Process Area
C-101 A/B	Fresh Feed Compressor (3-stage centrifugal)	B3–B4	Grade (compressor house)	Process Area
C-102 A/B	Recycle Gas Compressor	B5	Grade (compressor house)	Process Area
HE-101	Feed/Reactor Outlet Heat Exchanger	D4	Grade (shell-and-tube)	Process Area
E-102	Product Cooler (S4 → DS-01)	D6	Grade	Process Area
E-103	Condenser (MD-01 overhead)	H5 (elevated on column platform)	Elevated at column top (approx. 15 m)	Process Area
E-104	Reboiler (MD-01 bottom)	H7 (at column base)	Grade, adjacent to column skirt	Process Area
DS-01	High-Pressure Flash Separator	E6	Grade (on concrete saddle)	Process Area
MD-01	Methanol Distillation Column	H5–H7 (vertical)	Grade to top of column (~20 m)	Process Area
TK-101 A/B	CO ₂ Feed Storage Tanks (refrigerated horizontal bullets)	A1–B2	Grade (NW corner)	Raw Material Storage
TK-102	H ₂ Cylinder Storage Banks	A3	Grade (in ventilated cage)	Raw Material Storage
TK-103 A/B	Methanol Product Storage	I7–J8	Grade (SE)	Product Storage
TK-104	Crude Methanol Buffer Tank	G6	Grade	Process Area
P-101 A/B	Methanol Product Pump	I8 (at grade below TK-103)	Grade (below tank suction)	Product Storage
P-102 A/B	Reflux Pump	H5 (at column base)	Grade (below condenser outlet)	Process Area
FL-101	Emergency Flare Stack	J1 (SE corner)	Elevated (min. 30 m above grade)	Flare Area
KO-101	Flare Knockouts Drum	J2 (adjacent to flare)	Grade	Flare Area
CT-101	Cooling Tower (induced draft)	C9	Grade (SW, utility area)	Utility Area
B-101	Package Steam Boiler	D9	Grade (utility area)	Utility Area
AC-101	Instrument Air Compressor	D8	Grade (utility area)	Utility Area

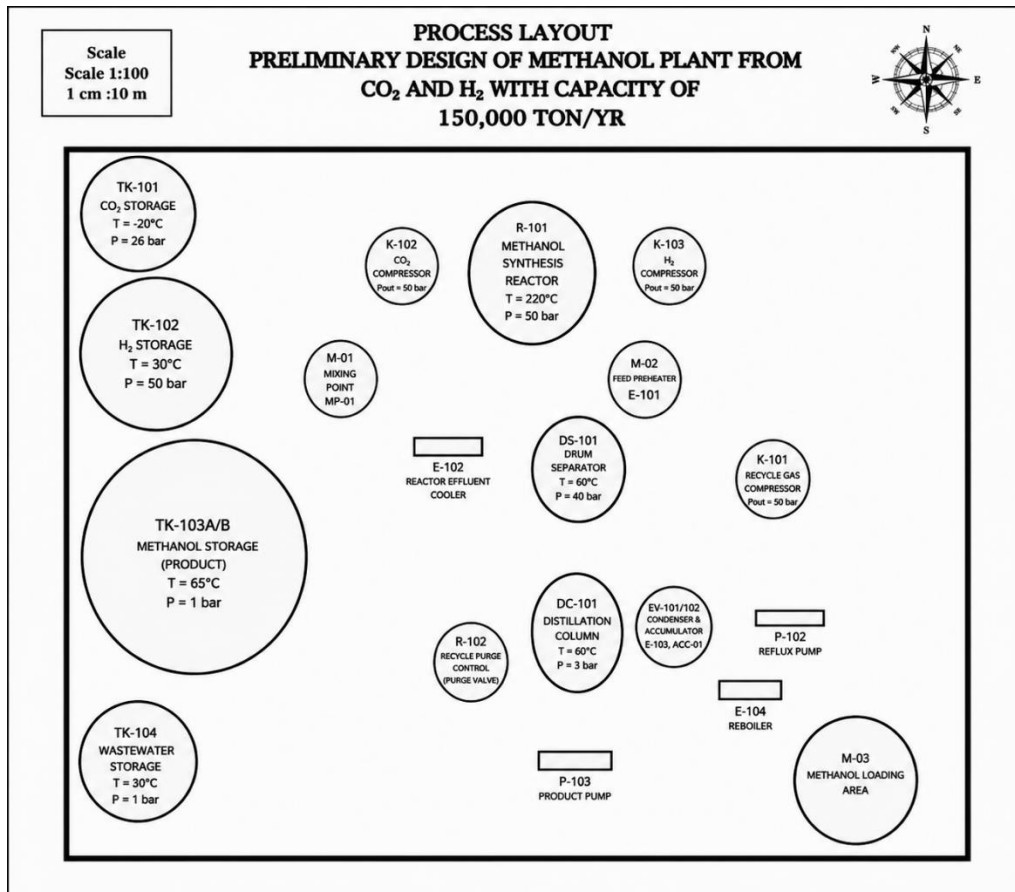


Figure 4.3 Process layout

4.4 Plant Organization

An organization is a system consisting of sub-systems or parts that are interrelated with each other in carrying out activity. This activity is not a temporary activity or just for a moment, but it is an activity that has a pattern or sequence that is carried out in a relatively regular and repetitive manner. Organization often interpreted as a group that together wants to achieve a common goal (Prion, 2007).

Anthony (1995), explains that an organization is a group of people who interact carry out various activities in a coordination to achieve goals, where basically individuals do not can achieve goals individually. This means that the organization's goals can be achieved through the arrangement/management carried out on several people as the executor of organizational work. As with organizations in a company/organization in an industry.

The success of a company/industry in achieving its goals very dependent on the management *of* the organization which includes planning, implementation and control, division of authority and responsibility.

4.4.1 Company Form

The proposed 150,000-ton-per-year methanol factory from SYNGAS and H₂ will be founded as a Limited Liability Company (PT), a type of business that obtains its funding through the sale of shares in which each partner contributes one or more shares. Shares are securities issued by the business or PT, and the owner of the shares has invested money in the business, indicating ownership. Shareholders in a limited business are only obligated to deposit the full amount specified in each share. The limited liability company structure was chosen for a number of reasons, including the following:

1. It is easy to get capital, namely by selling shares company
2. Shareholders' liability is limited so that the business runs smoothly production is solely handled by company management.
3. The owners and managers of the company are separate from each other. The company is the shareholder, while the management The company is a board of directors and staff who are supervised by the board of directors' commissioner.

4. The company's survival is more assured, because it does not affect by the termination of shareholders, directors along with staff and company employees.
5. Management efficiency. Shareholders can choose the people as a board of commissioners and capable directors and experienced.
6. Wider business field. A limited liability company can attract large capital from the community, so it can expand business.

4.4.2 Company Organizational Structure

An organization is a container or tool where people who have one vision carry out activities to achieve goals expected. Organizational structure is a systematic description about the duties and responsibilities and relationships between the parts in the company. The organizational structure of a company can various according to the shape and needs of each company. The leadership levels in this company are as follows:

- a. Shareholders
- b. Board of commissioners
- c. President Director
- d. Director
- e. Head of Department
- f. Head of Section
- g. Employees and Operators

As the company's owners, shareholders in A director, with assistance from an Operational Manager and a Manager of Finance and General Affairs, oversees the day-to-day operations of the corporation, while a Councilor Commissioner represents it. where the operations manager is in charge of field production, utilities, development, maintenance, and quality control. In the meantime, the field marketing, administration, general and security, health, occupational safety, and environmental departments are managed by the general and

financial managers. Each head the section will oversee several sections (Supervisors) and each will supervise and oversee several employees or staff companies in their respective fields. Company employees will divide into several groups of teams led by each squad leader, where the squad leader will be responsible to supervisors in each section.

The benefits of having this organizational structure are as follows:

1. Explain the authority, limitations of duties, responsibilities, and authority.
2. As official orientation material
3. More appropriate employee placement
4. Preparation of management development programs
5. Rearrange applicable work steps and work procedures if it is proven to be less smooth.

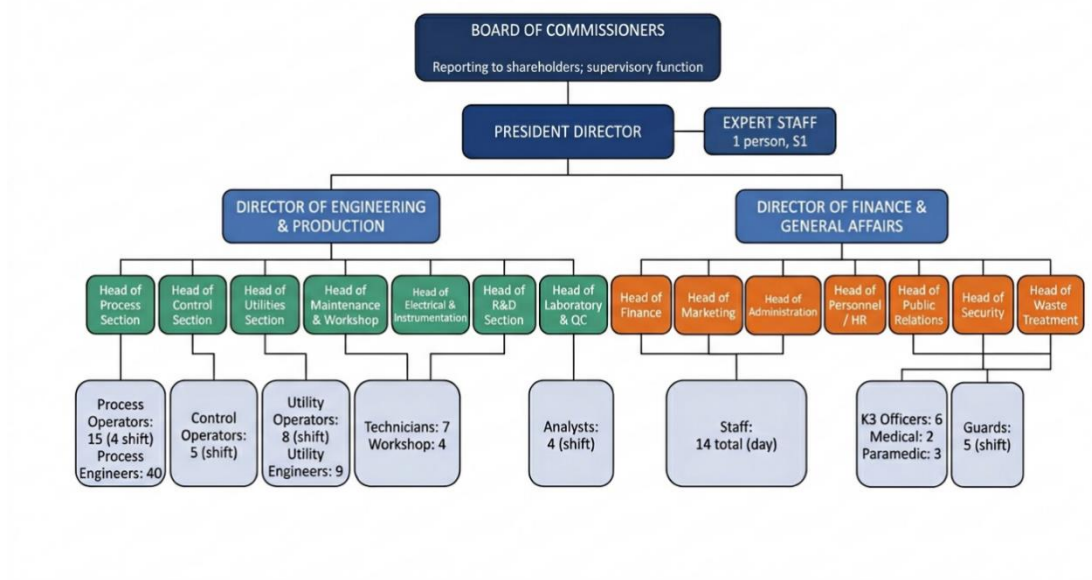


Figure 4.4 Corporate Organization

4.4.3 Duties and Authorities:

4.4.3.1 Shareholders

Shareholders (company owners) are several people which collect capital for the purposes of establishment and the running of the company's operations. The highest authority in companies that have the form of limited liability companies is a general meeting of shareholders. At the general meeting shareholders:

1. Appoint and dismiss the Board of Commissioners
2. Approve business results and balance sheets annual profit and loss of the company

4.4.3.2 Board of Directors:

The Board of Commissioners represents the interests of the shareholders and acts as the supervisory body over the management of the company. The duties of the Board of Commissioners include:

1. Evaluating and approving general policy plans, profit targets, and the allocation of funding sources
2. Supervising the performance of the President Director
3. Providing guidance to company management on important strategic matters

4.4.3.3 President Director

The main director is the highest leader in the company, and is fully responsible for the progress and decline company. The President Director is responsible to the Board of Directors. Commissioner for all actions and policies that have been appointed as the company's leader. President Director under the Director of Production and Engineering, as well as the Director Finance and General.

The main director is in charge of:

- a) Director of Engineering and Production

The duties of the Technical and Production Director are to lead implementation of factory activities related to production and operations, engineering, development, equipment maintenance,

procurement, and laboratories.

b) Director of Finance and General Duties

The Director of Finance and General Affairs is responsible to problems related to administration, personnel, finance, marketing, public relations, security and work safety.

4.4.3.4 Expert Staff

Expert staff consists of experts whose job is to assist the board of directors in carrying out his duties both in relation to technical and administrative. Expert staff are responsible to main director according to their respective fields of expertise each.

Duties and authorities:

- 1) Provide advice and suggestions in planning company development.
- 2) Improve the process of the factory or planning of tools and production development.
- 3) Increase work efficiency.

4.4.3.5 Head of Section

In general, the Head of Section's duties are to coordinate, organize and supervise the implementation of work in the environment of its parts according to the given lines by the company leadership. The head of the department can also act as a director's staff. The head of this section is responsible to respective directors. The section heads

consist of:

1) Head of Process and Utilities Section:

Coordinating factory activities in the field process and provision of raw materials and utilities.

2) Head of Maintenance, Electrical, and Instrumentation Section:

Responsible for maintenance activities and facilities supporting production activities.

3) Head of Research, Development and Control Division Quality:

Coordinate related activities with research, corporate development, and quality control.

4) Head of Finance and Marketing: Coordinating marketing and

procurement activities goods, as well as financial bookkeeping.

5) Head of Administration Section: Responsible for activities that

related to administration, personnel and corporate household.

6) Head of Public Relations and Security: Responsible for activities

that relationship between companies and society and maintain company security.

7) Head of occupational Health, Safety and Environment Section:

Responsible for factory security and employee health and safety

4.4.3.6 Head of Section

The section head is the implementer of work within the environment its parts according to the plan that has been arranged by the Head of each Section. Each section head is responsible to the respective section heads in accordance with

- 1) Head of Process Section: Directly lead and monitor the smooth running of the project production process.

- 2) Head of Raw Materials and Products Section: Responsible for providing materials raw materials and maintain the purity of raw materials, as well as control the products produced.
- 3) Head of Utilities Section: Responsible for water supply, steam, fuel, and compressed air are good for processes and instrumentation.
- 4) Head of Maintenance and Workshop Section: Responsible for maintenance and maintenance activities replacement of tools and supporting facilities.
- 5) Head of Electrical and Instrumentation Section: Responsible for providing electricity and the smooth operation of instrumentation tools.
- 6) Head of Research and Development Section: Coordinating activities that related to increased production and overall process efficiency.
- 7) Head of Laboratory and Quality Control Section: Organizing quality control for materials raw materials, auxiliary materials, products and waste.
- 8) Head of Finance Section: Responsible for bookkeeping and other matters relating to company finances.
- 9) Head of Marketing Section: Coordinating product marketing activities and procurement of factory raw materials.
- 10) Head of General Administration Section: Responsible for activities that related to corporate households and office administration.
- 11) Head of Personnel Section: Coordinate related activities with staffing.
- 12) Head of Public Relations Section: Organizing related activities

with corporate relations, government, and public.

