TECHNO-ECONOMIC STUDY OF CO_2 UTILIZATION FROM TAIL GAS STEAM REFORMING UNIT INTO METHANOL AT OIL REFINERIES IN INDONESIA

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ABSTRACT

The downstream oil and gas sector contributes significantly to CO_2 emissions, particularly from steam methane reforming (SMR), which produces 4.33 tCO_2e/kbbl of crude oil processed. This study investigates the technical and economic feasibility of utilizing CO_2 from SMR tail gas to produce methanol. The process simulation was conducted using Aspen HYSYS v12, modeling the methanol synthesis process through the hydrogenation of CO_2 with hydrogen (H_2) sourced from the Naphtha Reforming Unit. Results from the simulation indicate an optimal methanol yield of 81.16% at 235°C and 50 bar, producing 70 tons/hour (612 KTA) of methanol with a purity of 99.85% wt. The economic analysis reveals an IRR of 8.04%, an NPV of -26.5 million USD, and a pay-out time (POT) of 22 years, showing that the project is not yet economically feasible under current conditions. Sensitivity analysis highlights that methanol selling price and production capacity are the most critical factors impacting the project's viability, where a 20% increase in methanol price or production capacity could significantly improve IRR above the hurdle rate of 10.83%. This study provides insight into the potential of CO_2 utilization for methanol production, contributing to emission reduction and product diversification in refineries. Further optimization of feedstock costs and integration with downstream products like MTBE could enhance economic feasibility.

Keywords: *CO_2 Utilization, Methanol Synthesis, Tail Gas, Steam Methane Reforming, Techno-Economic Analysis, Sensitivity Analysis*

1. INTRODUCTION

1.1 Background

Energy transition in refineries focuses on reducing CO_2 emissions through strategies such as product diversification into petrochemicals, substituting low-carbon feedstocks, and adopting technologies like electrification, green hydrogen, and Carbon Capture and Utilization (CCU). According to Wood Mackenzie (2023), refinery decarbonization can be achieved through seven key initiatives, including substituting crude oil with low-carbon alternatives, optimizing Fluidized Catalytic Cracking (FCC) units, and utilizing hydrogen produced through Steam Methane Reforming (SMR).

The Steam Methane Reforming (SMR) process, a primary method for hydrogen production, emits 4.33 tCO_2e/kbbl of crude oil processed, making it one of the largest

CO_2 emission sources in refineries. Unlike combustion processes, CO_2 emissions from SMR tail gas have a higher concentration of CO_2 due to the absence of nitrogen (N2) from air combustion, presenting an ideal feedstock for methanol production. Methanol, a vital petrochemical product, can serve as a building block for formaldehyde, fuels, and Methyl Tertiary Butyl Ether (MTBE), while simultaneously offering a pathway to reduce CO_2 emissions.

1.2 Research Gaps

Current research on CCU primarily focuses on CO_2 derived from flue gas, resulting from combustion processes. Several studies have explored the conversion of CO_2 into acetic acid, formic acid (Medrano and Garcia, 2018), synthetic fuels (Panahi et al., 2018), methanol (Tagdisian et al., 2015), and dimethyl ether (Mevawala et al., 2019). However, limited research targets the utilization of CO_2 from SMR tail gas, which contains both CO_2 and residual hydrogen (H_2).

Nugroho (2022) investigated methanol synthesis using CO_2 from flue gas and hydrogen from Naphtha Reforming but did not consider tail gas as a feedstock. Methanol synthesis using SMR tail gas offers greater efficiency due to its higher CO_2 concentration and the presence of H2, which can improve the overall conversion rate. Furthermore, integrating this process with the Naphtha Reforming Unit's hydrogen stream eliminates the need for additional hydrogen production units, enhancing energy and operational efficiency.

1.3 Objectives

This study aims to evaluate the technical performance of methanol synthesis using CO_2 from SMR tail gas and H_2 from Naphtha Reforming. Assess the economic feasibility of methanol production in oil refineries, focusing on key parameters such as Net Present Value (NPV) and Internal Rate of Return (IRR) and conduct a sensitivity analysis to examine the impact of methanol price, production rates, and feedstock costs on project viability.

2. LITERATURE REVIEW

2.1 Overview of Oil Refining Processes

Oil refining is an essential process for converting crude oil into valuable end products such as gasoline, kerosene, diesel, and petrochemical feedstocks. Modern refineries incorporate advanced technologies to meet the growing demand for clean energy and petrochemical diversification, driven by environmental regulations and global energy transition trends (Speight, 2020). The complexity of refinery configurations depends on the desired products. Among the most versatile configurations are those with catalytic cracking and coking units, enabling higher conversion of heavy residues into lighter, value-added products.

Auxiliary units such as Steam Methane Reforming Units (SMR) play a critical role in providing hydrogen for hydrotreating, hydrocracking, and other catalytic processes.

These units also produce tail gas, a byproduct rich in CO_2 and hydrogen, which can be utilized for methanol synthesis.

