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Preliminary study of gas-to-liquid (GTL) plant development in Indonesia

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Abstract

Domestic fuel consumption has long been elevated with the numbers that could not stand straight (non-sustainable). The growth rate of fuel consumption in Indonesia reaches about 7% a year. With this continuous growth rate, its consumption has increased twice every ten years. But, the amount of oil production decreases from year to year and even will deplete in 15 – 20 years more (Ministry of Energy and Mineral Resources Republik Indonesia, 2010). In 2003, Indonesia faced the oil deficit for the first time where the consumption was higher than its production amount and in 2009 the national reserves couldn't meet the national needs that have to import oil about 176 kbpd. This condition continues to deteriorate. In 2010, Indonesia's oil production was recorded only 986 kbpd when the consumption level jumped to 1,304 kbpd or deficit about 318 kbpd (BP Statistical Review, 2011). On the other hand, Indonesia natural gas production in 2008 reached 7,883 MMSCFD which produced from 43 work areas of cooperation contract (KKS). Cumulatively, gas production in 2008 increased by 2.56% compared to 2007 which was only 7,686 MMSCFD (Ministry of Energy and Mineral Resources Republik Indonesia, 2009). The utilization of natural gas potential has been conducted such as for electricity generator, raw material for petrochemical industry, fuel for industry, household needs, etc. Unfortunately, direct utilization of natural gas for transportation has not developed well because most of the vehicles still use liquid fuel from oil. One technology that can be implemented to convert natural gas into liquid fuel (synthesis fuel) is by Gas to Liquid (GTL) technology by Fischer-Tropsch reaction. However, the high investment cost resists the establishment of GTL plant in Indonesia. One, which is addressed in this paper, is the preliminary study of establishment of GTL plant in Indonesia. The study is started by synthesizing and analyzing the process design of the plant by using Aspen Hysis V7.1. and followed by the economical consideration of the plant. The results of this study will show whether the GTL plant development in Indonesia can be implemented or not, which then become a potential consideration for Indonesia to build the GTL plant.

Keywords: Gas to Liquid, Development, Fischer-Tropsch, Preliminary Study

1. Introduction

1.1. Indonesia Oil Condition

Indonesia's oil production was formally governed by a quota allocation from OPEC. At the March 1991 OPEC ministerial meeting, Indonesia's quota was set at 1.445 million barrels per day, below the country's estimated production capacity of 1.7 million barrels per day. Indonesia's quota represented about 6 percent of total OPEC production. About 70 percent of Indonesia's annual oil production was exported on average during the late 1980s, but domestic consumption was increasing steadily and reached half of annual oil production by 1990 (Frederick and Worden, 1993).

Sumatra accounts for more than half of Indonesia's oil production, coming from Riau province, in West and Central Sumatra, from Aceh province in the North, which also oversees the administration of Natuna islands, and from South Sumatra. Java, the most populated island in Indonesia, is by far the biggest oil and gas consumer in this country. Parts of the island and the adjacent Java Sea form an oil and gas province. Numerous oil and gas

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fields have been located offshore. East Kalimantan, on the island of Borneo, is the other main petroleum-producing region (Indonesian Petroleum Association, 2009). The figure 1 below shows the Indonesia oil reserves in 2010 based on statistical data of Ministry of Energy and Mineral Resources RI.

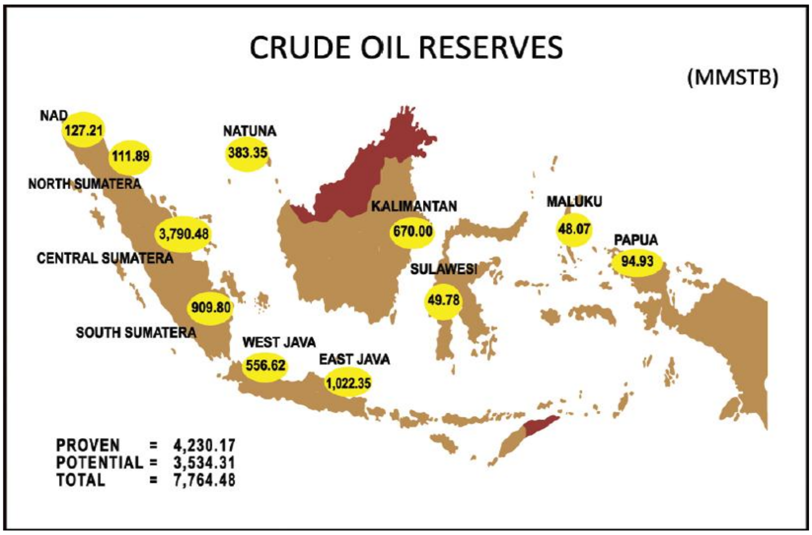

Figure 1. Indonesia Oil Reserves in 2010 (Indonesia Energy Statistic Ministry of Energy and Mineral Resources RI, 2010)

As the growth of petroleum consumption in Indonesia, with an average increase in consumption of 7% per year, the consumption of petroleum has been unable to be met by Indonesia's oil production (Ministry of Energy and Mineral Resources Republik Indonesia, 2011). In 2003, the deficit for the first time happened in Indonesia where oil consumption rate exceeds the production rate. In 2004, this deficiency can’t be covered again from the national reserve, so that Indonesia should also cover 176 kbpd of deficit by importing oil from overseas. This note continues to deteriorate. In 2010, Indonesia's oil production was recorded only 986 kbpd while the consumption level jumped to 1,304 kbpd or 318 kbpd of deficit (BP Statistical Review, 2011).

1.2. Indonesia Gas Condition

Indonesia Natural Gas Production in 2008 reached 7883 which produced from 43 MMSCFD Work Areas of Cooperation Contract (KKS). Accumulatively, in 2008, the gas production increased about 2.56% compared to 2007 which was only 7,686 MMSCFD (Ministry of Energy and Mineral Resources Republik Indonesia, 2009).

Indonesia enjoys a huge amount of untapped natural gas reserves, the result of 30 years exploration and investment. The two largest gas reserves are located in Arun, North Sumatra, and Badak, East Kalimantan. Gas from these two producing areas supports the production of LNG for export to East Asian industrial centers (Indonesian Petroleum Association, 2009). The figure 2 below shows Indonesia gas potential in 2010.