13) Head of Security Section: Organizing activities related to directly supervise security issues company

14) Head of occupational Health and Safety Section: Manage employee health issues and family, as well as addressing safety issues work in a company.

15) Head of Waste Processing Unit Section: Responsible for factory waste so that in accordance with waste quality standards.

4.4.6 Personnel System

a) Working System

The Methanol Plant is planned to operate 330 days a year with 24 hours a day. The working day of the production unit is the working day from Monday to Sunday. The remaining days that are not holidays will be used for repair or maintenance and shut down. Use Today aims to maintain the smooth running of the production process as well administration and marketing mechanisms, then employee working hours arranged with shift and non-shift systems.

b) Shift System

The shift work schedule applies to employees in the department production units and is carried out in rotation. The division of labor is divided in 4 groups, where each group will work according to with time between shifts in one week. Schedule arrangement shift work can be seen in Table below

Table 4.3 Shift Working Hours

Shift	Working Hours
I	08.00 – 16.00
II	16.00 – 00.00
III	00.00 – 08.00

Table 4.4 Two-Week Shift Rotation Schedule (4 Teams: A, B, C, D)

	1	2	3	4	5	6	7	8	9	10	11	12
A	D	D	D	D	D	H	A	A	A	A	A	H
B	N	N	N	N	H	D	D	D	D	D	H	A
C	A	A	A	H	N	N	N	N	N	H	D	D
D	H	H	H	A	A	A	H	H	H	N	N	N

c. Non-Shift System:

D = Morning Shift (07.00 – 15.00)

A = Afternoon Shift (15.00 – 23.00)

N = Night Shift (23.00 – 07.00)

H = Holiday

d. Workforce Details:

In human resource management (HR), the first thing that needs to be done is to conduct a job analysis to a position in a company organization. Next, compile the job

description details so that all company activities are covered in the job description, not there may be positions that overlap or are not included include. The need to make details of the number of workers is so find out how many workers there are the factory, and make it easier to check the workforce because the amount is certain (already calculated) beforehand.

The salary is determined based on the position within organization and expertise. Workforce requires job specifications concerning education level, work ability (skills), type gender and others to get the right person on the right.

Wags System:

Labor wages are adjusted to the class of worker depending on the population in the organizational structure and length of time working at the company. The wages received by employees consist of from:

- Basic salary.
- Position allowance.
- Attendance allowance (transportation) for non-shift staff.
- Health benefits with the provision of a company doctors and hospitals that have been appointed by the company for all employees according to their positions.

Due to the different employee groups, the wage system is divided into 3, namely:

- 1) Monthly system: Given to permanent employees
- 2) Daily system: Given to daily workers such as

direct laborers or workers who are needed only from time to time

- 3) Contract system: Given to contract workers and the amount is not fixed, depending on the type of work being done. Usually required during generations. Apart from regular salary, employees who work overtime are also given extra salary.

additional with calculation:

- Overtime on Sundays/holidays: For every hour, the amount is twice the hourly wage.
- Overtime on weekdays: For every hour, the amount is one and a half times the hourly wage.
- If employees are called to work at the factory outside of working hours work, will also be given additional salary.

Based on position, salary/wages of labor in the factory, it can be seen in Table below:

Table 4.5 Employee Salary Table

No.	Position	No. of Employees	Monthly Salary (Rp)	Monthly Subtotal (Rp)	Education Level
1	President Director	1	Rp 80.000.000	Rp 80.000.000	S2
2	Director of Engineering and Production	1	Rp 65.000.000	Rp 65.000.000	S2
3	Director of Finance and General Affairs	1	Rp 65.000.000	Rp 65.000.000	S2
4	Expert Staff	1	Rp 15.000.000	Rp 15.000.000	S1
5	Head of General Affairs Division	1	Rp 18.000.000	Rp 18.000.000	S1
6	Head of Marketing Division	1	Rp 18.000.000	Rp 18.000.000	S1
7	Head of Finance Division	1	Rp 18.000.000	Rp 18.000.000	S1
8	Head of Engineering Division	1	Rp 18.000.000	Rp 18.000.000	S1
9	Head of Production Division	1	Rp 18.000.000	Rp 18.000.000	S1
10	Head of R&D Division	1	Rp 18.000.000	Rp 18.000.000	S1
11	Head of Personnel Section	1	Rp 15.000.000	Rp 15.000.000	S1
12	Head of Public Relations Section	1	Rp 15.000.000	Rp 15.000.000	S1
13	Head of Security Section	1	Rp 15.000.000	Rp 15.000.000	S1
14	Head of Purchasing Section	1	Rp 15.000.000	Rp 15.000.000	S1
15	Head of Marketing Section	1	Rp 15.000.000	Rp 15.000.000	S1
16	Head of Administration Section	1	Rp 15.000.000	Rp 15.000.000	S1
17	Head of Cash/Budget Section	1	Rp 15.000.000	Rp 15.000.000	S1
18	Head of Process Section	1	Rp 15.000.000	Rp 15.000.000	S1
19	Head of Control Section	1	Rp 15.000.000	Rp 15.000.000	S1
20	Head of Laboratory Section	1	Rp 15.000.000	Rp 15.000.000	S1
21	Head of Utilities Section	1	Rp 15.000.000	Rp 15.000.000	S1
22	Head of Development Section	1	Rp 8.000.000	Rp 8.000.000	S1
23	Head of Research Section	1	Rp 8.000.000	Rp 8.000.000	S1
24	Personnel Employees	3	Rp 5.000.000	Rp 15.000.000	D3
25	Public Relations Employees	3	Rp 8.000.000	Rp 24.000.000	D3
26	Security Employees	5	Rp 8.000.000	Rp 40.000.000	SMA/SMK
27	Purchasing Employees	4	Rp 8.000.000	Rp 32.000.000	D3
28	Marketing Employees	4	Rp 8.000.000	Rp 32.000.000	D3
29	Administrative Employees	3	Rp 8.000.000	Rp 24.000.000	D3
30	Cash/Budget Employees	3	Rp 8.000.000	Rp 24.000.000	D3
31	Process Employees	40	Rp 8.000.000	Rp 320.000.000	D3/S1
32	Control Employees	5	Rp 8.000.000	Rp 40.000.000	D3/S1
33	Laboratory Employees	4	Rp 8.000.000	Rp 32.000.000	S1
34	Maintenance Employees	7	Rp 8.000.000	Rp 56.000.000	SMA/SMK
35	Utility Employees	9	Rp 10.000.000	Rp 90.000.000	D3/S1
36	Process Operators	15	Rp 10.000.000	Rp 150.000.000	SMA/SMK
37	Utility Operators	8	Rp 8.000.000	Rp 64.000.000	SMA/SMK
38	K3/HSE Employees	6	Rp 8.000.000	Rp 48.000.000	D3/S1
39	R&D Employees	3	Rp 8.000.000	Rp 24.000.000	D3
40	Secretary	5	Rp 8.000.000	Rp 40.000.000	S1
41	Medical Doctor	2	Rp 9.000.000	Rp 18.000.000	S1
42	Paramedic	3	Rp 7.000.000	Rp 21.000.000	D3
43	Driver	6	Rp 5.000.000	Rp 30.000.000	SMA
44	Cleaning Service	5	Rp 5.000.000	Rp 25.000.000	SMA
	TOTAL	166	—	Rp 1.663.000.000	

	TOTAL ANNUAL SALARY COST (× 12 months)			Rp 19.956.000.000	
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4.4.5 occupational Health Safety and Environment(K3/HSE)

occupational Health and Safety (K3) is one of the important factors that must be considered, K3 is a system that ensure that the process and process actors can run safely, and can reduce the risk of accidents that can °Ccur in the area factory operations. occupational safety includes machines, work tools, materials used in the production process, workplace processing and the environment that concerns all production or distribution processes of goods and services.

K3's goal is to prevent and reduce losses brought on by dangers, environmental damage, and fire. However, what is meant by an accident (injury/disability, death) that affects workers in connection with their jobs, beginning at home and ending at work along the customary route, including illnesses brought on by work-related activities.

4.4.5.1 Work Safety Foundations

The basis for occupational safety includes:

1. Idea: Pancasila
2. Structural: 1945 Constitution Article 27 paragraph 2

"Every citizen has the right to work and a decent living for humanity"

3. Operational:

- MPR Decree
- Employment Laws/Regulations
- Regulation of the Minister of Manpower

Law concerning basic provisions regarding labor is Law No.14 of 1969. Several

The important articles listed therein are:

- 1) Article 3: "Every worker has the right to work and decent income for humanity “
- 2) Article 8: "Every worker has the right to receive protection safety, health, morality, work ethics and treatment consistent with human dignity.
- 3) Article 10: "The government promotes good work protection.

Work safety norms (Law No.1 of 1979)

Health norms of company hygiene (Law No.12 years 1948)

Work norms (Civil Code, BK.III Chapter 71, etc.)

Provision of compensation, care and rehabilitation in the event of work accidents (Law No. 33 of 1947)

The law concerning occupational safety objectives is the Law

No. 1 of 1970 which contains:

- Protect workers' rights to safety in doing work for the welfare of life and increase national production and productivity.

- Maintain production sources and use them safely and efficiently.
- Ensure the safety of everyone inside workplace with the existence of established laws and regulations occupational safety. requirements, namely:
 - Prevent and reduce work accidents;
 - Prevent, reduce and extinguish fires;
 - Prevent and reduce the danger of explosions;
 - Provide an opportunity or way to save oneself during fire or other dangerous events;
 - Provide assistance in the event of a work accident;
 - Provide personal protective equipment to worker
 - Prevent and control the emergence or spread of the extent of temperature, humidity, dust, dirt, smoke, steam, gas, wind gusts, weather, radiation rays, sound, and vibration.
 - Prevent and control the emergence of diseases caused by work, both physical and psychological work, poisoning, infection and transmission;
 - Obtain sufficient appropriate lighting;
 - Maintain good air temperature and humidity;
 - Provide sufficient air freshening;
 - Maintain cleanliness, health and order;
 - Obtaining harmony between labor, work tools, environment, working methods and processes;
 - Secure and expedite loading and unloading work, treatment and storage of goods;
 - Prevent exposure to dangerous electric currents;
 - Adjust and improve security on jobs where the risk of accidents increases tall:

- Workers must receive protection from various things around him that can affect and disturb him in carrying out his work
- Provide information if requested by the supervisor or K3 expert;
- Use personal protective equipment;
- Express objections to the work being done if the K3 requirements for personal protective equipment are provided does not guarantee work safety;
- Successful implementation of occupational safety and health (K3) based on the K3 management policy taken by company leaders, including the following:
 - Firm leadership
 - K3 organization within the company's organizational structure
 - Adequate facilities and infrastructure
 - Integration of K3 into all company functions
 - Support all employees in implementing K3

The means of achieving K3 management is to minimize accidents accompanied by increased productivity high so that the company's goals can be achieved effectively optimal.

4.4.5.2 Hazardous Materials and Their Prevention

Hazardous substances are substances that during manufacture, processing, transportation, storage and its use can emit gas, dust, radiation and other forms that can cause irritation, radiation, blindness, explosion, corrosion, poisoning. Hazardous materials in certain amounts can cause damage to the

equipment. Materials

or hazardous tools that must be taken into account are:

1. Materials that are flammable and can explosive
2. Toxic substances that are harmful to health
3. Mechanical tools that can be dangerous work safety.

Considering some of the raw materials used and the resulting product is toxic, so take action. The safeguards that must be taken are:

- 1) Installation of symbols or symbols used as warning labels for each type of chemical used;
- 2) Storage in a closed room and provision of ventilation;
- 3) Provide masks, body protection, gloves, head protection (helmet), safety boots, and protection eye.
- 4) Providing training before starting work and making standard operating procedures that must be followed by everyone operator;
- 5) If spilled, use absorbent paper to absorb it.
- 6) Absorbent paper and clothing contaminated with these materials
The toxic material is isolated with airtight plastic air.
- 7) Contaminated surfaces (floors, clothing, etc.) washed with 60-70% ethanol, then washed with water;
- 8) If eye contact occurs, for those who wear contact lenses, The lenses should be removed, and the eyes rinsed with water. for 20-30 minutes, then immediately check with a doctor;
- 9) If it comes into contact with skin, flush the skin that has come

into contact with the toxic material with water, then wash with soap. If this happens irritation, see a doctor immediately.

In addition to preventive measures to deal with danger the above-mentioned toxic materials, the factory operators also need to be equipped with ear protection (earplugs), for protect ears from loud noises from factory equipment.

To avoid fires caused by electric currents, the following things need to be considered:

1. To prevent fires, fire extinguishers are provided.

some firefighting equipment such as fire box & indoor fire hydrants, as well as fire extinguishing units.

2. Use lightning protection for high equipment.
3. Using insulation on the electrical network.
4. Supervision of installed cables.
5. Installation of electrical installations does not hinder work.

To avoid dangers caused by tools production, some things that need to be considered include:

1. Use of process equipment that exceeds capacity;
2. Employ trained operators;
3. Conduct inspections and maintenance of production

equipment periodically.