2.2 Steam Methane Reforming (SMR) and CO_2 Emissions

Steam Methane Reforming (SMR) is the most widely used process for hydrogen production in refineries. Natural gas, refinery gas, or light hydrocarbons are reacted with steam at high temperatures (700-1000°C) over a nickel-based catalyst to produce syngas (H_2, CO, and CO_2). The main reactions are as follows:

- Steam Reforming: $CH_4 + H_2O \Leftrightarrow CO + 3\,H_2$ ΔH = +206 kJ/mol
- Water Gas Shift Reaction: $CO + H_2O \Leftrightarrow CO_2 + H_2$ ΔH = -41 kJ/mol

Hydrogen produced through SMR is essential for hydroprocessing and other catalytic operations. However, SMR emits significant CO_2, both as a reaction byproduct and from fuel combustion for heating. Stoichiometrically, 1 kg of hydrogen production generates approximately 5.5 kg of CO_2 (Jenkins and Mullinger, 2023).

CO_2 emissions from SMR tail gas are of particular interest due to their higher concentration compared to flue gases. Unlike combustion flue gases, tail gas contains minimal nitrogen (N_2), making CO_2 recovery and utilization more efficient. Pressure Swing Adsorber (PSA) systems, commonly used for hydrogen purification, produce tail gas with approximately 46.6% CO_2, 11.8% CO, and 24.9% H_2.

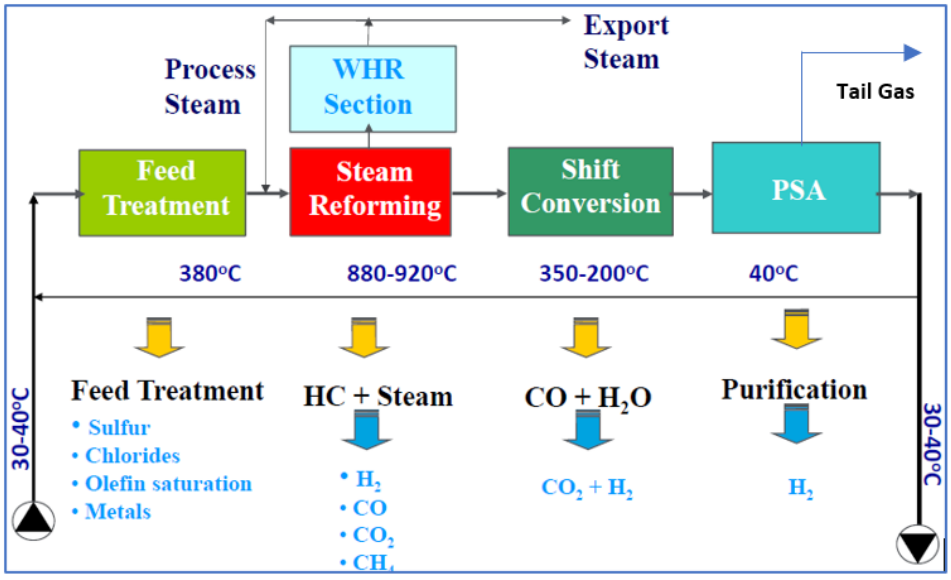


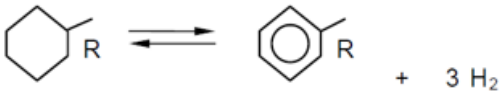
Figure 1: Blok Flow Diagram Steam Reforming Process

2.3 Naphtha Reforming and Hydrogen Production

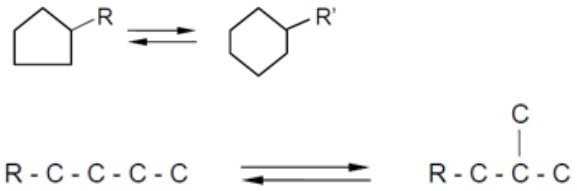
Catalytic Naphtha Reforming is a critical process for producing high-octane gasoline (reformate) and hydrogen as a valuable byproduct. The process involves dehydrogenation of naphthenes into aromatics and isomerization of paraffins over platinum-based catalysts at 480-550°C.

The key reactions include:

- **Dehydrogenation naphtene:**



- **Isomerization naphtene and paraffin :**



In addition to this reaction, other reactions occurred in naphtha reforming, such as dehydrocyclization of paraffins, hydrocracking, and demethylation. This process also produced hydrogen, C3 and C4 (LPG fraction) as side products.

Hydrogen yield is inversely proportional to reformate production, with typical hydrogen recovery ranging from 1-4% by weight of the feedstock (Kaiser et al., 2020). The hydrogen purity obtained from Naphtha Reforming Units (NRU) exceeds 99.9%, making it suitable for CO_2 hydrogenation processes.

Integrating hydrogen from NRU with CO_2 from SMR tail gas for methanol synthesis can optimize refinery operations, reduce emissions, and diversify product portfolios.

2.4 Methanol Synthesis from CO_2

Methanol is a key petrochemical intermediate used in the production of formaldehyde, olefins, and fuels. Commercial methanol synthesis typically involves syngas (CO and H_2) conversion over copper-based catalysts at 200-300°C and 50-100 bar (Goeppert, et al,. ,2014). The main reactions are:

- Hydrogenation of CO: $CO + 2H_2 \Leftrightarrow CH_3OH$ ΔH = -21.7 kcal/mol
- Hydrogenation of CO_2: $CO_2 + 3H_2 \Leftrightarrow CH_3OH + H_2O$ ΔH = -11.9 kcal/mol
- Reverse Water Gas Shift Reaction: $CO_2 + H_2 \Leftrightarrow CO + H_2O$ ΔH = 9.8 kcal/mol

CO_2 hydrogenation to methanol is thermodynamically favored at low temperatures and high pressures. However, CO_2's high stability and inert nature require optimized conditions and selective catalysts to achieve high conversion and yield.