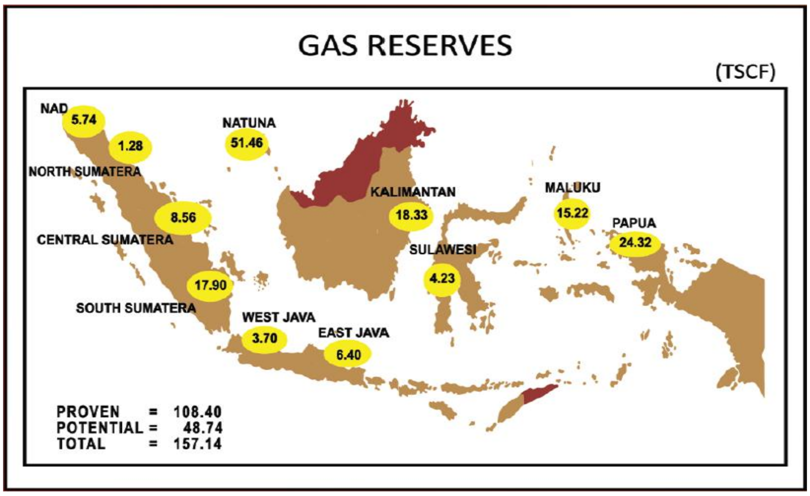

Figure 2. Indonesia Gas Reserves in 2010 (Indonesia Energy Statistic Ministry of Energy and Mineral Resources RI, 2010)

Of the Natural Gas produced in Indonesia, 54% is exported in the form of LNG and LPG, with the reminder targeted for domestic consumption. Annual production volume in 2008 was 8.3 billion cubic feet per day (or 1.4 MBOEPD), which will continue as development projects such as those in South Sumatra and Tangguh LNG in West Papua come on stream and reach full production. Indonesia's vast gas reserves of 164.9 TCF represent a great opportunity for the nation's future growth (Indonesian Petroleum Association, 2009).

1.3. GTL Technology

Based on the potential of Indonesia's natural gas reserves and also problems in the lack of oil production compared to consumption, one technology that can be implemented to produce Synthetic fuel (synthetic oil) is Gas-to-Liquid (GTL) technology. Gas-to-Liquid (GTL) fuels can be produced from natural gas by using Fischer-Tropsch chemical reaction process. The liquids produces include naphtha, diesel, etc. which is contained also in conventional oil.

The term gas-to-liquids refers to a small number of technologies designed to convert natural gas to liquid fuels, as alternatives to the traditional refining of crude oil and other natural gas commercialization routes. Natural gas is four times more expensive to transport than oil. Converting remote natural gas into a liquid reduces its cost. Declining GTL production costs, growing worldwide diesel demand, and stringent diesel exhaust emission standards and fuel specifications are driving the petroleum industry to revisit the GTL process to produce higher quality diesel fuels (California Energy Emission, 2006).

GTL fuel can be used neat or blended with today's fuel and used in existing engines and infrastructure. These fuels provide an opportunity to reduce dependence on petroleum based fuels and reduce tailpipe emissions. GTL fuel has virtually no sulfur, aromatics, or toxics. It can be blended with non-complying fuel to make the fuel cleaner so it will comply with new fuel standards (California Energy Emission, 2006). The graph below shows the projection of GTL plant development from working document of National Petroleum Council (NPC) in 2007.

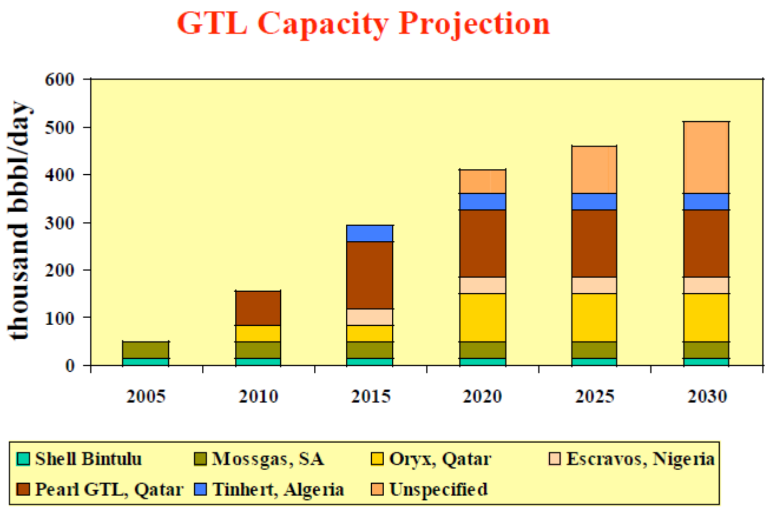

Figure 3. GTL Capacity Projection (National Petroleum Council, 2007)

2. Method

This paper will show the preliminary study of GTL plant development in Indonesia. The study is started by simulation conducted to design the process of GTL. The simulation uses Hysis V7.1. which then followed by the analysis of economic of the plant. The economic analysis is conducted to see the total capital investment of the designed GTL process by using Modular Guthrie Method. Furthermore, the study is continued by economic scenario and feasibility to show whether the plant is possible to be implemented or not. The scenario including scale up of the plant is made by using Six Tenth Rules method. This scenario is conducted to see the possible price of crude syn-fuel and also the feasibility by looking the Capital Charge Factor (CCF) of the plant. Finally, the sensitivity analysis is made to show which one of the component of the plant design and establishment cost which is sensitive to one parameter or more.

3. Discussion and Results

3.1. Process Description

Process in Gas-to-Liquid plant can vary from one design to another. Basically, it contains one unit of air separation, one unit of synthesis gas reformer, and one unit of Fischer-Tropsch reactor and other improvisation for optimization and further product upgrading. The block flow diagram of the process in this is illustrated by the following figure.

The process is started by air separation unit. This unit functions to separate Oxygen from air which then used in synthesis gas reforming process. Air separation unit has two types of process, cryogenic and non-cryogenic. Cryogenic air separation processes are routinely used in medium to large scale plants to produce nitrogen, oxygen, and argon as gases and/ or liquid products. Cryogenic air separation is the preferred technology for producing very high purity oxygen and nitrogen. Non-cryogenic processes use physical properties other than boiling point to separate and purify components of air at close-to-ambient temperature. Systems belong to one of two major technology categories: adsorption processes and membrane diffusion-separation systems (Universal Industrial Gas, 2011).

In this paper, air separation unit used in the process is the non-cryogenic process. The consideration on choosing this system is because the Oxygen needed in the synthesis gas reforming didn't need the high purity of Oxygen (99.9%). The 94% purity of oxygen can be considerable which can be produced by using non-cryogenic air separation unit. Equipment used in this process is pressure swing adsorption which run simultaneously and alternately from four high pressure adsorption vessel.

Adsorption air separation (Pressure Swing Adsorption/PSA) processes rely on the fact that under pressure gases tend to be attracted to solid surfaces or adsorbed. The higher the pressure, the more gas adsorbed. PSA processes can be used to separate gases in a mixture because different gases tend to be attracted to different solid surfaces more or less strongly. For example, if a gas mixture such as air is passed under pressure through a vessel containing an adsorbent bed that attracts Nitrogen more strongly than it does Oxygen, part or all of the Nitrogen will stay in the bed, and the gas coming out of the vessel will be enriched in Oxygen. When the bed reaches the end of its capacity to adsorb Nitrogen, it can be regenerated by reducing the pressure, thereby releasing the adsorbed Nitrogen. It is the ready for another cycle of producing Oxygen enriched air (Chemsystems Perp. Program – Nexant, 2010).