4. Create a manual control system and automatically on each unit, making it easier control in case of danger.

From research into accidents that occurs,

It turns out that human factors are the cause of accidents very large, where this is caused by negligence people in complying with work safety regulations, such as:

1. Activities that deviate from the regulations.
2. Not using work safety equipment.
3. Improper use of tools to anticipate the above matters, it is advisable factory management takes the following actions:
 1. Hold training or training regarding the nature and Dangers in the factory.
 2. Use protective equipment in the factory premises.
 3. Attach labels or danger symbols to make it easier recognition of the dangers of chemicals.
- Dangers posed by nature Hazards posed by nature include: floods, earthquake, hurricane, or lightning. To prevent the occurrence of dangers caused by nature, several things are necessary carried out include:
 - 1) Establish a factory with a strong foundation.
 - 2) Install lightning rods on buildings and process equipment the high.
 - 3) Install a notification alarm that works automatically if a natural disaster occurs.

- 4) Establish a factory in a location where it can be avoided dangers such as floods and earthquakes and provide a safe area within the factory premises.

Table 4.6 Methanol Physical and Chemical Hazard Data

Hazard Property	Value	Significance
Flash Point (closed cup)	11°C	Flammable at ambient temperature — requires flame-proof equipment throughout storage area
Lower Explosive Limit (LEL)	6.0 vol%	Any leak in confined space can create explosive atmosphere rapidly
Upper Explosive Limit (UEL)	36.5 vol%	Wide flammability range
Autoignition Temperature	464°C	Well above reactor operating temperature; not an autoignition risk in normal operation
Boiling Point	64.7°C	Low boiling point — significant vapor generation at ambient temperatures
TLV-TWA (ACGIH)	200 ppm (skin notation)	occupational exposure limit — all work areas must be below this threshold
STEL (Short-Term Exposure Limit)	250 ppm	15-minute maximum; TWA exceedance triggers immediate evacuation
Invisible Flame	Burns with invisible flame in daylight	CRITICAL: Fire may not be visually apparent; thermal imaging or detector required for fire detection
Toxicity Route	Inhalation, ingestion, skin absorption	Systemic toxicity — liver damage, optic nerve damage, blindness at high doses
IDLH (NIOSH)	6,000 ppm	Immediately Dangerous to Life and Health threshold

4.4.6 Factory Environment Settings

The arrangement of the factory environment is also an important factor. This affects work safety; attention needs to be paid specifically in factory environmental settings there is a physical environment and work environment

1. Physical Environment:

- a) Includes work equipment machines and raw materials:
Layout arrangement machines and tools in such a way that workers can carry out work with freedom of production and safety.
- b) Planning of factory machines and equipment by taking into account safety factor;
- c) The quality of the materials and equipment purchased is guaranteed.

2. Work Environment

- a. Place the machines in an orderly manner so that the distance between the machines is sufficient wide.
- b. clean factory yard.
- c. Sufficient lighting in the factory environment;
- d. Placement of unused materials or waste in its place;
- e. Installation of alarm systems and danger signs such as fire detectors and its instruments;
- f. Factory environment equipped with air ventilation sufficient and equipped with an exhaust fan to maintain air circulation.

CHAPTER V

UTILITIES

Utilities are a very important part in supporting production process in a chemical industry. Likewise, a process production in an industry will not run well without its utilities. The design of the utility unit plays a crucial role in maintaining the continuous operation of the plant. Several utility systems are required for a plant. There are several units in planning utilities so that the utility needs can be met, including:

- A. Water Supply and Treatment Unit
- B. Steam Generator Unit
- C. Power Generation Unit
- D. Compressed Air Supply Unit
- E. Fuel Supply Unit
- F. Waste Treatment Unit

Since Arun Lhokseumawe Special Economic Zone (SEZ) is situated in an industrial area, it is relatively easy to access necessary utilities or supporting facilities like fuel, power, and water. Within the SEZ, industrial utility systems and regional power plants facilitate the delivery of electricity. Because the factory is close to the seaside, seawater is readily available to meet the plant's water needs.

In the context of a chemical plant, utility systems are the group of auxiliary services that supply the process units with water, energy, inert gases, and the ability to handle waste. Unlike the core process equipment, they provide the infrastructure required for the process equipment to run profitably and safely rather than directly converting raw materials into completed goods.

5.1 Water Supply and Treatment Unit

In general, industrial water requirements are met using groundwater and seawater as sources. Boiler feed water and domestic water used comes from sea water which is not

far from the factory location. To prevent fouling in heat exchange equipment, river water treatment must be carried out both physically and chemically, before further processed, and seawater is desalinated before entering the factory. The reason for using seawater to meet the needs process support is due to the following factors:

- a. Sea water can be obtained in large quantities
- b. Easy to set up and process
- c. Sea location close to the factory

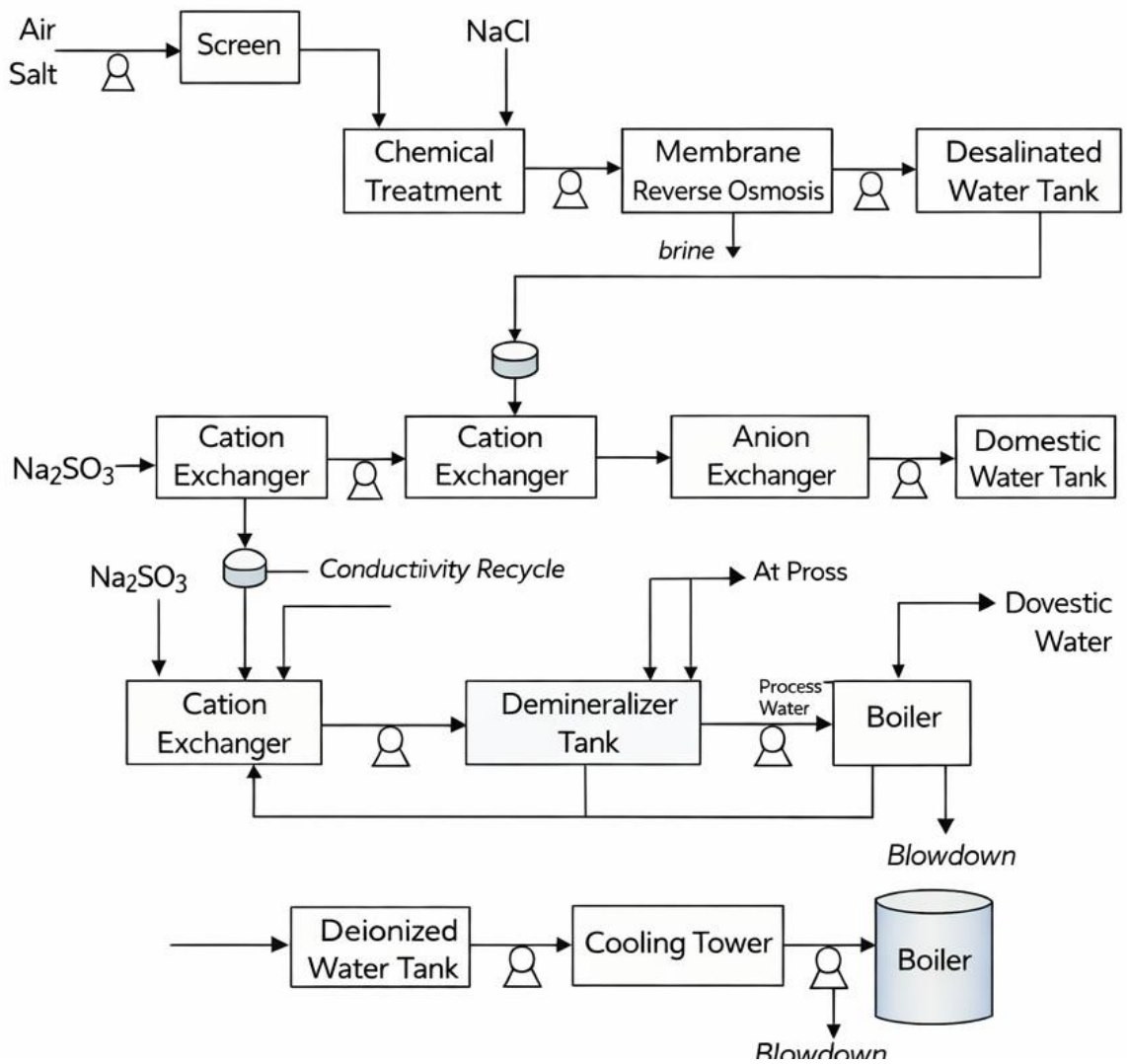


Figure 5.1 Seawater Treatment

5.1.1 Water Treatment Unit:

The seawater treatment process in the utility section can be processed using processing flow diagram in Figure below:

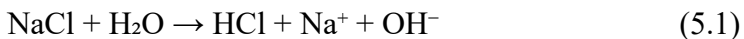
The system consists of screening, chemical pretreatment, reverse osmosis (RO), dichlorination, ion exchange demineralization, deaeration, and storage tanks. The overall objective of this treatment is to produce water that meets the quality requirements for process utilities and steam generation.

1 Seawater Intake and Screening:

Seawater is withdrawn from the intake system and passed through a coarse screen followed by a fine screen to remove suspended solids such as debris, shells, and marine organisms. This step protects downstream pumps, membranes from mechanical damage and fouling.

2 Chemical Pretreatment (Chlorination):

After screening, seawater is sent to a chemical treatment tank where sodium hypochlorite (NaCl) is dosed for disinfection. Chlorination prevents biological growth and biofouling in pipelines and RO membranes. Sodium hypochlorite reacts with water to form hypochlorous acid (HCl), which acts as the primary oxidizing biocides.



This pretreatment improves membrane performance and reduces biofilm formation on the RO surface (Baker, 2012).

3 Reverse Osmosis Desalination:

A high-pressure pump is used to pressurize the pretreated seawater before it is supplied into the reverse osmosis (RO) unit. By supplying pressure higher than the osmotic pressure of saltwater, usually between 55 and 70 bar, RO uses a semi-permeable membrane to extract dissolved salts from water. Two streams are produced:

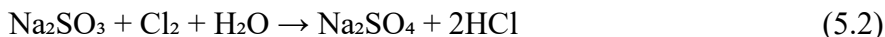
- a) Permeate: low-salinity desalinated water.
- b) Brine: a concentrated stream of rejected material.

For seawater applications, RO systems usually achieve water recovery of 35–45% and salt rejection of 99.2–99.7%. The water is fit for additional polishing and utility use when the total dissolved solids (TDS) drop from about 35,000 ppm to less than 500

ppm (Elimelech & Phillip, 2011). The desalinated water stored in a water tank to provide ensure stable supply to downstream units.

4 Dichlorination:

Dichlorination is carried out by adding sodium sulfite (Na_2SO_3), which reacts with chlorine according to the following reaction:



To preserve demineralization efficiency and safeguard the ion exchange resins, the residual chlorine content is lowered to less than 0.1 mg/L. (Baker, 2012).

5 Ion Exchange Demineralization:

Demineralization is performed using a cation exchanger followed by an anion exchanger to remove dissolved ionic species from the desalinated water. The DM water is stored in a demineralized water tank for use in steam generation and process applications.

6 Deaeration:

Then the demineralized water is sent to a deaerator to remove dissolved oxygen and carbon dioxide, that cause corrosion in boiler tubes and steam systems. In short, Seawater is first filtered to remove suspended solids and then chlorinated to prevent biological contamination. Pre-purified seawater is desalinated by reverse osmosis to produce permeate and brine. The permeate is stored and dechlorinated before desalination by ion exchange. Desalinated water is degassed and used as boiler feed water, and some of the desalted water is used for cooling towers and domestic drinking water supplies. Boiler and cooling tower blowdowns are vented to maintain acceptable levels of dissolved solids. This integrated treatment system ensures that the plant is supplied with water quality suitable for all utility needs while minimizing scale formation, corrosion, and contamination. (Elimelech and Phillip, 2011).

5.1.2 Identify Water Requirements

The overall water demand in this plant is categorized into several uses, including:

a. Sanitary Water

According to WHO standards, the daily water requirement per person ranges between 100–120 liters. For industrial or office environments, the water demand per person is

set at 120 liters per day (Sularso, 2001). This plant has 166 employs, resulting in a total domestic water need are:

$$120 \times 166 = 830 \text{ kg/hr} \quad (5.3)$$

b. Service Water

Service water is used for general facility needs such as mosques, cafeterias, laboratories, and other utilities. It is set at 10% of the sanitary water demand, amounting to 222 Kg/hr.

$$830 \times 10\% = 83 \text{ kg/hr} \quad (5.4)$$

c. Fire Hydrant Water

According to NFPA 15 standards, fire hydrant water is required for 12 hours at a rate of 260 kg/min, resulting in a total hydrant water demand of 15600 Kg/hr.

$$260 \times 60 = 15,600 \text{ kg/hr} \quad (5.5)$$

d. Cooling Water

The cooling water requirements for each piece of equipment in this plant are presented in Table 5.1.