2.5 Catalyst for Methanol Synthesis

Copper-based catalysts ($Cu/ZnO/Al_2O_3$) are widely used for methanol synthesis due to their high activity and selectivity. The role of each component is as follows:

- Cu: Provides active sites for hydrogenation reactions.
- ZnO: Enhances dispersion of Cu and stabilizes the catalyst.
- Al2O3: Acts as a structural support, improving catalyst durability.

The reaction mechanism involves bifunctional adsorption of CO_2 on ZnO and H_2 dissociation on Cu, leading to methanol formation (Fisher and Bell, 1998). Kinetic studies by Graaf et al. (1986) and Lim et al. (2009) have provided equilibrium constants and rate expressions to model the methanol synthesis process accurately.

2.6 State-of-the-Art Methanol Synthesis Technology

Kiss (2016) proposed an advanced methanol synthesis scheme integrating a stripper unit to remove water from hydrogen feed streams and unconverted CO_2 from the methanol-water product mixture. This process is particularly effective when using wet hydrogen streams, such as those from Naphtha Reforming Units, improving conversion efficiency and product purity.

Several studies have optimized methanol synthesis using simulation tools such as Aspen HYSYS and Aspen Plus. Rosha et al. (2021) demonstrated process optimization using biogas-derived CO_2, achieving stoichiometric conversion efficiencies. Francis et al. (2022) used response surface methodology to optimize temperature, pressure, and feed rates, highlighting the potential for reducing thermodynamic limitations in CO_2 hydrogenation.

2.7 Economic Evaluation of Methanol Production

Economic feasibility analysis of methanol plants involves key parameters such as Net Present Value (NPV), Internal Rate of Return (IRR), and Pay-Out Time (POT). According to Mellichamp (2019), the criteria for economic viability are:

- NPV: Positive value indicates profitability.
- IRR: Must exceed the hurdle rate (>10.83%).
- POT: Shorter payback periods are preferred.

Sensitivity analysis is critical to evaluate the impact of methanol price, production rates, and feedstock costs on project viability. Kiss (2016) highlighted the importance of process integration and cost optimization to improve economic performance.

The integration of CO_2 from SMR tail gas and hydrogen from NRU for methanol synthesis presents a promising pathway for emission reduction and product diversification in refineries. The use of advanced catalytic processes and energy-efficient schemes, such as those proposed by Kiss (2016), offers significant potential for improving conversion efficiency and economic viability. This study builds upon existing research by simulating methanol synthesis under refinery-specific conditions and evaluating its techno-economic feasibility.

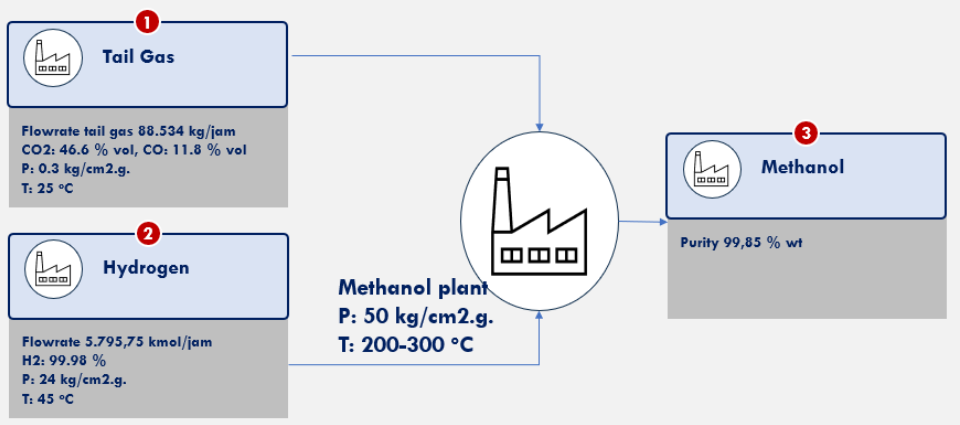


Figure 2: Methanol Plant from CO_2

3. METHODOLOGY

The research methodology flowchart for the study on CO_2 utilization from SMR tail gas to produce methanol in an Indonesian refinery is illustrated in Figure 3.