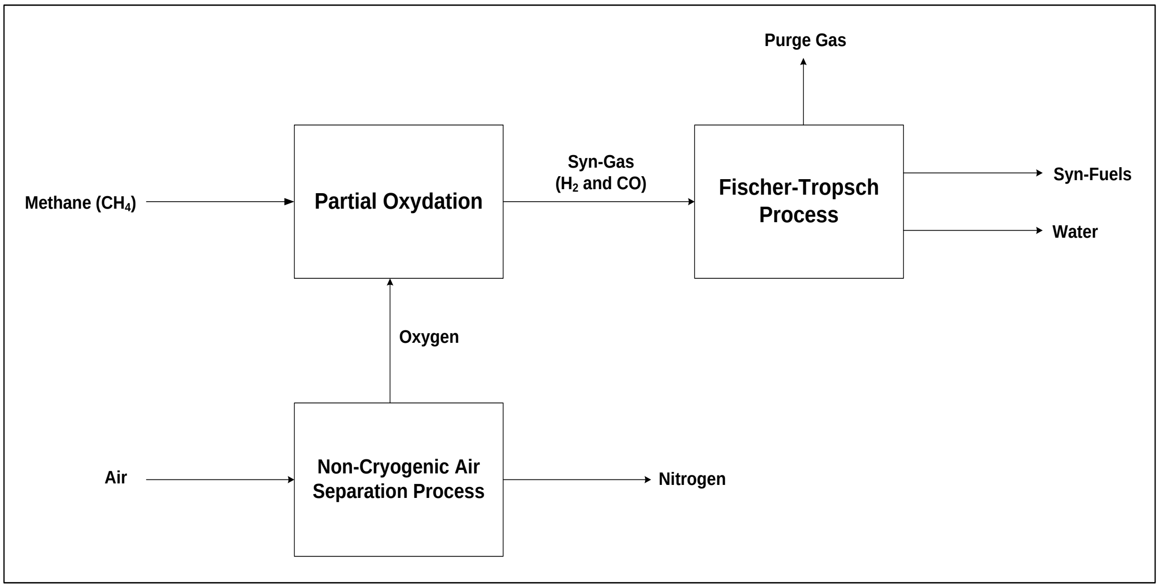

Figure 4. Block Flow Diagram

The process is continued by synthesis gas reforming. This process aims to produce a synthesis gas (CO and H_2) that will be used for Fischer-Tropsch process. There are many ways to produce synthesis gas such as steam methane reforming, partial oxidation process, CO_2 reforming process, or even combination of those processes. The comparison of those types of synthesis gas reforming is shown by the following tables.

Table 1. Performance Comparison Table of Syn-Gas Production Technology (Holladay et al, 2009)

Technology	Advantages	Disadvantages
Steam reforming	• Does not require oxygen • Low process temperature • Highest ratio of CO/H_2 gas	• Highest emission of CO_2
Autothermal reforming	• The process temperature is lower than the POX • Low methane slip	• Requiring air or oxygen • Commercial use is still limited
Partial oxidation	• Does not require a catalyst • Reducing the need of desulfurization • Low methane slip	• Low H_2/CO ratio • Requiring high process temperature

Table 2. Comparison of Synthesis Gas Reforming Process

	Steam Methane Reforming	**Partial Oxidation**	**CO_2 Reforming**
Temperature, ^oC	800-900	1000-1450	900-1000
Pressure, bar	20-30	30-85	10
H_2/CO Ratio	3-6	1.6-2	1
CH_4 Conversion, %	65-95	95-100	---
Oxygen	---	High	---
Steam Consumption	High	Optional	Optional
Investment, %	100	80-110	---
Emission	High	Low	Low
Scale	Big	Small to Big	Medium
Status	Commercial	Commercial	Commercial

The process chosen in this paper for synthesis gas reforming is partial oxidation process. The consideration on taking partial oxidation as the process is because the ratio of H_2:CO produced in this process is nearly to 2 which is needed in the Fischer-Tropsch reaction to produce synthesis fuel. In partial oxidation, the feed is burned in a restricted amount of Oxygen. The chemical reaction of partial oxidation is as follows:

$$CH_4 + \tfrac{1}{2} O_2 \rightarrow CO + 2 H_2 \qquad \Delta H = -35.9 \text{ kJ/mole } (\Delta H < 0)$$

Oxygen and natural gas are separately heated and mixed at the burner tip of a combustion lance held within the partial oxidation pressure vessel. The temperature in the flame is very high. This temperature favors the formation of Carbon Monoxide over Carbon Dioxide. Consequently, partial oxidation produces a synthesis gas containing very little Carbon Dioxide with almost no Methane slippage. The gas stoichiometric ratio is about 1.8 H_2:CO. Increasing Oxygen input and adding some Carbon Dioxide can move this value down to about 1, and addition of water can increase the value to about 2 (Seddon, 2006).

After synthesis gas formed, the next process is the Fischer-Tropsch process by following the reaction to produce synthesis fuel. The principal chemical reactions of the FT synthesis are:

Synthesis: $nCO + 2nH_2 \rightarrow -[CH_2]_n- + nH_2O$

Methanation: $CO + 3H_2 \rightarrow CH_4 + H_2O$

Water Gas Shift: $CO + 3H_2 \rightarrow CO_2 + H_2$

In the synthesis reaction, $-[CH_2]_n-$ represents all of the hydrocarbon molecules from Methane to Paraffin wax of high molecular weight. Methanation represents Methane made in the process over and above that produced by the synthesis reaction. The presence of Water Gas Shift (WGS) reaction during synthesis can be an advantage in helping to generate any extra Hydrogen needed. Carbon Dioxide is also a product when the WGS reaction is present (Seddon, 2006).

3.2. Process Simulation

The simulation is conducted by using Hysis V7.1. The raw materials of the process are methane gas and also air. The product will be syn-fuel with other side products such as purge gas and water. The figure bellow shows the main process flow diagram of the GTL plant.