Table 5.1 Cooling Water Requirement

Equipment	Code	Amount (kg/hr)
Reactor Effluent Cooler	CL-01	1,850,000
Syngas Intercooler	CL-02	720,000
Methanol Condenser	CD-01	1,980,000
Distillation Condenser	CD-02	1,150,000
Compressor Aftercooler	CL-03	650,000
Purge Gas Cooler	CL-04	380,000
Total Cooling Water Circulation		6,730,000
a make-up water of 3%,		201,900
Total Cooling Water required		6,931,900

The total cooling water requirement is 6,730,000 kg/hr, with a make-up water requirement of 3%, amounting to 201,900 kg/hr. GPSA guidelines So, the total need for this is

$$6,730,000 + 201,900 = 6,931,900 \text{ kg/hr} \quad (5.6)$$

The Total need of water in this unit can be calculated in the table below:

Table 5.2 Total Plant Water Demand Summary

No	Information	Water Requirements (kg/hr)
1	Sanitary Water	830
2	Service Water	83
3	Fire Hydrant Water	15,600
4	Cooling Water	6,931,900
5	Total	6,948,339

5.2 Steam Generation Unit:

The steam generation unit provides the thermal energy required for process heating in the methanol plant. The main steam consumers include the methanol purification reboiler, feed preheaters, start-up heating systems, and general plant heating services. Steam is selected as the heating medium due to its high latent heat, constant condensation temperature, ease of distribution, and good heat transfer characteristics (Towler & Sinnott, 2013). The steam requirements for each piece of equipment in this methanol plant design are detailed in the following table:

Table 5.3 Steam Requirement

Equipment	Code	Amount (kg/hr)
Methanol Distillation Reboiler	RB-01	9,500
Light Ends Column Reboiler	RB-02	4,800
Heavy Ends Column Reboiler	RB-03	3,900
Deaerator Heating Steam	DA-01	800
Tank Heating Coils	HT-01	1,200
General Process Heating	PH-01	2,600
Total Steam Demand		22,800

The make-up water required for steam generation is 10% of the total steam demand:
 Make-up water = $10\% \times 22,800 \text{ kg/hr} = 2,280 \text{ kg/hr}$ Total boiler water requirement.

$$22,800 + 2,280 = 25,080 \text{ kg/hr} \quad (5.7)$$

Steam distributes from the boiler house via an insulated main steam at 20 bar. Every steam consumer specified is supplied by a separate branch line. Condensate recovery lines minimize heat losses and make-up water use by returning condensate to the deaerator. According to Smith (2016), the projected condensate return rate is 80% of the total steam produced, which is in line with appropriate condensate management practices.

Table 5.4 Boiler Specification

Parameter	Value	Basis / Description
Boiler type	Fire-tube package boiler	Suitable for $\leq 30 \text{ ton/h}$ steam capacity (Kern, 1950)
Capacity	25,080 kg/hr	Total demand + 10% standby margin
Pressure	20 atm	Saturated steam
Temperature	212–215 °C	Saturation temperature at 20 bar
Total heat duty	13.52 MW	From steam latent heat
Heating surface area	1,620 m ²	(Q/q), heat flux = 30,000 kJ/h·m ²
Boiler efficiency	85 %	Typical industrial package boiler (Perry)
Fuel type	Natural gas	Plant fuel
Fuel consumption	1,145 kg/hr	Based on LHV = 50,000 kJ/kg
Boiler feedwater flowrate	26,334 kg/hr	Steam + 5% blowdown
Blowdown rate	5 %	Standard value (Perry)
Shell diameter	2.6 m	From volume–area relation
Shell length	7.8m	(L = 3D) assumption
Boiler volume	46.3 m ³	From heating surface density
Steam outlet quality	Dry saturated	With steam separator
Safety devices	Safety valve, low-level trip, flame detector	Boiler code requirement

5.3 Power Generation Unit

The power supply unit is responsible for providing electricity to meet the demands of both the plant and office facilities. The electricity required for the methanol plant is supplied by the State Electricity Company (PLN), with a generator serving as a backup source such as, Krakatau Daya PLTU Electric and diesel generator. This setup ensures a continuous power supply, even in the event of disruptions from PLN. The generator used is an alternating current (AC) generator. The main grid supply from PLN is connected at 20 kV, which is the typical medium-voltage industrial supply in Indonesia. It is then stepped down via a main transformer to 6.6 kV for distribution to large motor drives (compressors) and then to 380 V three-phase for smaller motors and general plant supplies. The safety instrumented system (SIS), critical analyzers, and distributed control system (DCS) all have their own UPS (Uninterruptible Power Supply), selected based on the following considerations:

- a) The generated electrical power is sufficiently high.
- b) Voltage levels can be adjusted (increased or decreased) as required.

The electricity within the plant is utilized for various purposes, including:

- a) Process and utility operations
- b) Lighting
- c) Air conditioning (AC)
- d) Laboratory, workshop, and other facility operations
- e) Instrumentation and control equipment

The Power generator is considered as an important unit, because a loss of electrical power can cause serious operational disruptions in the plant, including:

- a) Trip compressors
- b) Stop feed supply
- c) Cause reactor shutdown

In daily operations, PLTU electricity is used 100%. But if the power goes out, the operation will use up 100% diesel powered electricity. This electricity is distributed and divided into three categories.

- a) Process Equipment need:

Table 5.5 Process Equipment Power

Equipment	Power (kW)	Count	Total (kW)
Syngas compressor (H ₂ + CO ₂)	8,500	1	8,500
Recycle gas compressor	1,200	1	1,200
Methanol product pump	45	2	90
Distillation reflux pump	30	2	60
Distillation bottoms pump	35	2	70
Feed water circulation pump	25	2	50
Total Process Load			9,970

b) Utilities Equipment

Table 5.6 Utilities Equipment Power

Equipment	Power (kW)	Count	Total (kW)
Cooling water pump	250	3	750
Cooling tower fan	110	3	330
Boiler feedwater pump	90	2	180
Condensate pump	20	2	40
Instrument air compressor	150	2	300
Service water pump	35	2	70
Fire water pump (electric)	160	1	160
Total Utilities Load			1,830

c) Supporting Equipment:

Table 5.7 Supporting Equipment Power

Equipment	Power (kW)	Count	Total (kW)
Plant lighting	120	1	120
Control system (DCS/PLC)	80	1	80
Laboratory equipment	40	1	40
Workshop equipment	60	1	60
HVAC (control room & offices)	200	1	200
Administration building load	75	1	75
Total Supporting Load			575

d) Backup generator

Table 5.8 Backup Generator Specification

Parameter	Specification
Type	Krakatau Daya PLTU diesel generator
Diesel or Gas engine	Diesel engine
Capacity	4,500 kW (4.5 MW)
Voltage	6.6 kV
Frequency	50 Hz
Fuel consumption	0.24 L/kWh ($\approx$ 1,080 L/h at full load)
Startup	Automatic start, < 15 seconds

The overall electricity needs in this methanol plant design with almost 12,375 KW/day can be seen in below table:

Table 5.9 Overall Electricity Needs

Category	Total (kW)
Process	9,970
Utilities	1,830
Supporting	575
Connected Load	12,375

5.4 Compressed Air Supply Unit

Compressed air is required to operate pneumatic equipment, such as valve actuation for opening and closing mechanisms. It is also used for instrumentation systems to ensure accurate and reliable operation. The air processing system ensures that the compressed air is moisture-free, dry, oil-free, and free from solid particles to maintain system efficiency and prevent contamination. For the preliminary design of this methanol plant, the estimated compressed air requirements are:

Pressure: 7 atm

Temperature: 35 – 40 °C

Flow rate: 64.5 m³/h

Table 5.10 Compressor Specification

Parameter	Value	Description
Compressor type	Oil-free rotary screw	Instrument air requirement (Perry)
Number of units	3 (2 duty + 1 standby)	Reliability design
Total system capacity	64.5 m ³ /h	Plant requirement
Discharge pressure	7 atm	Standard instrument air pressure
Suction pressure	1 atm	Atmospheric
Discharge temperature	90–110 °C	Before after-cooler
After-cooler outlet temperature	35–40 °C	Cooling water system
Specific power	0.11 kW	Perry's typical value
Operating power	46 kW	2 compressors running
Control system	Load/unload with pressure control	Energy efficient
Air quality	Oil-free, particulate < 1 µm	ISA standard

The compressed air system is designed to supply 64.5 m³/h at 7 bar to meet instrument and plant air requirements. Air demand was estimated based on the number of control valves, analyzers, and service air usage, with allowances for leakage and future expansion. Three oil-free rotary screw compressors are provided in a 2 duty + 1 standby configuration to ensure continuous operation. Meeting instrument air quality standards (Perry; ISA).

5.5 Fuel Supply Unit

5.5.1 Purge Gas Composition (S6)

The fuel supply unit is in charge of supplying fuel to the plant's boiler and any fueled heaters. The main source of fuel gas recovery in this methanol plant is the purge stream S6, which is eliminated from the recycle loop at the DS-01 drum separator to avoid the buildup of CO and inert components. The mass balance of stream S6 is given in the table below:

Table 5.11 Stream S6 Purge Gas Composition (from Mass Balance)

Component	Flow (kg/hr)	Flow (kmol/h)	Mole Fraction	Lower Heating Value (kJ/mol)
CO ₂	3,720.3	84.53	0.1073	0 (non-combustible)
H ₂	1,097.4	544.35	0.6909	241.8
CH ₃ OH	351.7	10.98	0.0139	726.1
H ₂ O	45.6	2.53	0.0032	0 (non-combustible)
CO	4,072.0	145.38	0.1845	283.0
Total	9,287.0	787.77	1.0000	—

The purge gas stream S6's blended lower heating value (LHV) can be determined by adding the mole-fraction-weighted total of each, using the below equation

$$\text{LHV} = \sum (y^i \times \text{LHV}^i) \quad (5.8)$$

$$\text{LHV} = (0.1073 \times 0) + (0.6909 \times 241.8) + (0.0139 \times 726.1) + (0.0032 \times 0) + (0.1845 \times 283.0) \quad (5.9)$$

$$\text{LHV} = 167.0 + 10.1 + 52.2 = 229.3 \text{ kJ/mol} \quad (5.10)$$

The total mole flow of S6 is 787.77 kmol/h, and the total energy content of the purge gas is:

$$Q_r = \dot{n}_r \times \text{LHV}_{\text{f}}^{\text{e,D}} = 787.77 \times 1000 \text{ mol/h} \times 229.3 \text{ kJ/mol} \div 3,600 \text{ s/h} \quad (5.11)$$

$$Q_r = 50,130 \text{ kW} = 50.1 \text{ MW} \quad (5.12)$$

The purge gas has enough energy to fuel the boiler at full capacity with a significant excess, given the boiler's total heat duty of 13.52 MW in table 5.4. The extra energy from the purge gas (about 36.6 MW) can export to the fuel gas header of the plant for heating process, supplemental burning, or, if no other use is found, safe flared. Additionally, to lowering site CO₂ emissions, using purge gas as fuel instead of flaring offers a substantial financial advantage.

5.5.2 Distribution and Header of Fuel Gas

The purge stream S6 from DS-01 has collected by a fuel gas header and then supplied to the boiler burner at a controlled pressure of 1.5–3.0 bar. If DS-01 liquid levels are high, entrained liquid droplets may be removed from the purge stream by a fuel gas knockout drum upstream of the burner. In the case that purge gas flow is insufficient

during startup or shutdown, natural gas from the plant fuel supply is available as a backup fuel source.

5.6 Waste Treatment Unit

Wastewater treatment is mandatory under Indonesian environmental law and is also an operational necessity for this plant. Stream S9, the water-rich bottoms stream from the distillation column DC-01, which is the main source of effluent in this methanol facility. According to the mass balance, stream S9 includes 120.3 kg/hr of methanol (0.88 weight percent methanol, or around 8,800 mg/L) and has a total flowrate of 13,688.8 kg/hr. Wastewater treatment is required prior to disposal because this methanol concentration is significantly higher than the allowable discharge limit (usually ≤ 10 mg/L COD-equivalent under Indonesian Government Regulation PP No. 22/2021 on Water Quality). The table below show thee wastewater source in this plant:

Table 5.12 Wastewater Sources and Estimated Flowrates

Source	Stream / Origin	Estimated Flow (kg/hr)	Key Contaminants
Distillation bottoms	S9 from DC-01	13,688.8	Methanol 0.88 wt% (~8,800 mg/L)
Cooling tower blowdown	CW system	~67,300 (1% of circulation)	Dissolved solids, bi°Cide residuals
Boiler blowdown	Boiler BFW system	~1,254 (5% of steam)	High dissolved solids (TDS)
Sanitary wastewater	Personnel facilities	~83	BOD, suspended solids
Total WWTP feed	—	~82,400 kg/hr	Multi-contaminant

5.6.1 Treatment Strategy:

Since methanol is a water-soluble and biodegradable organic compound, and it is the main contaminant in S9, steam stripping the concentrated S9 stream to recover methanol for recycling is the recommended treatment method. This is followed by biological treatment of the combined wastewater streams for final polishing before discharge. Because methanol is volatile in comparison to water, steam stripping of S9 is especially successful. By using a stripping column at 1 bar with steam as the stripping medium can reduce methanol in the bottoms water to below 50 ppm. Economic value

can be recovered by recycling the stripped methanol vapor into the distillation column feed.

Before being released, cooling tower and boiler blowdowns must be softened or diluted due to their high concentration of dissolved inorganic particles such as calcium, magnesium, etc. Treatment options include pH adjustment, coagulation/flocculation, and final precipitation, depending on the final concentrations and the receiving waterbody classification under Indonesian regulations. Before entering the site drainage system, sanitary wastewater from the staff facilities is treated in a special package biological treatment unit (septic tank + biofilter).