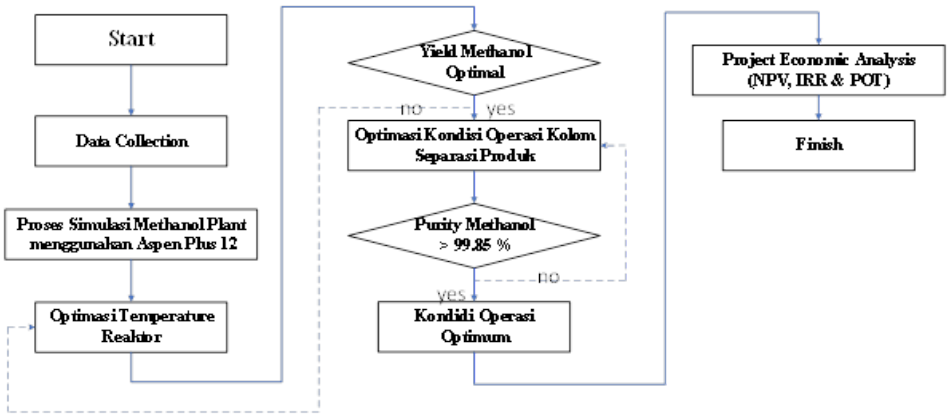


Figure 3 Flowchart Research

This study utilized Aspen Plus V-12 software to develop the process simulation for CO_2 hydrogenation to methanol. Primary data, including CO_2 and H_2 stream specifications, were sourced from refinery operations, ensuring the model reflects real-world feed conditions. Secondary data, such as thermodynamic and kinetic parameters, were derived from previous studies and engineering best practices to enhance simulation accuracy and reliability. The research was divided into two main stages: the first focused on the process simulation of CO_2 hydrogenation to methanol, while the second stage involved the economic evaluation of the methanol plant. Key variables influencing the process are categorized into fixed, independent, and dependent variables, as summarized in Table 3.1. The fixed variables include the CO_2 flowrate and composition obtained from steam methane reforming (SMR) and the H_2 flowrate from naphtha reforming, as well as the catalyst type, $CuO/ZnO/Al_2O_3$. Independent variables encompass the H_2/CO_2 molar ratio, set at 3, reactor temperature ranging between 200-300°C, and reactor pressure maintained at 50 bar. Dependent variables include methanol yield, methanol purity of ≥99.85 wt%, and overall yield and conversion efficiency. Together, these variables form the foundation for evaluating the technical performance and economic feasibility of the methanol synthesis process.

Tabel 1 Variable Research

Fixed Variables	Independent Variables	Dependent Variables
- H_2 Flowrate and Concentration (From Naphtha Reforming Side Reaction)	- Reactor Inlet Temperature (200–300°C; Cu-based catalysts)	- Methanol Yield
- H_2/CO_2 Molar Ratio = 3	- Distillation Column Operating Conditions	- Methanol Purity ≥ 99.85 wt% (Grade AA; Commercial Standard)
- CO_2 Flowrate and Composition (From Steam Reforming Tail Gas)		
- Catalyst Type: $CuO/ZnO/Al_2O_3$		
- Reactor Type: Isothermal		
- Reactor Pressure: 50 bar		
- Liquid Hourly Space Velocity (LHSV)		

3.1 Process Simulation

The methanol synthesis process from CO_2 and H2 was simulated using Aspen Plus V-12. The reaction kinetics, based on Graaf et al. (1986), included the following reversible reactions:

R1: $CO + 2H_2 \Leftrightarrow CH_3OH \quad \Delta H_{298} = -21{,}7\ kcal/mol$ [3. 1]

R2: $CO_2 + H_2 \Leftrightarrow CO + H_2O \quad \Delta H_{298} = 9{,}8\ kcal/mol$ [3. 2]

R3: $CO_2 + 3H_2 \Leftrightarrow CH_3OH + H_2O \quad \Delta H_{298} = -11{,}9\ kcal/mol$ [3. 3]

Graaf et al. (1986) demonstrated that the chemical equilibrium for methanol formation can be well described using relationships derived from Gibbs free energy. This model assumes ideal gas behavior but corrects for non-ideal conditions at high pressures using the **Soave-Redlich-Kwong (SRK)** equation of state. Lim et al. (2009) later updated Graaf's experimental data, providing slightly adjusted equilibrium constants as follows:

$$lnK_1 = \frac{9{,}8438\ x\ 10^4\ J/mol}{RT} - 29{,}07 \quad [3.\ 4]$$

$$K_1 = 2{,}3717\ x\ 10^{-13} exp\left(\frac{9{,}8438\ x\ 10^4}{RT}\right)[atm^{-2}] \quad [3.\ 5]$$

$$lnK_1 = -52{,}096 + \frac{11840}{T}\ dengan\ K_A[Pa^{-2}] \quad [3.\ 6]$$

$$lnK_2 = \frac{-4{,}3939x10^4\ J/mol}{RT} + 5{,}639 \quad [3.\ 7]$$

$$K_2 = 2{,}8118\ x\ 10^2 exp\left(\frac{-4{,}3939\ x\ 10^4}{RT}\right) \quad [3.\ 8]$$

$$lnK_2 = 5{,}639 + \frac{-5285}{T}\ dengan\ K_B[-] \quad [3.\ 9]$$

$$K_3 = K_1 x K_2 \quad [3.\ 10]$$

$$K_3 = 6{,}6688\ x\ 10^{-11} exp\left(\frac{5{,}4499\ x\ 10^4}{RT}\right) \quad [3.\ 11]$$

$$lnK_3 = -46{,}457 + \frac{6555}{T}\ dengan\ K_B[Pa^{-2}] \quad [3.\ 12]$$

The reaction kinetics for methanol production via CO_2 hydrogenation using a Cu/Zn/Al/Zr catalyst were initially provided by Graaf et al. (1988) and later refined by An et al. (2009), resulting in updated rate constants:

$$k_1 = 4{,}0638\ x\ 10^{-6} exp\left(\frac{-11695}{R.T}\right), \frac{kmol}{Pa^{1.5}.\ kg_{cat}.\ s} \quad [3.\ 13]$$

$$k_2 = 9{,}0421\ x\ 10^{8} exp\left(\frac{-11695}{R.T}\right), \frac{kmol}{Pa^{0.5}.\ kg_{cat}.\ s} \quad [3.\ 14]$$

$$k_3 = 1{,}5188\ x\ 10^{-33} exp\left(\frac{-11695}{R.T}\right), \frac{kmol}{Pa.\ kg_{cat}.\ s} \quad [3.\ 15]$$

$$K_{CO} = 8{,}3965\ x\ 10^{-11} exp\left(\frac{118270}{R.T}\right), Pa^{-1} \quad [3.\ 16]$$

$$K_{CO_2} = 1{,}7214\ x\ 10^{-10} exp\left(\frac{81287}{R.T}\right), Pa^{-1} \quad [3.\ 17]$$

$$\frac{K_{H_2O}}{K_{H_2}^{\frac{1}{2}}} = 4{,}367\ x\ 10^{-12} exp\left(\frac{84000}{R.T}\right), Pa^{-0.5} \quad [3.\ 18]$$

The methanol plant process flow diagram used in this simulation is based on the Aspen Plus Flowsheet with energy-efficient methanol production through CO_2 hydrogenation. The flow diagram is illustrated in Figure 4 (Dimian, 2019).

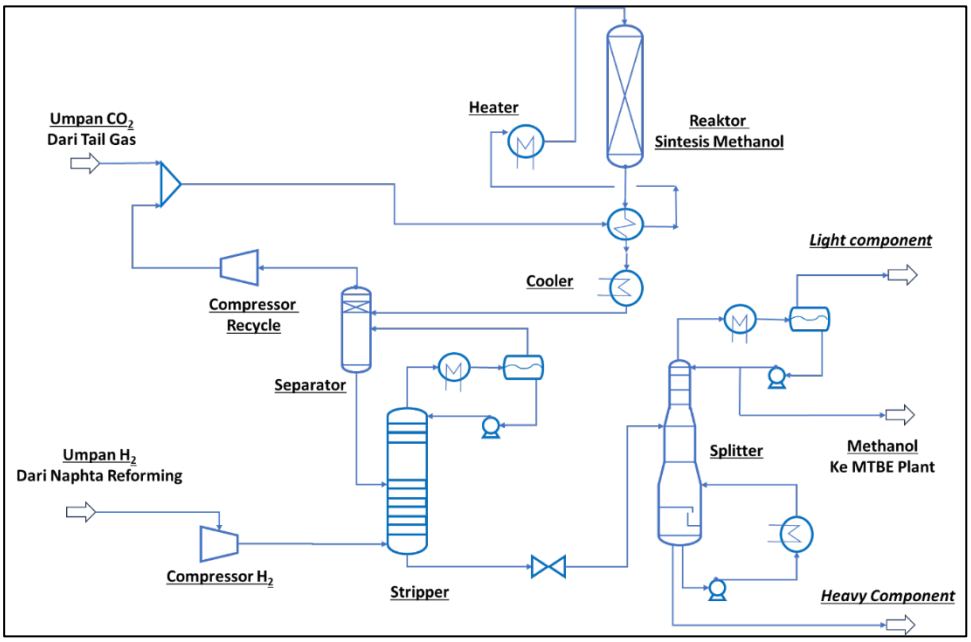


Figure 4: Process Flow Diagram of the Methanol Plant

The model employed an isothermal reactor using $Cu/ZnO/Al_2O_3$ catalyst at pressures of 50 bar and optimized temperature ranges (200-300°C). A downstream distillation column was modeled to achieve methanol purity of ≥99.85 wt%.

3.2 Economic Evaluation

The investment cost estimation for the construction of the methanol plant was conducted using ASPEN Process Economic Analyzer (APEA) V-12. The cost estimation in APEA is based on the process flow diagram (PFD) and equipment configurations simulated in Aspen Plus V-12. The economic analysis assumes the following project conditions plant Life is 25 years, EPC Duration is 3 years and engineering Start Date is January 1, 2025

Economic feasibility was assessed using key parameters Internal Rate of Return (IRR) Net Present Value (NPV) Pay-Out Time (POT). Sensitivity analysis was conducted to evaluate project viability under varying economic conditions. Parameters analyzed include sales price (± 20 % from the base case), production rate (± 20 % from the base case), and feed cost (- 10 to + 30% from the base case).

4. RESULTS AND DISCUSSION

4.1 Methanol Synthesis Process

The methanol synthesis process adopts the Kiss (2016) scheme, simulated using Aspen Plus 12, using two key reactant streams: tail gas from Steam Methane Reforming (SMR) and hydrogen from Naphtha Reforming.