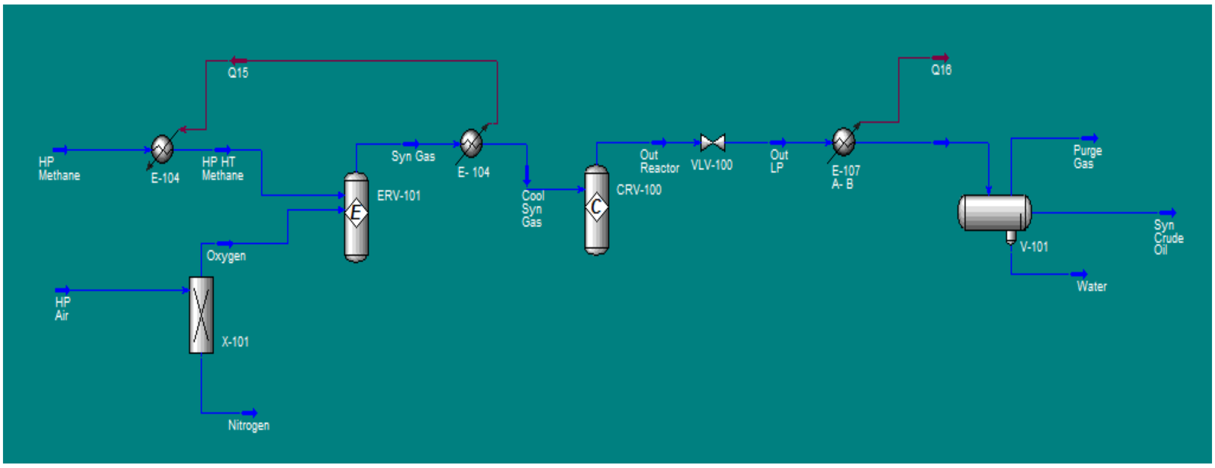

Figure 5. Process Flow Diagram by Using Hysis V7.1

Before the raw materials enter the main process, there are pre-treatment processes which are conducted so that the raw materials meet the condition close to the main process condition. Methane is compressed by using three stage compressors with intercoolers to 30 bars and air is compressed also by using three stage compressors with intercoolers to 30 bars pressure which then separated to produce Oxygen in high purity. The process flow of the pre-treatment process is shown in figures below.

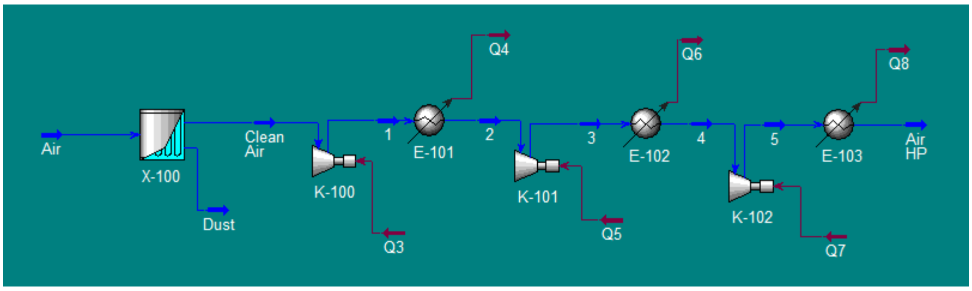

Figure 6. Air Pre-Treatment Process

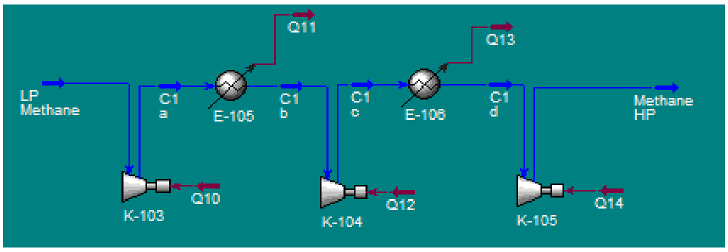

Figure 7. Methane Pre-Treatment Process

The results of the process simulation design are shown in table 3 below. The table shows the raw materials and also products condition and components. The methane used as raw materials is about 12.51 MMSCFD (10,000 kg/hr) and the amount of air is followed by the stoichiometric amount which is about 29.77 MMSCFD (42,770 kg/hr). The air is assumed only Nitrogen and Oxygen substances with 79% of mole Nitrogen and 21% of mole Oxygen.

Table 3. Raw Materials and Products Conditions and Components

Parameter	**Raw Materials**		**Products**		
	Air	**Methane (CH_4)**	**Syn-Fuel**	**Purge Gas**	**Water**
P (bar)	1.01	1.01	1.12	1.12	1.12
T (^oC)	33.00	15.00	30.00	30.00	30.00
Mole Flow Rate (kmol/hour)	1,484.00	623.30	58.63	54.16	621.2
Standar Ideal Liquid Flow Rate (Barrel/day)	7,465.00	5,042.00	1,680.98	267.2	1,693.00
Mole fraction of CH_4	-	1.00	-	-	-
Mole Fraction of O_2	0.21	-	-	-	-
Mole Fraction of N_2	0.78	-	-	0.43	-
Mole Fraction of CO	-	-	-	-	-
Mole Fraction of H_2	-	-	-	0.52	-
Mole Fraction of Gasoline	-	-	0.52	0.01	-
Mole Fraction of Kerosene	-	-	0.21	-	-
Mole Fraction of Diesel	-	-	0.27	-	-
Mole Fraction of H_2O	-	-	-	0.04	1.00

The syn-fuel components in the process are simplified into only three major components of fuels such as gasoline (represented by n-Octane), kerosene (represented by Naphtalene), and diesel (represented by n-C16). The amount of gasoline is more than other components. The total fuel components produced are as follows:

Table 4. Crude Synthetic Fuel Produced

Fuel Component	**Volume (Barrel/Day)**
Gasoline	745.56
Kerosene	248.62
Diesel	686.80
Total	**1,680.98**

3.3. Economic Analysis

In the economic analysis will be explained about Total Capital Investment (TCI) calculation, economic scenario and feasibility, and sensitivity analysis. This study is conducted to show whether GTL plant can be implemented or not based on several considerations such as CCF (Capital Charge Factor) and cost payback period. Furthermore, in this economic study will show the crude Synthetic fuel price from GTL process and compare to the Brent crude oil price per April 2012. There are several assumptions used in this economic study, such as:

- Equipment cost is a cost in a variation of time. Hence, the cost used is based on year 2015 by using CE Index.
- The plant is run in 300 days per year.
- IRR (Internal Rate of Return) used is 15%, or 0.15.
- The production capacity is 1,681 barrel per day of crude syn-fuel.
- Cost is in USD ($).

3.3.1. Total Capital Investment Calculation

Total Capital Investment calculation can be done by using many ways. In this paper, the method used to calculate the TCI is Modular Guthrie method which can be formulated by the following equation (Seider et al, 2003).

$$C_{TCI} = 1{,}18\left(C_{TBM} + C_{site} + C_{buildings} + C_{ofssite\ facilities}\right) + C_{WC} \quad (1)$$

Where C_{TCI} is total capital investment cost, C_{TBM} is total bare module cost, C_{site} is site development cost, $C_{building}$ is building establishment cost, and $C_{offsite\ facilities}$ is offsite facilities development cost.