5.6.2 Waste Water Treatment Process (WWTP) :

The hydraulic retention time (HRT) method is used to determine the size of the biological treatment basin. The four streams make up the whole wastewater input to the WWTP, with a combined flowrate of:

$$F_{WW} = 13,688.8 + 67,300 + 1,254 + 83 \approx 82,326 \text{ kg/hr} \quad (5.13)$$

Assuming the density of wastewater $\rho = 1,000 \text{ kg/m}^3$, then the volumetric flowrate is $Q_{WW} = 82.3 \text{ m}^3/\text{h}$. Applying a standard aerobic biological treatment HRT of 24 hours (Metcalf & Eddy, Wastewater Engineering), the minimum aeration basin volume is:

$$V_{WWTP} = Q_{WW} \times \text{HRT} = 82.3 \text{ m}^3/\text{h} \times 24 \text{ h} = 1,975 \text{ m}^3 \quad (5.14)$$

As a result, the WWTP is built with a minimum biological basin volume of roughly $1,975 \text{ m}^3$. The treated effluent is intended to meet the discharge standard of PP No. 22/2021 and the industrial wastewater quality criteria set by the Ministry of Environment of the Republic of Indonesia by achieving a methanol/COD-equivalent concentration of $\leq 10 \text{ mg/L}$ after biological treatment and final clarification.

CHAPTER VI

ECONOMIC EVALUATION

In this preliminary chemical plant design, an economic analysis is required to estimate the feasibility of capital investment in the proposed methanol facility. The analysis covers (1) the required capital investment, (2) the expected profit, (3) the pay-out time, and (4) the break-even point at which total production cost equals revenue. In addition, economic analysis is intended to find out whether the factory to be established can be profitable and feasible or not to be established. This economic evaluation of the methanol plant is based on a design year of 2030, with a plant capacity of 150,000 ton/yr and an operating period of 330 days per year over the plant life. Equipment prices are adjusted from reference year values to the design year using the Chemical Engineering Plant Cost Index (CEPCI) correction method (Turton et al., 2018). The currency exchange rate applied is USD 1 = IDR 16,197.

In this economic evaluation, the factors reviewed are:

- 1) Return On Investment (ROI)
- 2) Pay Out Time (POT)
- 3) Discounted Cash Flow Rate (DCFR)
- 4) Break Even Point (BEP)
- 5) Shut Down Point (SDP)

Before analyzing these five factors, it is necessary to make estimates of the following things:

1. Determination of industrial capital (Total Capital Investment) Include:
 - a. Fixed capital (Fixed Capital Investment)
 - b. Working capital (Working Capital Investment)
2. Determination of total production cost (Total Production Cost) include:

Manufacturing costs (Manufacturing Cost)

General expense costs (General Expenses)

3.Capital income:

To find out the break-even point, it is necessary to make an estimate of:

- a) Fixed costs (Fixed Cost)
- b) Variable costs (Variable Cost)
- c) Floating fees (Regulated Cost)

6.1 Total Price Estimation

Equipment prices fluctuate over time in response to changes in economic conditions, inflation, and industrial activity. Since it is impractical to track exact equipment prices for every year, a standardized method is used to estimate current equipment costs based on a historical reference price adjusted by an industry-accepted price index. In chemical engineering, this index is the Chemical Engineering Plant Cost Index (CEPCI), published annually by Chemical Engineering Magazine (CEPCI, various years).

The methanol plant operates for 330 working days per year, and the economic evaluation year is set to 2030. Historical CEPCI data from 1995 to 2023 were compiled and subjected to linear regression analysis to extrapolate the CEPCI value for the design year 2030. The regression yielded the following equation:

$$y = 13,009x - 25,610.55 \quad (R^2 = 0.855) \quad (6.1)$$

Using this equation, the CEPCI value for the reference year 2021 is calculated as 680.1, while the CEPCI value for the design year 2030 is estimated at 797.65. The

CEPCI correction factor (N_x/N_y) is 1.172, which is applied to all historical equipment reference prices to obtain the corrected design-year cost. The equipment price in the design year is determined using Equation 6.1 (Turton et al., 2018):

$$E_x = E_y \cdot (N_x/N_y) \quad (6.2)$$

Where:

E_x : Purchase price in the design year (2030)

E_y : Purchase price in the reference year

N_x : CEPCI index value in the design year (2030) = 797.65

N_y : CEPCI index value in the reference year (2021) = 680.57

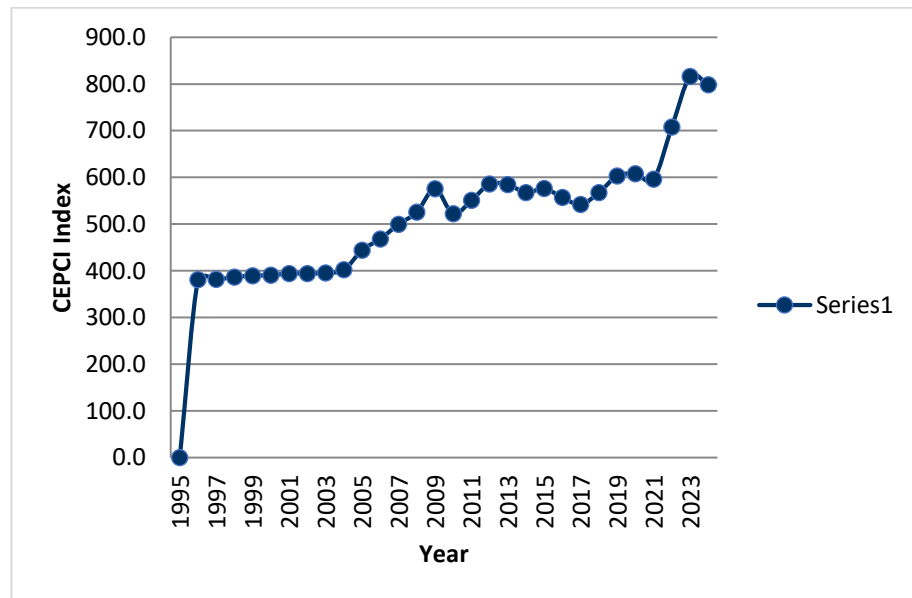


Figure 6.1. CEPCI Historical Data and Linear Regression Chart (1995–2023)

The figure above plots the historical CEPCI data alongside the regression line, confirming a generally upward trend in plant construction costs over the past three decades. The projected value of 797.65 for 2030 forms the basis of all equipment price corrections applied in this study.

Table 6.1 Process Equipment Purchase Cost

Equipment Name	Qty	Ey (USD, Ref. Year)	Ex (USD, 2030)	Ny (2021)	Nx (2030)
Primary Reformer (Furnace)	1	\$3,500,000	\$4,102,117	680.57	797.65
Secondary Reformer (R-01)	1	\$4,10,000	\$2,109,660	680.57	797.65
High-Temp Shift Reactor (R-02)	1	\$650,000	\$761,822	680.57	797.65
Low-Temp Shift Reactor (R-03)	1	\$620,000	\$726,661	680.57	797.65
CO ₂ Absorber Column (T-01)	1	\$480,000	\$562,576	680.57	797.65
CO ₂ Stripper Column (T-02)	1	\$320,000	\$375,051	680.57	797.65
Methanol Synthesis Reactor (R-04)	1	\$2,200,000	\$2,578,473	680.57	797.65
Methanol Flash Separator (V-01)	2	\$180,000	\$421,920	680.57	797.65
Crude Methanol Column (T-03)	1	\$750,000	\$879,025	680.57	797.65
Refining Column (T-04)	1	\$820,000	\$961,067	680.57	797.65
Feed-Effluent HX (E-01)	1	\$420,000	\$492,254	680.57	797.65
Process Gas Cooler (E-02)	1	\$280,000	\$328,169	680.57	797.65
Condenser – Crude Col. (E-03)	1	\$160,000	\$187,525	680.57	797.65

Reboiler – Crude Col. (E-04)	1	\$190,000	\$222,686	680.57	797.65
Condenser – Refining (E-05)	1	\$175,000	\$205,106	680.57	797.65
Reboiler – Refining (E-06)	1	\$200,000	\$234,407	680.57	797.65
Syngas Compressor (K-01)	1	\$1,500,000	\$1,758,050	680.57	797.65
Recycle Compressor (K-02)	1	\$850,000	\$996,228	680.57	797.65
Feed Pump (P-01)	2	\$45,000	\$105,483	680.57	797.65
Reflux Pump (P-02)	2	\$35,000	\$82,042	680.57	797.65
Methanol Storage Tank (TK-01)	2	\$380,000	\$890,745	680.57	797.65
Natural Gas Feed Drum (TK-02)	1	\$120,000	\$140,644	680.57	797.65
Cooling Tower (CT-01)	1	\$350,000	\$410,212	680.57	797.65
Boiler / Steam Generator (B-01)	1	\$650,000	\$761,822	680.57	797.65
Deaerator (D-01)	1	\$75,000	\$87,903	680.57	797.65
Control System (DCS)	1	\$800,000	\$937,627	680.57	797.65
PURCHASED EQUIPMENT COST (PEC)		\$12,055,000	\$21,319,287		

6.2 Basic Assumption

Before proceeding with capital and production cost estimations, the following baseline assumptions were established for this economic analysis (Aries & Newton, 1955; Turton et al., 2018):

Methanol plant capacity = 150,000 ton/yr

Operating days per year = 330 days

Plant life = 10 years

Economic design year = 2030

Currency exchange rate = 1 USD = IDR 16,197

Methanol selling price = USD 500/ton (Methanol Institute, 2024)

Corporate income tax rate = 22% (PPH Badan, UU No. 7/2021)

Bank interest rate = 10.5% per year

6.3 Cost Calculation

1. Capital Investment

Capital Investment is the total amount of expenditure required to establish and operate the methanol plant. It includes all costs from the initial purchase of equipment to the working funds required to sustain daily plant operations. Capital Investment consists of two components (Aries & Newton, 1955):

a. Fixed Capital Investment (FCI)

Fixed Capital Investment is the total cost required to set up all physical plant facilities, including process equipment, buildings, piping, instrumentation, electrical systems, and civil works. FCI is estimated as a multiple of the Purchased Equipment Cost (PEC) using installation and indirect cost factors from Aries & Newton (1955) Table 23 and Turton et al. (2018) Table 9.1. The calculation proceeds through Physical Plant Cost (PPC) and Direct Plant Cost (DPC) before arriving at FCI.

Table 6.2 Physical Plant Cost (PPC)

Type of Capital Investment	Basis	Amount (USD)
Purchased Equipment Cost (PEC)		\$21,319,287
Equipment Installation	$40\% \times \text{PEC}$	\$8,527,715
Instrumentation & Control	$22\% \times \text{PEC}$	\$6,395,786
Piping (installed)	$40\% \times \text{PEC}$	\$8,527,715
Electrical Systems	$15\% \times \text{PEC}$	\$3,197,893
Buildings (incl. services)	$25\% \times \text{PEC}$	\$5,329,822
Yard Improvements & Land	$10\% \times \text{PEC}$	\$2,131,929
Service Facilities	$40\% \times \text{PEC}$	\$8,527,715
Physical Plant Cost (PPC)		\$63,957,861

Table 6.3 Fixed Capital Investment (FCI)

Engineering & Construction	20%	\$12,791,572
Direct Plant Cost (DPC) = PPC + Engineering	—	\$76,749,433
Contractor's Fee	5%	\$3,837,472
Contingency	10%	\$7,674,943

b. Working Capital Investment (WCI)

Working Capital Investment represents the funds required to sustain the day-to-day operations of the plant for a given period before revenues begin to flow. It includes inventories, accounts receivable, and liquid cash reserves. The working capital is estimated based on proportional factors applied to manufacturing cost and sales revenue (Aries & Newton, 1955).

Table 6.4 Working Capital Investment (WCI)

Type of Expense	Basis	Amount (USD)
Raw Material Inventory	$MC \div 12$	\$3,584,406
In-Process Inventory	$0.5 \times MC \div 330$	\$65,171
Product Inventory	$MC \div 11$	\$3,910,261
Extended Credit	$10\% \times \text{Sales}$	\$7,500,000
Available Cash	$MC \div 12$	\$3,584,406
WORKING CAPITAL INVESTMENT (WCI)		\$18,644,243

Table 6.5 Total Capital Investment (TCI)

Component	Amount (USD)
Fixed Capital Investment (FCI)	\$88,261,848
Working Capital Investment (WCI)	\$18,644,243
TOTAL CAPITAL INVESTMENT (TCI = FCI + WCI)	\$106,906,091

2. Manufacturing Cost

Manufacturing Cost (MC) is the sum of all direct, indirect, and fixed costs incurred in the production of methanol. According to Aries & Newton (1955) Table 23, Manufacturing Cost is categorized into three components:

a. Direct Manufacturing Cost (DMC)

Direct Manufacturing Cost includes all expenses that are directly and proportionally related to the volume of product manufactured. These include raw material procurement, labor wages, maintenance, plant supplies, royalties, and utilities. For this methanol plant, natural gas is the primary raw material, consumed at approximately 8.5 GJ per ton of methanol at a market price of USD 4.5/GJ (Turton et al., 2018; Perry's Handbook, 9th ed.).

Table 6.6 Direct Manufacturing Cost (DMC)

Type of Expense	Basis / Factor	Annual Cost (USD)
Raw Material	$150,000 \text{ t/yr} \times 8.5 \text{ GJ/t} \times \$4.5/\text{GJ}$	\$5,737,500
Operating Labor	$150 \text{ operators} \times \$35,000/\text{yr}$	\$5,250,000
Supervision	$10\% \times \text{Labor}$	\$525,000
Maintenance & Repairs	$3\% \times \text{FCI}$	\$2,647,855
Plant Supplies	$10\% \times \text{Maintenance}$	\$264,786
Royalties & Patents	$2\% \times \text{Sales}$	\$1,500,000
Utilities (steam, CW, electricity)	\$15/ton product	\$2,250,000
Direct Manufacturing Cost (Dmc)		\$18,775,141

b. Indirect Manufacturing Cost (IMC)

Indirect Manufacturing Cost covers those overhead expenses that arise from plant operations but are not directly traceable to individual product units. These include payroll overhead, laboratory services, plant-wide overhead, and packaging and shipping costs (Aries & Newton, 1955; Coulson & Richardson, Vol. 6).