Table 2: Operating Condition of Feedstock Methanol Plant

Parameter	Unit	Tail Gas
Fase		Vapor
Laju alir	Kg/jam	88.534
Temperatur	°C	25
Tekanan	Kg/cm^2.g	0,3
Komposisi		
CO	%mol	11,8%
CO_2	%mol	46,6%
H_2O	%mol	0,9%
H_2	%mol	24,9%
CH_4	%mol	15,6%
N_2	%mol	0,1%

Parameter	Unit	Hydrogen
Fase		Vapor
Laju alir	Kmol/hr	5.795,75
Temperatur	oC	45
Tekanan	Kg/cm2.g	24
Komposisi		
CO	%mol	
CO_2	%mol	
H_2O	%mol	
H_2	%mol	99,93 %
CH_4	%mol	0,05 %
N_2	%mol	0,02 %

The feed comprises tail gas at 88,534 kg/hr with 46.6% CO_2 content at a low pressure of 0.3 kg/cm^2.g. The process also involves hydrogen from naphtha reforming, with a purity of 99.93% vol at 24 kg/cm^2.g and 45°C.

Given the high-pressure reactor conditions (50 kg/cm^2.g), tail gas compression is required with a compression ratio of 167. Notably, the presence of 11.8% CO in the feed improves methanol yield as CO converts more readily compared to CO_2 due to CO_2's inert nature.

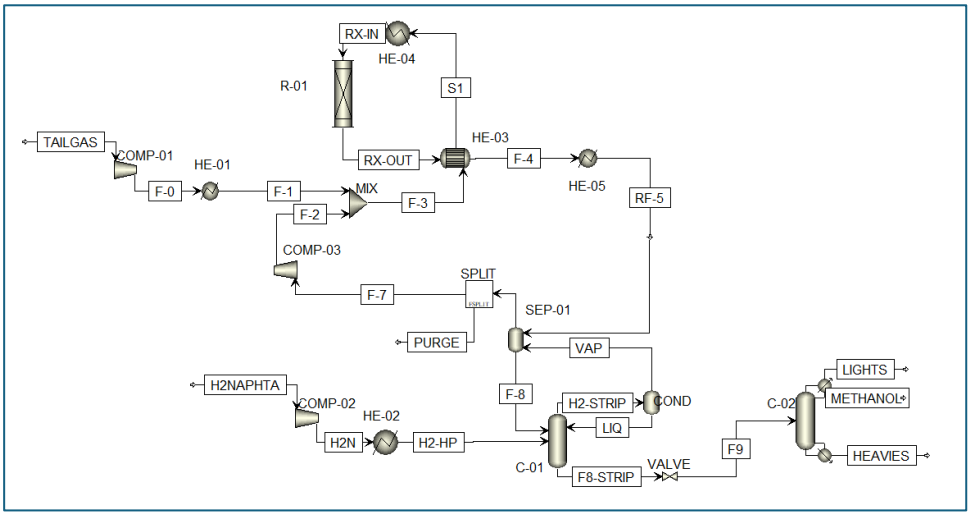


Figure 5: Methanol Simulation Aspen Plus

4.2 Effect of Reactor Temperature on Methanol Yield

Simulations were conducted over a reactor temperature range of 200–300°C at a constant pressure of 50 kg/cm^2.g. The optimum temperature for methanol synthesis is 235°C, achieving a maximum methanol yield of 81.16% (equivalent to 70 tons/hr or 612 KTA).

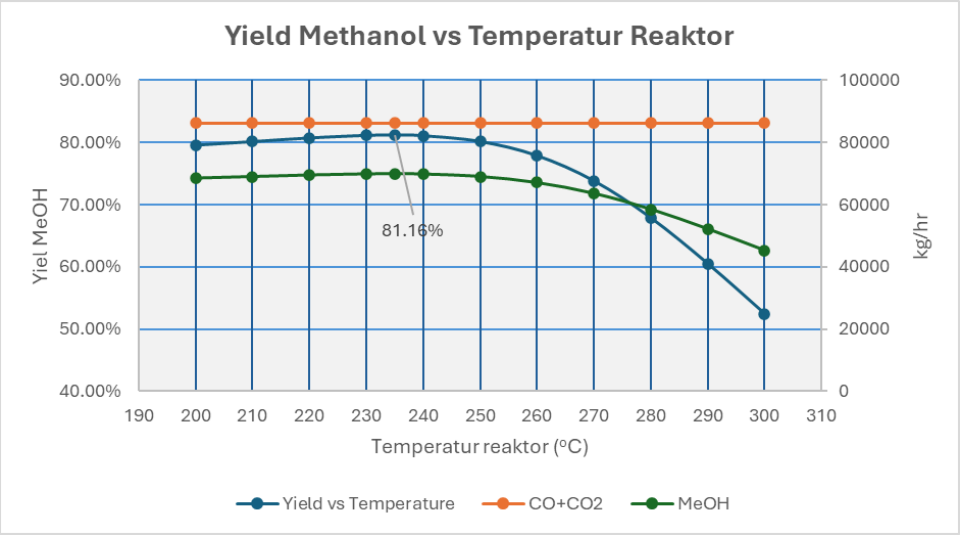


Figure 6: Yield Methanol Vs Temperature Reactor

The involvement of CO from tail gas enhances yield due to its easier conversion to methanol compared to CO_2, which is more inert. However, temperature control is critical as Reactions R1 and R2 (CO and CO_2 hydrogenation) are exothermic, and yield decreases above 250°C due to equilibrium limitations (Kiss, 2016). The reverse water-gas shift reaction (R3) is endothermic, benefiting from higher temperatures.