Total bare module cost can be calculated by summing all bare module cost of equipment used in the process. The calculation of TBM cost is started by calculating purchase equipment cost from all equipment in the year basis of 2015. The cost in 2015 is forecasted by using CE index. The bare module cost can be expressed by the following equation.

$$Total\ Bare\ Module\ Cost = \sum_{i=equipment} (Bare\ Module\ Cost) \qquad (2)$$

The calculation of purchased equipment cost and total bare module cost are shown in the appendix. The breakdown of Bare Module Equipment cost is shown in the figure 8 below. The biggest portion of the bare module cost is in syn-gas reforming process which includes air separation unit and also syn-gas reformer. The highest cost of equipment is reactors and compressors used in the process. In air separation unit and Methane preparation, price is high enough because the existence of compressors. Meanwhile, in FT Process and syn-gas reformer, price is high because the existence of reactors.

Site development cost is a cost to develop site. The type of the site is grass root plant with the amount of the cost is about 10 – 20% of total bare module cost. Building cost is a cost to build a building outside the process area in the plant. Building cost for non-process of grass root plant is about 20% of total bare module cost. Offsite facilities cost can be calculated by summing the utility cost and 5% of total bare module cost. Contingency is about 15% of TBM cost and contractor fee is about 3% of TBM cost. Working Capital Cost (C_{WC}) can be calculated by using following formula.

$$C_{WC} = 0{,}176 \times 1{,}18(C_{TBM} + C_{site} + C_{building} + C_{offsite\ facilities}) \qquad (3)$$

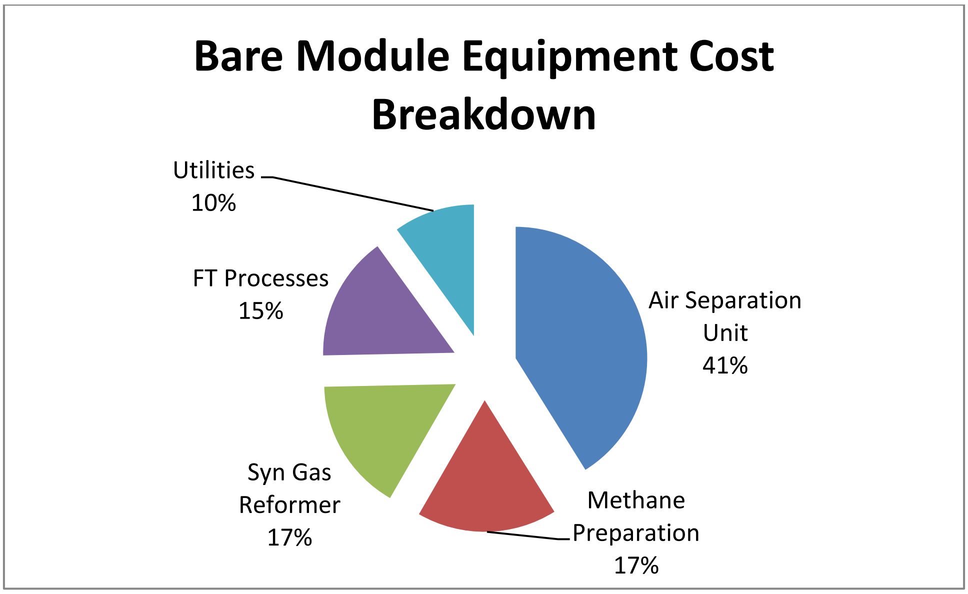

Figure 8. Bare Module Equipment Cost Breakdown

After calculating the component of Total Capital Investment cost, the TCI amount can be achieved by adding all of the components. The result from the calculation is shown in the following table and the breakdown of the TCI cost is shown in figure 9.

Table 5. Total Capital Investment Cost

Component	Value in $
Total Bare Modul Cost (C TBM) ($)	43,359,435.27
Site Development Cost (C site) ($)	8,671,887.05
Building Cost (C building) ($)	8,671,887.05
Offsite Facilities Cost (C offsite facilities) ($)	3,915,303.92
Contingency ($)	6,503,915.29
Contractor fee ($)	1,300,783.06
Working Capital (C WC) ($)	13,419,972.84
Total Capital Investment ($)	**97,474,516.89**

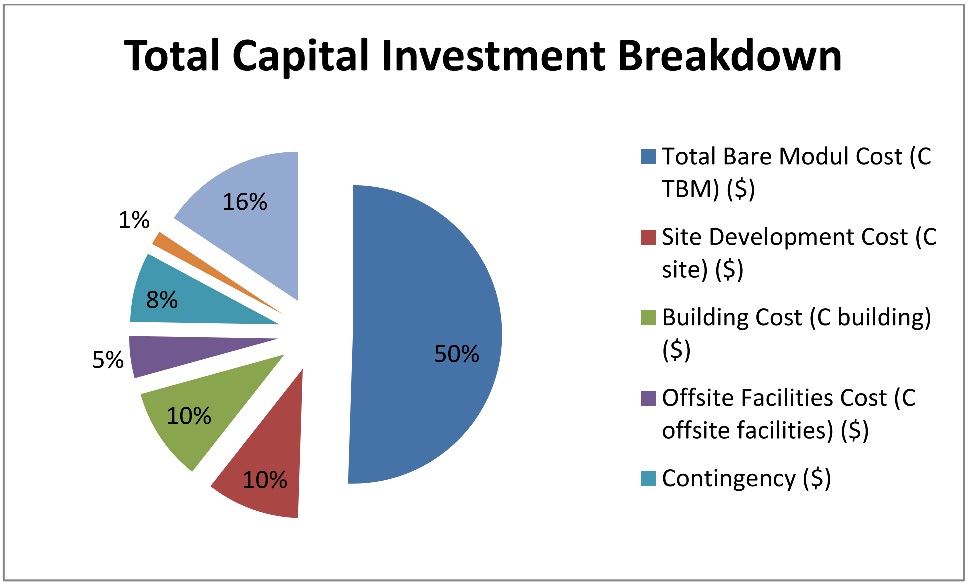

Figure 9. Total Capital Investment Breakdown

The previous study from Energy Information Administration, Department of Energy US, stated that the TCI cost for GTL plant is in the range of $81,950,000.00 up to $114,730,000.00 for production capacity of 1,681 barrels per day (Figure 10). The calculation of TCI in this paper shows that the amount is $97,474,516.89 which is between the TCI costs of GTL by EIA. This shows that the design of the plant is considerable and feasible to be implemented.