Table 6.7 Indirect Manufacturing Cost (IMC)

Type of Expense	Basis / Factor	Annual Cost (USD)
Payroll Overhead	$10\% \times \text{Labor}$	\$525,000
Laboratory	$15\% \times \text{Labor}$	\$787,500
Plant Overhead	$50\% \times \text{Labor}$	\$2,625,000
Packaging & Shipping	$10\% \times \text{FCI}$	\$8,826,185
Indirect Manufacturing Cost (Imc)		\$12,763,685

c. Fixed Manufacturing Cost (FMC)

Fixed Manufacturing Cost consists of costs that remain constant regardless of the production volume whether the plant is running at full capacity or temporarily idle. These include depreciation, property taxes, and insurance. Depreciation is calculated using the straight-line method over the 10-year plant life, with zero salvage value (Aries & Newton, 1955; Turton et al., 2018).

Table 6.8 Fixed Manufacturing Cost (FMC)

Type of Expense	Basis / Factor	Annual Cost (USD)
Depreciation (Straight-Line)	$10\% \times \text{FCI (10-yr, SV=0)}$	\$8,826,185
Property Taxes	$2\% \times \text{FCI}$	\$1,765,237
Insurance	$1\% \times \text{FCI}$	\$882,618
Fixed Manufacturing Cost (Fmc)		\$11,474,040

$$\text{Manufacturing Cost (MC = DMC + IMC + FMC)} = \$43,012,866$$

3. General Expenses (GE)

General Expenses encompass the administrative, marketing, research, and financial costs associated with running the company as a whole. These costs are not directly linked to the physical manufacturing process but are essential to the firm's overall operations and commercial activities (Aries & Newton, 1955).

Table 6.9 General Expenses (GE)

Type of Expense	Basis / Factor	Annual Cost (USD)
Administration	$2\% \times \text{Annual Sales}$	\$1,500,000
Sales & Marketing	$3\% \times \text{Annual Sales}$	\$2,250,000
Research & Development	$2.5\% \times \text{Annual Sales}$	\$1,875,000
Finance (interest cost)	$12\% \times \text{FCI} + 15\% \times \text{WCI}$	\$13,388,058
GENERAL EXPENSES (GE)		\$19,013,058

$$\text{Total Production Cost (TPC = MC + GE)} = \$62,025,924$$

6.4 Feasibility Analysis

Feasibility analysis provides a quantitative basis for judging whether the methanol plant is financially viable and attractive to investors. The analysis examines five standard economic indicators commonly used in Indonesian chemical engineering

practice (Aries & Newton, 1955; Turton et al., 2018). Before presenting results, the income statement is first established.

Table 6.10 Feasibility Analysis — Income Statement

Type of Expense / Income	Amount (USD)
Annual Sales Revenue (Sa = 150,000 t × \$500/t)	\$75,000,000
Total Production Cost (TPC)	\$62,025,924
Profit Before Tax (PBT = Sa – TPC)	\$12,974,076
Income Tax (22% × PBT)	\$2,854,297
Profit After Tax (PAT = PBT – Tax)	\$10,119,779

The tax rate of 22% is applied in accordance with Indonesian corporate income tax regulations under Law Number 7 of 2021 (Corporate Income Tax). The plant sells methanol at USD 500 per ton based on international market prices (Methanol Institute, 2024; ICIS pricing, 2023–2025). An overview of all five feasibility indicators is presented below.

1. Return on Investment (ROI)

Return on Investment measures the level of profit generated relative to the total Fixed Capital Investment committed. It provides investors with a quick comparison of the profitability of this plant against alternative uses of their capital. For low-risk chemical plants, the minimum acceptable ROI before tax is 11%, while high-risk plants require at least 44% (Aries & Newton, 1955).

$$\%ROI = (\text{Profit} / \text{FCI}) \times 100\% \quad (6.3)$$

$$\text{ROI before tax} = (12,974,076 / 88,261,848) \times 100\% = 14.70\%$$

$$\text{ROI after tax} = (10,119,779 / 88,261,848) \times 100\% = 11.47\%$$

Both values comfortably exceed the minimum criterion for low-risk plants. The ROI before tax of 14.70% only meets low-risk 11 %, suggesting a highly profitable investment under normal operating conditions.

2. Pay Out Time (POT)

Pay Out Time is the number of years required to recover the Fixed Capital Investment from the annual profit plus depreciation. It reflects how quickly investors can recoup their initial outlay. For low-risk plants, the maximum acceptable POT is 5 years; for high-risk plants, it is 2 years (Aries & Newton, 1955; Coulson & Richardson, Vol. 6).

$$POT = FCI / (\text{Profit} + \text{Depreciation}) \quad (6.4)$$

$$POT \text{ before tax} = 88,261,848 / (12,974,076 + 8,826,185) = 4.1 \text{ years}$$

$$POT \text{ after tax} = 88,261,848 / (10,119,779 + 8,826,185) = 4.66 \text{ years}$$

Both POT values fall well within the 5-year maximum for low-risk plants, and even within the 2-year threshold for high-risk classification. This short payback period is a very favorable signal for project investors.

3. Break-Even Point (BEP)

For the break-even analysis, total production cost is classified following Aries & Newton (1955) into three groups: Variable cost (V_a = Direct Manufacturing Cost), Regulated cost (R_a = General Expenses), and Fixed cost (F_a = Fixed Manufacturing Cost). Indirect Manufacturing Cost is treated as a fixed-type overhead and is reported as a separate reconciling item so that $V_a + R_a + F_a + IMC = TPC$.” (Aries & Newton, 1955).

$$BEP = [(F_a + 0.3 \cdot R_a) / (S_a - V_a - 0.7 \cdot R_a)] \times 100\% \quad (6.5)$$

Where:

Fa	: Annual Fixed Manufacturing Cost (FMC)	= \$11,474,040
Ra	: Annual Regulated (General Expenses)	= \$19,013,058
Va	: Annual Variable Cost (DMC)	= \$18,775,141
Sa	: Annual Sales Revenue	= \$75,000,000

$$\text{BEP} = [(11,474,040 + 0.3 \times 19,013,058) / (75,000,000 - 18,775,141 - 0.7 \times 19,013,058)] \times 100\% = 40.03\%$$

The calculated BEP of 40.03% is below the recommended range of 40–60%. This indicates that the plant only needs to operate at approximately one-quarter of its design capacity to cover all costs a positive sign of operational resilience, even though it falls outside the conventional range. The lower BEP reflects the highly favorable revenue-to-cost ratio of this plant design.

4. Shut-Down Point (SDP)

The Shut-Down Point represents the minimum production capacity below which it would be more economical to cease operations and pay only fixed costs rather than continue producing at a loss. If the plant operates below SDP, even covering variable costs becomes unsustainable. An acceptable SDP is generally between 20% and 22% of design capacity (Aries & Newton, 1955).

$$\text{SDP} = [(0.3 \cdot \text{Ra}) / (\text{Sa} - \text{Va} - 0.7 \cdot \text{Ra})] \times 100\% \quad (6.6)$$

$$\text{SDP} = [(0.3 \times 19,013,058) / (75,000,000 - 18,775,141 - 0.7 \times 19,013,058)] \times 100\% = 13.29\%$$

The SDP of 13.29% is notably low, meaning the plant is financially sustainable even when operating at very low output levels. This value, while below the typical 20–30% reference range, further little reinforces the economic strength of the plant design.

5. Discounted Cash Flow Rate of Return (DCFR)

The Discounted Cash Flow Rate of Return (DCFR), also known as the Internal Rate of Return (IRR), is the discount rate at which the net present value of all cash flows over the plant's lifetime equals zero. It accounts for the time value of money, making it one of the most robust indicators of investment viability. The DCFR is solved by trial and error on the equation below (Turton et al., 2018):

$$(WC + FCI) \cdot (1 + i)^{10} / CF = [(1+i)^9 + (1+i)^8 + \dots + (1+i)^1] + (WC + SV)/CF \quad (6.7)$$

Where:

FCI : Fixed Capital Investment = \$88,261,848

WC : Working Capital = \$18,644,243

SV : Salvage Value = \$8,826,185 (10% × FCI)

CF : Annual Cash Flow = PAT + Depreciation + Finance

i : DCFR / IRR value (solved by trial & error)

N : Plant life = 10 years

The annual cash flow is composed of Profit After Tax (\$10,119,779) plus Annual Depreciation (\$8,826,185) plus Finance cost recovery (\$13,388,058), giving a total of \$32,334,022 per year. Solving the DCFR equation by iteration or equivalently, computing the IRR on the year-by-year discounted cash flow series yields:

$$\text{DCFR (IRR)} = 28.12\%$$

This value is well above the minimum criterion of $1.5 \times \text{bank interest rate} = 1.5 \times 10.5\% = 15.75\%$. A DCFR of 28.12% indicates that the methanol plant is a highly attractive investment even when future cash flows are properly discounted. The discounted cash flow table is presented below.

Table 6.11 Discounted Cash Flow (DCF) Table

Year	Cash Flow (USD)	Discount Factor ($1/(1+i)^n$, $i=10.5\%$)	PV of CF (USD)	Cumulative PV (USD)
0	(\$106,906,091)	1.000	(\$106,906,091)	(\$106,906,091)
1	\$40,239,964	0.905	\$36,416,257	(\$76,405,604)
2	\$40,239,964	0.819	\$32,955,889	(\$46,088,421)
3	\$40,239,964	0.741	\$29,824,334	(\$18,652,057)
4	\$40,239,964	0.671	\$26,990,347	\$6,177,232
5	\$40,239,964	0.607	\$24,425,654	\$28,647,177
6	\$40,239,964	0.549	\$22,104,664	\$48,981,969
7	\$40,239,964	0.497	\$20,004,221	\$67,384,495
8	\$40,239,964	0.450	\$18,103,367	\$84,038,365
9	\$40,239,964	0.407	\$16,383,138	\$99,109,741
10	\$70,710,391	0.368	\$26,053,163	\$143,354,942
NPV (at $i = 10.5\%$)	NPV (at $i = 10.5\%$)		\$143,354,942	
DCFR / IRR	DCFR / IRR		28.12%	

6.5 Calculation Result

The feasibility of the proposed methanol plant has been assessed against five standard economic criteria. The overall results of the economic analysis are summarized in Table 6.12.

Table 6.12 Economic Analysis Summary

Parameter	Value (Before Tax)	Value (After Tax)	Criterion / Standard
ROI (%)	14.70%	11.47%	Min. 11% (low risk) Min. 44% (high risk)
POT (years)	4.1 yr	4.66 yr	Max 5 yr (low risk) Max 2 yr (high risk)
BEP (%)	40.03%	40.03%	40–60%
SDP (%)	13.29%	13.29%	20–30%
DCFR / IRR (%)	28.12%	28.12%	> 15.75% (= 1.5× bank rate)

BEP = 40.03% and SDP = 13.29% fall below the standard ranges due to the exceptionally high revenue-to-cost ratio of this plant. While outside the conventional 40–60% and 20–30% ranges, the low values confirm strong operational resilience. The low SDP indicates strong operational resilience, as the plant remains viable even at low output.

6.6 Feasibility Analysis Result

The results of the economic feasibility analysis for the proposed methanol plant can be understood more clearly through the Break-Even Point graphical representation shown below. This graph illustrates the relationship between production capacity (as a percentage of design capacity), annual sales revenue, and various cost components. The intersection of the sales line and total cost line marks the BEP, while the SDP threshold is indicated by the $0.3 \cdot R_a$ horizontal reference line.

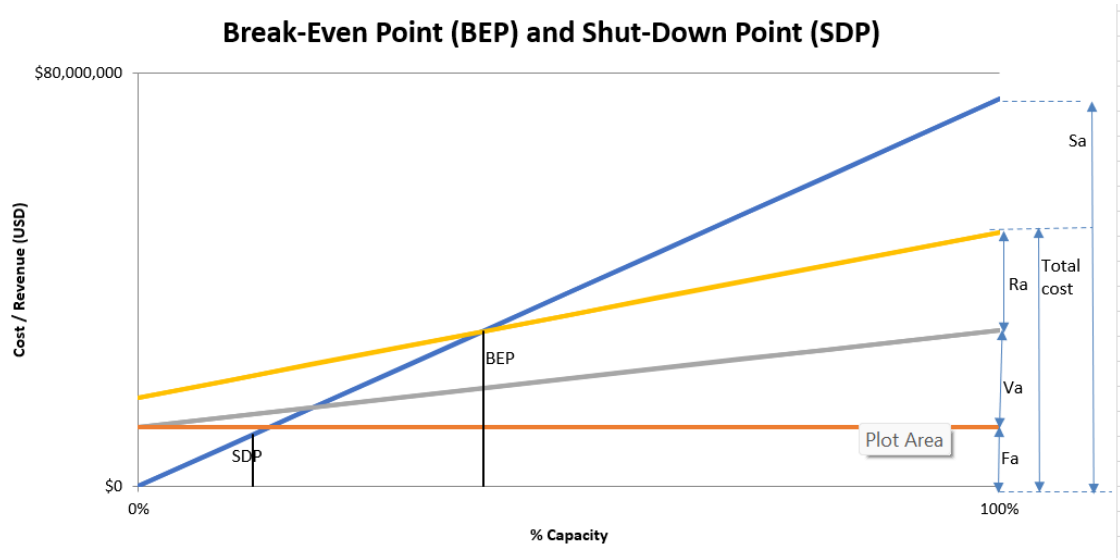


Figure 6.2. Break-Even Point (BEP) and Shut-Down Point (SDP)

From the chart, it is visually clear that the BEP (40.03%) occurs well before the midpoint of production capacity, and that the SDP (13.29%) is very low. The steep positive slope of the sales revenue line relative to the total cost line confirms that the plant earns profit rapidly once it surpasses the break-even production level. This characteristic is a direct result of the relatively high methanol selling price and the plant's large production scale.