Methanol purification is carried out in distillation column C-02 to achieve a product purity of 99.85% wt. The feed to the column is a methanol-water mixture with minor dissolved gases, having a flow rate of 110,385 kg/hr, methanol content of 63.33% wt, and water content of 36.63% wt. The feed enters the column at tray number 23 out of 30 trays. The operating conditions of the column were determined through simulations in Aspen Plus and further optimized trials. The column operates at a pressure of 5 kg/cm^2.g, with a top temperature of 62.7°C and a bottom temperature of 124°C. The reflux ratio is set at 0.81, ensuring efficient methanol purification while minimizing energy consumption, while the boil-up ratio is 1.7, maintaining sufficient vapor circulation within the column. A partial vapor-liquid condenser is used to separate light gases from methanol vapor at the top of the column, while a kettle-type reboiler ensures stable and uniform heating at the bottom. Under these optimized operating conditions, the column produces methanol at a flow rate of 69,805 kg/hr

(approximately 70 tons/hr or 612 KTA) with a purity of 99.85% wt, meeting the global Grade AA methanol specifications.

4.3 Economic Analysis and Sensitivity

The investment assumptions include a plant life of 25 years, an EPC duration of 3 years, and the start of basic engineering on January 1, 2025. The estimated Inside Battery Limit (ISBL) capital cost is USD 89,885,000.

The total capital investment is broken down as follows:

Table 3: Project Cost Estimate

No	Komponen Biaya	Rumus	Angka
1.	*FIXED CAPITAL INVESTMENT (FCI)*		
	a. Inside Battery Limmit (ISBL) Investment	C_2	$89,885,000
	b. Outside Battery Limmit (OSBL) Invevestment	30% C_2	$44,942,500
	c. Engineering Cost	20% (ISBL + OSBL)	$40,448,250
	d. Contingency	10% (ISBL + OSBL)	$13,482,750
	TOTAL		**$188,758,500**
2	*WORKING CAPITAL*	5 % FCI	$9,437,925
	TOTAL		$9,437,925
3	*VARIABLE COST OF PRODUCTION*	*Raw Material Cost + Utility + Consumable*	$215,066,426
	TOTAL		**$215,066,426**
3	*FIXED COST OF PRODUCTION*		
	a. Operating labor	Pedoman Inkindo	$299,871
	b. Supervision	25 % Operating Labor	$74,968
	c. Direct Salary Overhead	50 % operating labour + Supervision	$187,419
	d. Maintenance	4 % ISBL Investment	$3,595,400
	e. Property taxes and insirance	1,5 % ISBL Investment	$1,797,700
	f. General plant overhead	65 % *total labour* (operating, supervision and maintenance)	$2,580,655
	g. Allocated environment charges	1 % (ISBL + OSBL *Cost*)	$1,348,275
	TOTAL		**$9,884,288**

The project economic feasibility analysis assumes 100% equity financing and considers revenue only from methanol sales. The methanol price is based on the Global Methanol Long-Term Price Forecast from Wood Mackenzie, with an average price of USD 398/ton from 2024–2050.

The economic feasibility of the methanol plant project was evaluated based on key financial indicators. Over the plant's lifespan, the total revenue is estimated to be USD 6,089,400,000**,** while the variable cost of production, which includes raw materials, utilities, and consumables, amounts to USD 5,376,660,660. The fixed cost of production is USD 247,107,206, which includes expenses for labor, maintenance, and overhead. Despite the projected revenue, the analysis resulted in a negative Net Present Value (NPV) of USD -26,485,422.10, indicating a net loss over the investment period. The Internal Rate of Return (IRR) was calculated at 8.04%**,** which falls below the hurdle rate of 10.83%**,** making the project financially unviable under current conditions.

Furthermore, the Pay Out Time (POT), representing the time required to recover the investment, was found to be 22 years, which is considerably longer than acceptable benchmarks for petrochemical projects. These results highlight the economic challenges of the project, primarily driven by high production costs and significant capital investment.

The key contributors to the project's economic limitations are High Variable Cost of Production and High Fixed Capital Investment. Variable Cost includes feedstock and utility costs, amounting to USD 215,066,426. Feedstock costs dominate, with hydrogen priced at USD 1.4/kg, resulting in an annual cost of USD 143,336,726 (67% of variable costs). Utilities contribute USD 71,729,700 (33% of variable costs). Inaddition, the total fixed capital cost reaches USD 188,758,500, driven by extreme operating conditions—50 kg/cm^2.g pressure and 200–300°C temperature—which increase the material and equipment costs.

A sensitivity analysis was performed to evaluate the impact of critical economic parameters on the Internal Rate of Return (IRR). The sensitivity analysis results are illustrated in Figure 7, showing the IRR response to parameter variations.