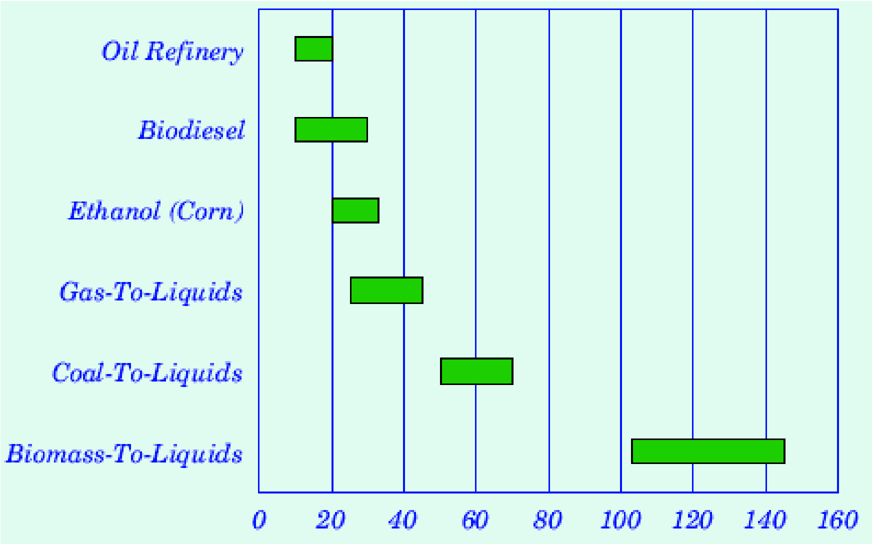

Figure 10. Range of Capital Investment Cost for Synthetic Fuel Plant (1000 USD per barrel per day of capacity)(EIA, Department of Energy USA)

3.3.2. Economic Feasibility Study

This study is done to give a decision whether the process design of GTL plant is feasible to be implemented or not which can give benefits and also to look the pattern of feasibility factor and crude syn-fuel price when the plant is scaled up. This paper prefers to avoid trial-and-error calculations in preliminary process designs, and yet it would like to account for the time value of money in some ways in the profitability and feasibility analysis. To accomplish this goal, it can be defined the Capital Charge Factor (CCF) as (Douglas, 1988):

$$Revenue - Total\ Production\ Cost = CCF\ (TCI\ Cost) \tag{4}$$

From the previous part of this paper, the TCI cost has been defined which is $97,474,516.89. Then, the revenue can be achieved by assuming the selling cost per barrel of crude syn-fuel is the same with the price of Brent crude oil in April 2012 which is $121.80 per barrel. The production cost can be calculated by assuming the price of natural gas is $6.5/MMBTU, electricity is $0.14/kWh, and utility water is $0.15/ton. Those prices are considered by the actual domestic price in Indonesia per April 2012. By considering also maintenance, labor, tax, insurance, interest, and plant overhead, it can result the amount of CCF which only 0.30.

For a new project, a minimum CCF number for the plant to be said potentially feasible is 0.33. Basically, the 0.3 for CCF in this plant design can be said not feasible. But, by doing the scale up projection of the plant, we can get the relation of CCF with production capacity in the following figure.

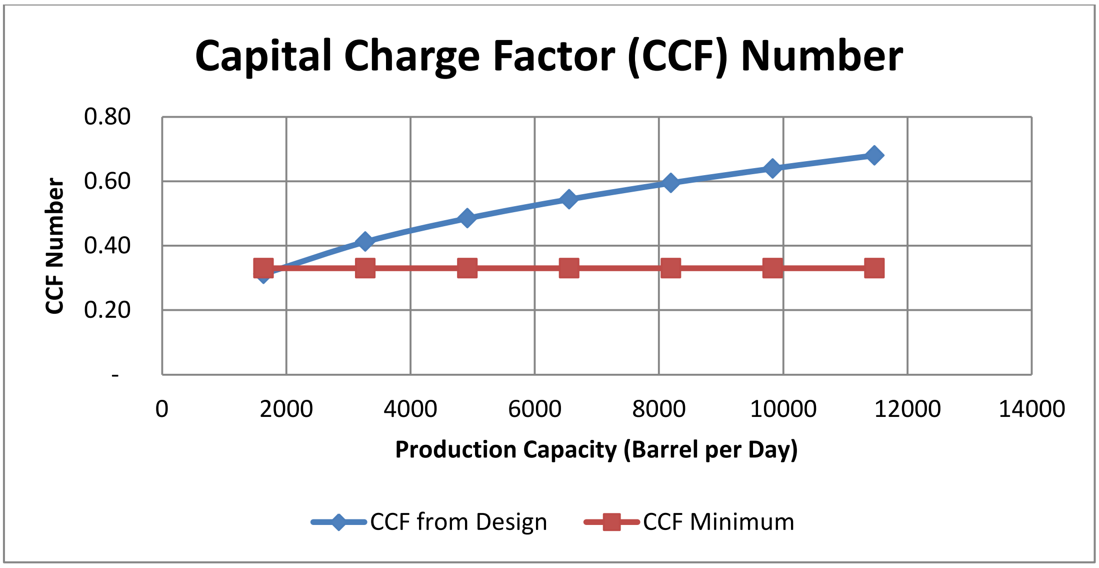

Figure 11. Capital Charge Factor Number

From the figure above, when the production capacity increase, the number of CCF is also increase which makes the plant become very feasible to be implemented because the CCF is bigger than 0.33. While the CCF bigger, the payback period relationship can be achieved by using the following equation (Douglas, 1988), and the result is the payback period become faster when the production capacity bigger (figure 12). From the CCF and payback period graphs, it is shown that the plant design is feasible and even potential to develop when the production capacity is more than 2,000 barrels per day.

$$CCF = \frac{[0.25(1+i)^4 + 0.295i - 0.298](1+i)^N - 0.225i + 0.048}{0.676[(1+i)^N - 1]} \quad (5)$$