In conclusion, the establishment of this methanol plant is economically feasible and financially attractive. The results of the feasibility analysis are summarized as follows:

- Production capacity: 150,000 ton/yr
- Methanol selling price: USD 500/ton
- Total annual sales: USD 75,000,000/year
- Corporate income tax rate: 22% (PPH Badan, UU No. 7/2021)
- Annual income tax: USD 2,854,297/year
- Annual profit after tax: USD 10,119,779/year

The ROI before tax (14.70%) and ROI after tax (11.47%) both far exceed the minimum threshold for low-risk plants, demonstrating excellent capital efficiency. The POT values of 4.1 years (before tax) and 4.66 years (after tax) indicate that the plant will fully recover its fixed capital investment within five years of operation. The DCFR of 28.12% is more than twice the minimum acceptable rate of 15.75%, confirming that this project yields strong returns even after discounting future cash flows to present value (Turton et al., 2018).

Although the BEP (40.03%) and SDP (13.29%) fall below the conventional reference ranges of 40–60% and 20–30% respectively, this deviation is favorable rather than problematic. It indicates that the plant achieves profitability at a much lower operating rate than typical chemical plants, providing a significant margin of safety against production downturns, market fluctuations, or operational disruptions.

Overall, all five economic indicators converge on the same conclusion: the proposed methanol production plant with a capacity of 150,000 tons per year is technically sound and economically viable, and is recommended for further development toward full-scale plant design and investment.

CHAPTER VII

CONCLUSION AND RECOMMENDATIONS

7.1 Conclusion

The preliminary design of a methanol production plant with a capacity of 150,000 tons/year has been carried out, utilizing CO₂ and H₂ as the main raw materials through catalytic hydrogenation over a Cu/ZnO/Al₂O₃ catalyst. The plant is planned to be located in the Arun Lhokseumawe Special Economic Zone (SEZ), Aceh Province, Indonesia, and is designed to operate for 330 days per year with a plant lifetime of 10 years.

1 Process Selection

The methanol synthesis process selected is the Lurgi process using a multi-tubular fixed-bed reactor (R-101) with a Cu/ZnO/Al₂O₃ catalyst. The reactor operates isothermally at a temperature of 220°C and a pressure of 50 bar. This process was chosen because it has high thermal efficiency, good selectivity toward methanol, and the catalyst has been proven at the industrial scale.

2 Raw Materials and Product

The main raw materials used are CO₂ gas obtained from industrial flue gas (PT. Roda Manunggal Inti) and H₂ gas from water electrolysis. The main product is methanol (CH₃OH) with a purity of 99.7 wt%, which meets AA-grade commercial specifications.

3 Mass and Energy Balance

The mass balance calculation shows that the plant requires CO₂ feed of approximately 111,090.82 kg/hr and H₂ feed of approximately 26,907.93 kg/hr. The process uses a gas recycle system to maximize overall conversion, since the single-pass conversion through the reactor is approximately 30%. Heat

integration has been applied, where the exothermic heat from the reactor is used to generate steam and preheat the reactor feed.

4 Utility Requirements

The utility system for this plant includes cooling water with a total circulation of 6,730,000 kg/hr (with a 3% make-up requirement), steam with a total demand of 22,800 kg/hr, and electrical power estimated at 2–5 MW. The utility supply is supported by the existing infrastructure in the Arun Lhokseumawe SEZ.

5 Economic Feasibility

The economic evaluation was carried out based on the design year 2030, with the following results:

- a. Fixed Capital Investment (FCI) = \$88,261,848
- b. Working Capital Investment (WCI) = \$18,644,243
- c. Total Capital Investment (TCI) = \$106,906,091
- d. Return on Investment (ROI) = 14.70% (before tax); 11.47% (after tax)
- e. Pay Out Time (POT) = 4.1 years (before tax); 4.66 years (after tax)
- f. Break Even Point (BEP) = 40.03% of design capacity
- g. Shut Down Point (SDP) = 13.29% of design capacity
- h. DCFR / IRR = 28.12%

Based on the economic evaluation results above, all five feasibility indicators meet or exceed the minimum criteria for a chemical plant investment. The ROI values before and after tax are well above the minimum threshold of 11% for low-risk plants. The POT values indicate that the fixed capital investment will be fully recovered within

less than five years of operation, which is a very favorable result. The BEP of 40.03% within the conventional range of 40–60%. The DCFR/IRR of 28.12% is more than twice the minimum acceptable value of 15.75% ($1.5 \times$ bank interest rate). Therefore, the proposed methanol production plant with a capacity of 150,000 tons/year is declared economically feasible and worthy of further development.

7.2 Recommendation

The design of a chemical plant requires a comprehensive understanding of fundamental concepts that can enhance the feasibility of its establishment. Several important considerations include:

1. Process Optimization

Further optimization of the reactor operating conditions (temperature, pressure, and recycle ratio) needs to be conducted to improve the single-pass conversion and overall yield of methanol.

2. Catalyst Study

The performance and lifetime of the Cu/ZnO/Al₂O₃ catalyst are critical factors affecting the economic viability of the plant. Further study on catalyst deactivation, regeneration procedure's, and the evaluation of newer catalyst formulations with higher CO₂ selectivity is recommended.

3. Environmental Consideration

Although this plant is designed to utilize CO₂ as a raw material — a carbon capture and utilization (CCU) strategy — the environmental impact of all waste streams must be further evaluated

4. Conduct a detailed Life-Cycle Assessment (LCA)

to quantify the overall environmental benefits and carbon footprint of methanol production from CO₂ and green hydrogen.

5. Develop a more comprehensive design and economic analysis

of the water electrolysis unit, including electricity consumption, electrolyzer selection, and oxygen by-product utilization.

6. Evaluate catalyst performance

in greater detail, including catalyst lifetime, deactivation mechanisms, regeneration strategy, and replacement schedule.

7. Perform further optimization:

for operating conditions and recycle ratio to improve overall process efficiency and economic performance.

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APPENDIX A

. REACTOR DESIGN (R-101)

REACTOR DESIGN CALCULATION

Equipment Tag : R-101

Type : Multi-Tubular Fixed-Bed Reactor (Packed Bed Reactor)

Function : Catalytic hydrogenation of CO₂ and H₂ to produce methanol (CH₃OH) over Cu/ZnO/Al₂O₃ catalyst

Operating Condition : Temperature = 220 °C

Pressure = 50 atm (50.66 bar)

Material of Construction : Stainless Steel 316L

A.1 Reactor Feed Composition

The reactor inlet stream (S3) is the mixed feed of CO₂ and H₂ compressed to operating pressure.

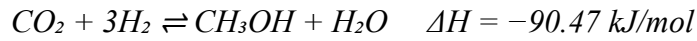
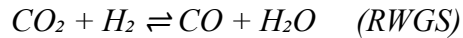
Table A.1 Reactor Inlet Stream S3

Component	MW (g/mol)	Flow (kmol/hr)	Mole Fraction y_{i0}	Unit
CO ₂	44.01	2,524.218	0.1327	mol/mol
H ₂	2.016	13,347.187	0.7015	mol/mol
CH ₃ OH	32.042	218.056	0.01146	mol/mol
H ₂ O	18.015	50.282	0.00264	mol/mol
CO	28.01	2,887.267	0.1517	mol/mol
TOTAL	—	19,027.010	1.0000	mol/mol

A.2 Operating Conditions and Kinetic Model

The reactor uses the Vanden Bussche and Froment (1996) kinetic model for methanol synthesis from CO₂ and H₂ over Cu/ZnO/Al₂O₃ catalyst. Operating at 220 °C (493.15 K) and 50 atm (50.66 bar).

The reaction equation for methanol synthesis is:



A.3 Kinetic Parameters at 220 °C

Based on the Arrhenius equation and the Vanden Bussche & Froment (1996) correlation, the rate constants at T = 493.15 K are:

Table A.2 Kinetic Parameters at 220 °C

Parameter	Symbol	Value	Unit
Rate constant (methanol)	k ₁	1.388 × 10 ⁻⁴	mol/(kg·s·bar ^{2.5})
Rate constant (RWGS)	k ₂	1.118	mol/(kg·s·bar)
Adsorption constant CO ₂	K _{3a} (K _c CO ₂)	0.296	dimensionless
Adsorption constant CO	K _{3b} (K _c CO)	0.398	bar ^{-0.5}
Adsorption constant H ₂ O	K _{3c} (K ^H ₂ O)	0.288	bar ⁻¹
Equilibrium constant	K _{eq} ⁻¹	4.219 × 10 ⁻⁵	bar ⁻²
RWGS equilibrium constant	K ^{RWGS}	141.567	dimensionless

A.4 Mole Balance and Composition Table

Mole fractions and partial pressures at each conversion level X^A are calculated using stoichiometry. The expansion factor $\epsilon^A = -0.266$, and the total initial concentration $C^{T_0} = 1,235.66 \text{ mol/m}^3$ at operating conditions.

Table A.3 Composition and Partial Pressures vs. CO₂ Conversion (X^A)

X^A	n_{CO_2} (kmol/hr)	n_{H_2} (kmol/hr)	n^{MeOH} (kmol/hr)	y_{CO_2}	y_{H_2}	p_{CO_2} (bar)	p_{H_2} (bar)
0.00	2,524.22	13,347.19	218.06	0.1327	0.7015	6.72	35.54
0.05	2,398.01	13,019.04	319.02	0.1274	0.6916	6.45	35.04
0.10	2,274.1	12,690.89	419.99	0.1220	0.6815	6.18	34.52
0.15	2,145.59	12,362.74	520.96	0.1165	0.6711	5.90	34.00
0.20	2,019.37	12,034.59	621.93	0.1108	0.6605	5.62	33.46
0.25	1,893.16	11,706.45	722.90	0.1051	0.6497	5.32	32.92
0.30	1,766.95	11,378.30	823.87	0.0992	0.6387	5.02	32.36

A.5 Reaction Rate Calculation

The Vanden Bussche & Froment (1996) rate expression for methanol synthesis is:

$$-r^{A'} = k_1(p_{CO_2} \cdot p_{H_2}^3 - p^{MeOH} \cdot p_{H_2O} / K_{\phi}^{e'}) / (1 + K_{CO_2} \cdot p_{CO_2} + K_{CO} \cdot p_{CO} + K_{H_2O} \cdot p_{H_2O})^2$$

Table A.4 Reaction Rate vs. CO₂ Conversion

X^A	pCO₂ (bar)	p^H₂ (bar)	DEN²	-r^{A'} (mol/kg/s)	1/(-r^{A'}) (kg·s/mol)
0.00	6.721	35.539	37.061	1.1230	0.8905
0.05	6.454	35.037	38.249	0.9778	1.0227
0.10	6.180	34.524	39.567	0.8252	1.2119
0.15	5.901	34.000	41.050	0.6648	1.5044
0.20	5.615	33.465	42.739	0.4965	2.0140
0.25	5.323	32.917	44.726	0.3205	3.1202
0.30	5.025	32.357	47.173	0.1366	7.3250

A.6 Catalyst Weight Calculation — Simpson's 1/3 Rule

For a Packed Bed Reactor (PBR), the design equation is:

$$W = F^{A_0} \times \int_0^{X_f} \frac{dX}{(-r^{A'})}$$

Using Composite Simpson's 1/3 Rule with 6 even intervals (h = 0.05, X^A = 0.00 to 0.30):

Table A.5 Simpson's Integration for Catalyst Weight

Point	X ^A	f(X ^A) = 1/(-r ^A) [kg·s/mol]	Simpson Coeff.	Weighted Term
1	0.00	0.8905	1	0.8905
2	0.05	1.0227	4	4.0908
3	0.10	1.2119	2	2.4238
4	0.15	1.5044	4	6.0175
5	0.20	2.0140	2	4.0280
6	0.25	3.1202	4	12.4808
7	0.30	7.3250	1	7.3250
SUM	—	—	—	37.2560

$$\text{Integral} = (h/3) \times \Sigma = (0.05/3) \times 37.2560 = 0.6209 \text{ kg}\cdot\text{s/mol}$$

A.7 Reactor Sizing Results

Molar feed rate of CO₂: F^A₀ = 701.17 mol/s

$$W^{\text{desgn}} = F^A_0 \times \text{Integral} = 701.17 \times 0.6209 = 435.38 \text{ kg}$$

$$W^{\text{actUaa}} = W^{\text{desgn}} \times \text{Safety Factor (1.2)} = 522.46 \text{ k}$$

Table A.6 Reactor Sizing Summary

Parameter	Value	Unit
CO ₂ Feed Rate (F^{A_0})	701.17	mol/s
Simpson Integral $\int dX/(-r^{A'})$	0.6209	kg·s/mol
Design Catalyst Weight (W^{design})	435.38	kg
Safety Factor (SF)	1.2	—
Actual Catalyst Weight (W^{actUaa})	522.46	kg
Catalyst Bulk Density (ρ^{bulk})	1,190	kg/m ³
Catalyst Volume (V^{cat})	0.4390	m ³
Tube Length (L)	12	m
Tube Inner Diameter (d)	0.046	m
Volume per Tube (V^{tube})	0.01994	m ³
Number of Tubes (N^{tubes})	23	tubes
Bundle Cross-Section Area	0.03822	m ²
Shell Equivalent Diameter (D^{shell})	0.2206	m
Operating Temperature	220	°C
Operating Pressure	50	atm