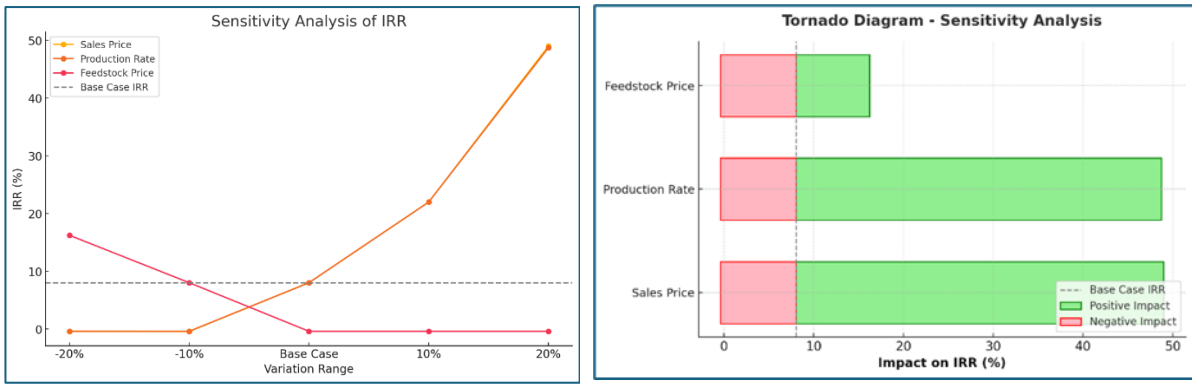


Figure 7: Sensitivity Analysis IRR

The sensitivity analysis revealed that the methanol sales price has the most significant impact on the Internal Rate of Return (IRR). A 20% increase in the methanol price raises the IRR to 49%, making the project highly feasible, while a 20% decrease reduces the IRR to -0.37%**,** rendering the project economically unviable. Similarly, production capacity plays a critical role in improving project profitability**.** An increase of 10% in production capacity raises the IRR to 22%**,** while a 20% increase results in a substantial IRR of 48.72%**,** highlighting the importance of maximizing production rates. Conversely, a 20% reduction in production capacity causes the IRR to remain at **-**0.37%**,** reinforcing the need to maintain or enhance output levels. On the other hand, feedstock cost has a moderate impact on IRR. A 10% reduction in feedstock cost improves the IRR to 16.24%**,** but an increase in feedstock cost by 10–30% significantly suppresses the IRR, making the project economically unviable. These findings emphasize that optimizing the methanol sales price and production capacity are key

strategies to improve project feasibility, while controlling feedstock costs remains a supporting factor.

The sensitivity analysis confirms that methanol sales price has the most significant impact on project feasibility, followed by production capacity and feedstock cost. Optimizing the sales price through long-term contracts and increasing production capacity are critical strategies to improve the IRR. Although feedstock costs have a moderate impact, reducing hydrogen costs through negotiations or alternative sources can also enhance project viability.

5. CONCLUSIONS AND RECOMMENDATIONS

5.1 Conclusions

Based on the research and analysis conducted on the utilization of CO_2 from the Steam Reforming Unit tail gas for methanol production at an oil refinery, the following key conclusions were obtained:

a. Technical Performance of Methanol Synthesis:
   The methanol synthesis process utilizing CO_2 from the Steam Reforming Unit tail gas and H_2 from the Naphtha Reforming Unit achieved the desired results. The maximum methanol yield was obtained at a reactor temperature of 235°C, with a yield of 81.16% or equivalent to 612 KTA (70 tons/hour). The process employs a $CuO/ZnO/Al_2O_3$ catalyst under an operating pressure of 50 bar. The methanol purification process using a distillation column produced methanol with a purity of ≥99.85% wt (Grade AA), meeting global commercial specifications.

b. Economic Evaluation:
   The economic evaluation showed that the methanol plant project resulted in an IRR of 8.04%, a negative NPV of USD -26,485,422.10, and a Pay Out Time (POT) of 22 years. These values do not meet the investment feasibility criteria, as the IRR remains below the hurdle rate of 10.83%. The primary factors contributing to the project's low economic viability are the high feedstock cost and significant fixed capital investment. Feedstock, primarily hydrogen, accounts for 67% of the variable production costs, while the fixed capital investment reaches USD 188,758,500.

c. Sensitivity Analysis:
   The sensitivity analysis demonstrated that methanol sales price is the most significant variable affecting IRR. A 20% increase in methanol price from the baseline (USD 398/ton) raises IRR to 49%, well above the hurdle rate. The second most influential variable is production capacity. A 20% increase in production capacity results in an IRR of 48.72%, surpassing the hurdle rate. Feedstock cost has a moderate impact on IRR. A 10% reduction in feedstock price improves IRR to 16.24%, while increases of 10–30% suppress IRR to below 0%, making the project economically unfeasible.

5.2 Recommendations

a. Economic Optimization of the Project:
Optimize the production capacity of the plant to achieve economies of scale, targeting an increase of 10–20% to improve the IRR.

b. Feedstock Cost Management:
Evaluate the use of alternative feedstock sources that are more cost-effective or seek lower-cost hydrogen supplies through integration with other refinery units.

c. Integration with Downstream Products:
Integrate the methanol plant with an MTBE production facility to enhance economic value. MTBE can be used as a gasoline blending component to increase octane numbers, thereby expanding the diversification of petrochemical products within the refinery.

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