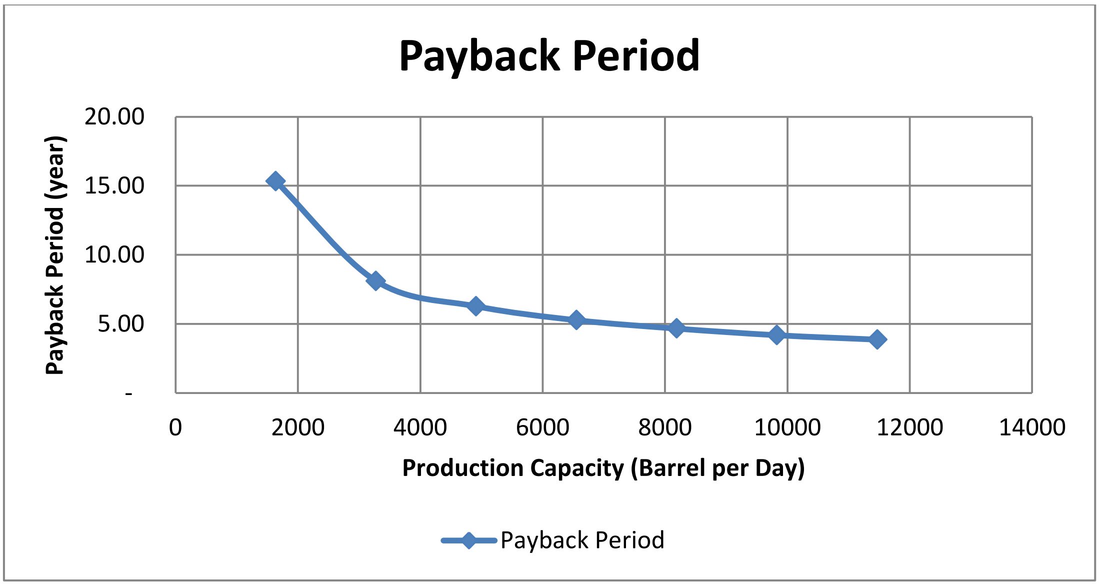

Figure 12. Payback Period per Production Capacity

3.3.3. *Crude Syn-Fuel Price Analysis Study*

The previous section about economic feasibility study shows whether the process design is feasible or not in the condition of selling value of crude syn-fuel is $121.8/barrel based on Brent crude oil price in April 2012. In this section, the price will be decided from minimum CCF when the process design is said to be feasible. The crude syn-fuel price is affected by the revenue. The revenue can be achieved when CCF, production cost, and capital cost are known.

The figure 13 below shows that the increase of production capacity will increase total revenue. When the total revenue increases, the price of syn-fuel will decrease as shown in the figure 14.

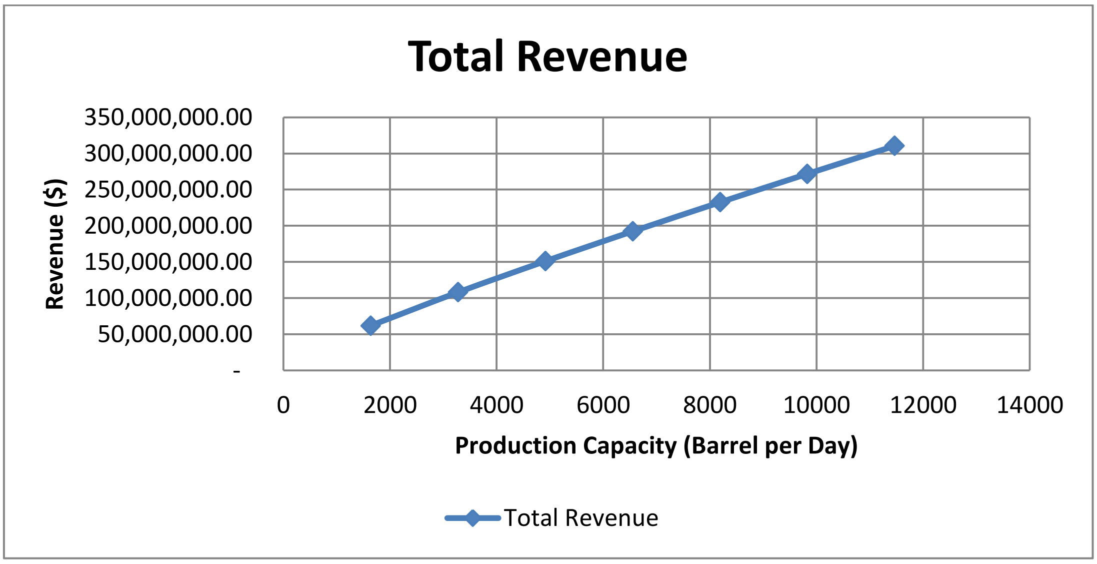

Figure 13. Total Revenue per Production Capacity

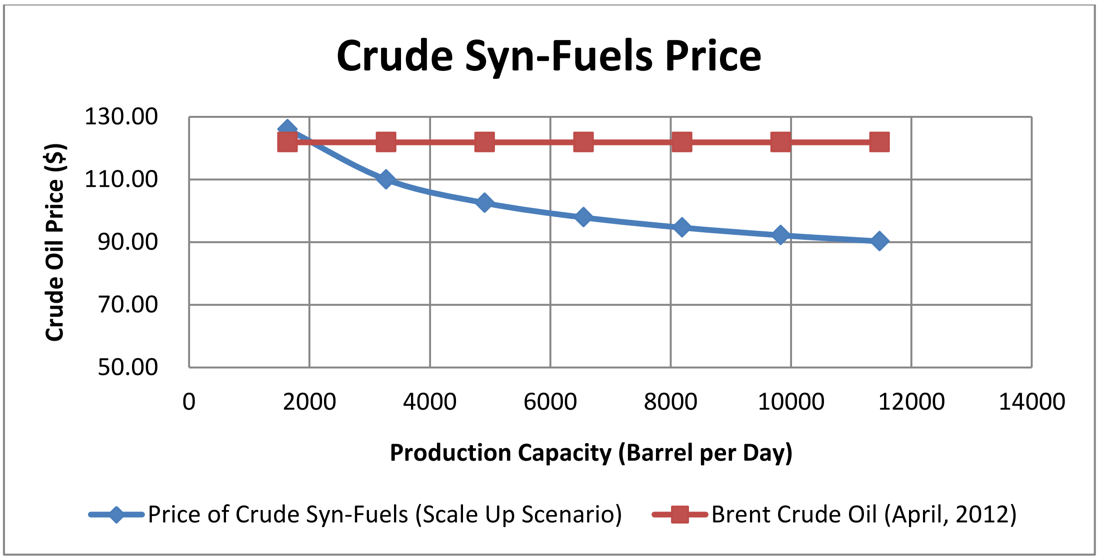

Figure 14. Crude Syn-Fuel per Production Capacity

From the figure 14 above, we can get that the price of crude syn-fuel is cheaper than the price of conventional crude oil based on Brent crude oil price. This is shown that the development of GTL plant can be a solution for Indonesia oil depletion. Besides the price is cheaper than the conventional oil price, the syn-fuel is cleaner because there is no amount of Sulfur which becomes a trouble to the environment.

3.3.4. Sensitivity Analysis

Sensitivity analysis (SA) is the study of how the uncertainty in the output of a model (numerical or otherwise) can be apportioned to different sources of uncertainty in the model input (Saltelli, 2008). In this paper, sensitivity analysis is conducted to show which cost element is the most sensitive to the crude syn-fuel price and the Capital Charge Factor which then decide whether the plant design is feasible or not. The figure below shows how TCI and raw material cost influence the syn-fuel price. From the graph, TCI looks more sensitive than raw material cost to affect the price. When the deviation of both components reaches until 50% and -50%, the TCI influences the price more than 20% while raw material cost only less than 20%. By this information, we can say that the change of TCI and raw material cost will influence the syn-fuel price cost directly. The price of syn-fuel will increase when TCI and raw material cost increases. Meanwhile, it will decrease when both components decrease.