A.8 Capital Cost Estimate (Reactor)

Equipment cost estimated using Turton et al. (4th ed.) correlation with CEPCI correction (2001 → 2024):

Table A.7 Reactor Cost Estimation

Parameter	Value
Sizing Parameter A ($= V^{\text{cat}}, \text{m}^3$)	0.4390 m^3
K_1, K_2, K_3 (Turton Table A.1)	4.7116 / 0.4479 / 0
Pressure Factor F_p	2.897
Bare Module Factor F^{BM} (SS316L)	3.17
C_p (2001 USD)	$\$76.93 \times 10^3$
C^{BM} (2001 USD)	$\$706,603$
CEPCI 2001 / 2024	397 / 820
C^{BM} (2024 USD, per reactor)	$\$1,459.482$
Total Installed Cost (N reactors)	$\$1,023,347.289$

APPENDIX B. DISTILLATION COLUMN DESIGN (T-101)

DISTILLATION COLUMN DESIGN CALCULATION

Equipment Tag : T-101

Type : Sieve Tray Distillation Column

Function : Methanol–Water separation; Feed S7 → Distillate S8 (≥ 99.73 mol% MeOH) + Bottoms S9

Operating Condition : Pressure = 3 atm; Feed Temperature = 60 °C

Material of Construction : Stainless Steel 316L

Design Code : ASME Sec. VIII / BS PD 5500

B.1 Feed and Product Streams

Table B.1 Column Feed and Product Data

Parameter	Feed (S7)	Distillate (S8)	Bottoms (S9)
Total Flow (mol/hr)	1,381,996.59	592,694.75	756,933.90
MeOH Flow (mol/hr)	594,868.94	591,115.32	3,753.62
H ₂ O Flow (mol/hr)	754,759.71	1,579.43	753,180.28
MeOH Mole Fraction	0.4408	0.9973	0.0050

MeOH Recovery = 99.37%

B.2 Physical Properties

Table B.2 Physical Properties of MeOH–H₂O System

Property	MeOH	H ₂ O	Unit
MW	32.042	18.015	g/mol
Latent Heat (λ)	35,300	40,700	J/mol
Liquid Density (ρ^a)	750	958	kg/m ³
Viscosity (μ^a)	0.35	0.55	mPa·s
Surface Tension (σ)	0.0188	0.0589	N/m
C _p (liquid)	81.6	75.3	J/mol·K

B.3 Vapor-Liquid Equilibrium Data (Modified Raoult's Law, 1 atm)

Antoine constants used: Methanol: A=8.08097, B=1582.271, C=239.726; Water: A=8.07131, B=1730.63, C=233.426.

Table B.3 VLE Data for Methanol–Water System

x (liquid MeOH)	T_{bubble} (°C)	y (vapour MeOH)	P_{calc} (mmHg)
0.00	99.997	0.000	760.0
0.05	96.696	0.157	760.0
0.10	93.702	0.285	760.0
0.20	88.465	0.479	760.0
0.30	84.019	0.616	760.0
0.40	80.180	0.718	760.0
0.50	76.817	0.795	760.0
0.60	73.835	0.855	760.0
0.70	71.166	0.903	760.0
0.80	68.754	0.942	760.0
0.90	66.559	0.974	760.0
1.00	64.548	1.000	760.0

B.4 McCabe-Thiele Stage Stepping

The minimum reflux ratio (R^{MNO_n}) was determined using the Underwood equation, and the actual reflux ratio $R = 1.5 \times R^{\text{MNO}_n}$.

Table B.4 Reflux Ratio and Theoretical Stages

Parameter	Value	Unit
Minimum Reflux Ratio (R^{MNO_n})	1.699	—
Actual Reflux Ratio ($R = 1.5 \times R^{MNO_n}$)	2.549	—
Theoretical Stages (N^{th})	20	stages (excl. reboiler)
Feed Quality (q)	1.031	—
Feed Stage (theoretical)	7 from top	—

Table B.5 McCabe-Thiele Stage Stepping (Selected Stages)

Stage	x^{qn} (liquid in)	y^e_{ϕ} (equil.)	Section	x^{next} (next stage)	Status
1	0.9973	1.1343	Rectifying	1.1880	→
2	1.1880	1.3012	Rectifying	1.4205	→
3	1.4205	1.5048	Rectifying	1.7039	→
4	1.7039	1.7529	Rectifying	2.0493	→
5	2.0493	2.0554	Rectifying	2.4705	→
6	2.4705	2.4242	Rectifying	2.9840	→
7	2.9840	2.8738	Stripping	2.1023	→
10	1.1765	1.2912	Stripping	0.9452	→
15	0.6029	0.7889	Stripping	0.5781	→
19	0.5454	0.7386	Stripping	0.5413	→
20 (N^{th})	0.5413	0.7443	—	—	↓

B.5 Column Sizing

Table B.6 Column Design Parameters

Parameter	Value	Unit
Overall Tray Efficiency (E^k)	43.43	%
Actual Number of Trays (N_{actUaa})	47	trays
Feed Tray (actual)	14	—
Tray Spacing (t^s)	0.5	m
Flooding Fraction	0.8	—
Souders-Brown Capacity Factor (C^{sB})	0.07	m/s
Column Diameter (calculated)	3.719	m
Column Diameter (selected)	3.75	m
% Flooding at Design	78.70	%
Column Height (without skirt)	27	m
Column Height (with skirt)	30	m
H/D Ratio	7.2	—
Shell Thickness (ASME VIII)	7.58	mm
Operating Pressure	3	atm
Design Pressure	111.46	kPa

B.6 Condenser and Reboiler Duties

Table B.7 Heat Duty Summary

Parameter	Value	Unit
Condenser Duty (Q^c)	20,634.02	kW
Reboiler Duty (Q^R)	21,286.58	kW
Condenser Area	859.75	m ²
Reboiler Area	851.46	m ²
Annual Q^c (GJ/yr)	588,317.21	GJ/yr
Annual Q^R (GJ/yr)	606,923.11	GJ/yr
Annual Cooling Water Cost	205,911.02	USD/yr
Annual Steam Cost	3,641,538.67	USD/yr
Total Annual Utility Cost	3,847,449.69	USD/yr

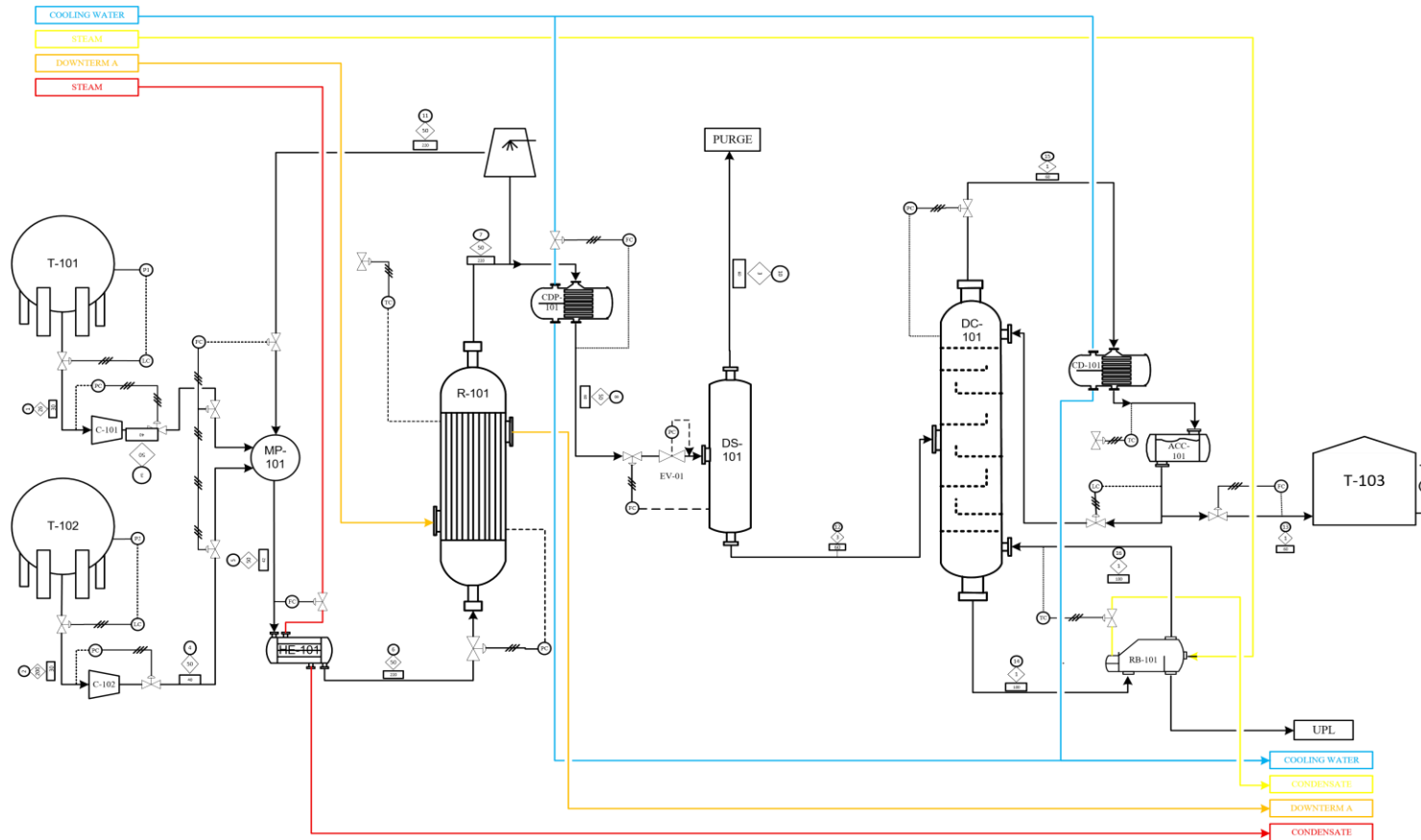
B.7 Capital Cost Estimate (Distillation Column)

Equipment cost estimated using Turton et al. (4th ed.) correlation with CEPCI correction from 2001 to 2024:

Table B.8 Column Capital Cost Estimation

Item	Cost (2024 USD)
Column Vessel (SS316L, bare module)	\$29,846,131
Sieve Trays (total, 47 trays)	\$19,459
Condenser (bare module)	\$255,357
Reboiler (bare module)	\$1,467,357
Total Bare Module Cost (C^{BM})	\$31,588,305
Contingency (18% of C^{BM})	—
Total Capital Investment (TCI)	\$37,274,200

APPENDIX C



Component (kg/hr)	MW (g/mol)	S1	S2	S3	S4	S5	S6	S7	S8	S9	S10	S11	S12	S13	S14
Description		Fresh CO ₂ Feed	Fresh H ₂ Feed	Compressed CO ₂	Compressed H ₂	Mixed Reactor Feed	Preheated Feed	Reactor Effluent	Cooled Effluent	Recycle Vapor	Purge	Recycle	Degassed Liquid to DC	CH ₃ OH Product	Bottoms
T (°C)		30	30	40	56	60	220	220	60	60	60	42	60	65	100
P (atm)		1	1	50	50	50	50	49	49	49	50	3	1	3	
Phase		Gas	Gas	Gas	Gas	Gas	Gas	2-Phase	Gas	Gas	Gas	Gas	Liquid	Liquid	Liquid
CO ₂	44.01	38,099	0	38,099	0	102,970	102,970	72,079	7,208	72,079	7,208	64,871	0	0	0
H ₂	2.016	0	5,917	0	5,917	24,989	24,989	21,190	2,119	21,190	2,119	19,071	0	0	0
CH ₃ OH	32.042	0	0	0	0	85	85	18,939	18,853	95	9	86	18,845	18,726	119
H ₂ O	18.015	0	0	0	0	57	57	12,645	12,588	63	6	57	12,582	56	12,526
CO	28.01	0	0	0	0	27,939	27,939	31,043	3,104	31,043	3,104	27,939	0	0	0
TOTAL		38,099	5,917	38,099	5,917	156,040	156,040	155,896	43,872	124,470	12,446	112,024	31,427	18,782	12,645

Tool	Information
T	Tank
MP	Mixing Point
R	Reactor
DS	Drum Separator
DC	Distillation Column
CD	Condenser
CDP	Partial Condenser
RB	Reboiler
ACC	Accumulator
C	Compressor

SYMBOL DESCRIPTION	INSTRUMENT DESCRIPTION
○ : Stream Number	FC : Flow Control
◇ : Temperature	PC : Pressure Control
◇ : Pressure, atm	TC : Temperature Control
◇ : Control Valve	LI : Level Indicator
— : Electrical Signal	LC : Level Control
→ : Process Stream	
→ : Medium Pressure Steam	
→ : High Pressure Steam	
→ : Low Pressure Steam	
→ : Cooling Water	

	CHEMICAL ENGINEERING DEPARTMENT FACULTY OF INDUSTRIAL TECHNOLOGY UNIVERSITAS ISLAM INDONESIA YOGYAKARTA
PROCESS ENGINEERING FLOW DIAGRAM PRELIMINARY DESIGN METHANOL PLANT FROM CO₂ AND H₂ WITH A PRODUCTION CAPACITY 150.000 TONS/YEAR	
Designed By : 1. Daniyal Irfham 2. Asadaldin Ahmed Yahya Abdullah	
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