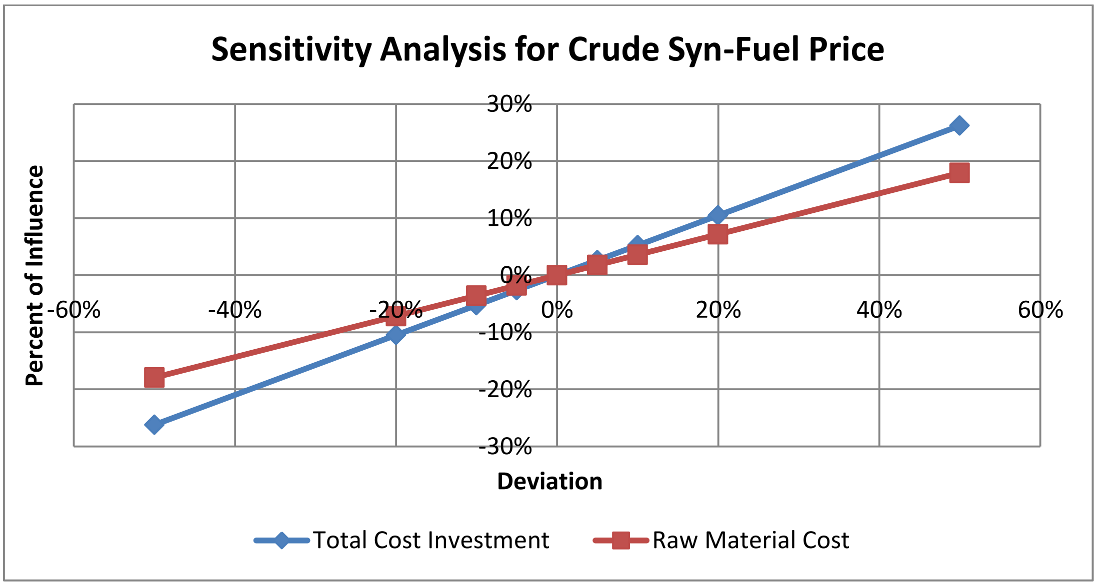

Figure 15. Sensitivity Analysis for Crude Syn-Fuel Price

The following graph shows the influence of TCI, raw material cost, and crude syn-fuel price to the Capital Charge Factor (CCF). From the graph, the CCF will decrease when TCI and raw material cost increase. The value of TCI and production cost will affect the CCF value. When TCI and production cost increase, the CCF will decrease that shows the feasibility of the plant design is low or even unfeasible. Meanwhile, the TCI and production cost decrease, the CCF will increase which also increase the feasibility of the design.

Crude syn-fuel price will also increase the value of CCF. The price will influence the revenue of the plant which supports more benefits for plant. When the revenue increases, the CCF will increase also. When the CCF value becomes higher, the feasibility is also. This shows that the change of syn-fuel price will affect the CCF value.

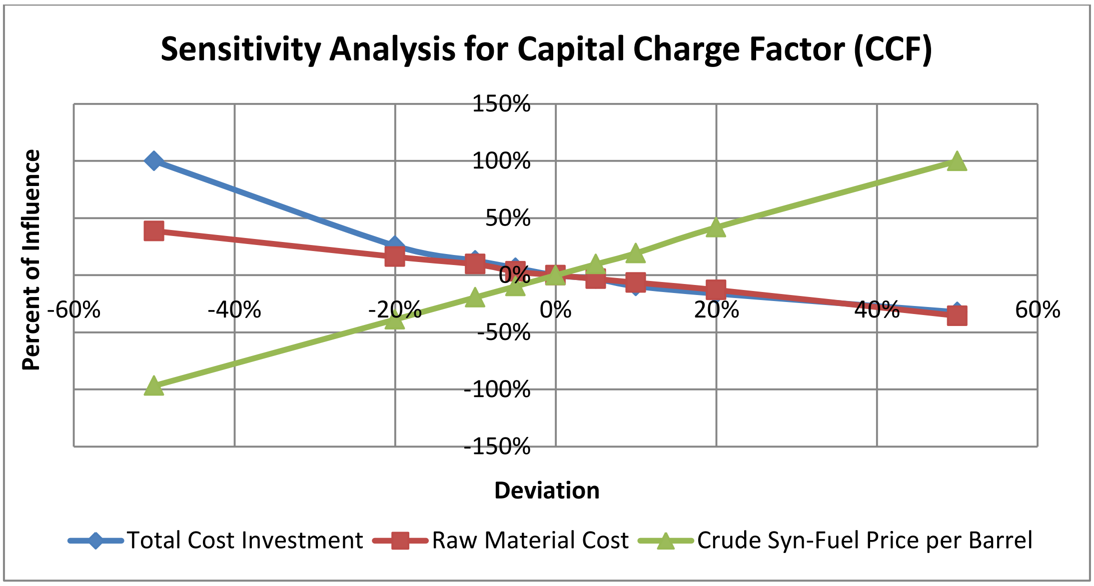

Figure 16. Sensitivity Analysis for CCF

This sensitivity analysis is very useful to see which cost components that have a great influence because of its change. After knowing the most sensitive components, the decision can be made when there are problems, such as problem in syn-fuel price and CCF.

4. Conclusion

Based on the potential of Indonesia's natural gas reserves and also problems in the lack of oil production compared to consumption, one technology that can be implemented to produce Synthetic fuel (synthetic oil) is Gas-to-Liquid (GTL) technology. Gas-to-Liquid (GTL) fuels can be produced from natural gas by using Fischer-Tropsch chemical reaction process. The liquids produces include naphtha, diesel, etc. which is contained also in conventional oil.

From the study, the process design made by simulation results the number of raw materials used and products produced in GTL plant. The methane used as raw materials is about 12.51 MMSCFD (10,000 kg/hr) and the amount of air is followed by the stoichiometric amount which is about 29.77 MMSCFD (42,770 kg/hr). The syn-fuel components produced in the process are simplified into only three major components of fuels such as gasoline (represented by n-Octane), kerosene (represented by Naphtalene), and diesel (represented by n-C16). The total amount of product is about 1,681 barrel per day.

From the preliminary study of the plant development, TCI cost of the plant is $97,474,516.89. This value is considerable based on the projection of TCI cost study by EIA Department of Energy, USA. The feasibility study of the plant stated that the production capacity of 1,681 barrel per day is still not feasible enough to be implemented. But, after doing the scale up projection of the design, the number of CCF is also increase which makes the plant become very feasible to be implemented and will give a lot of benefits from its development. Furthermore, the design results lower crude syn-fuel price than conventional oil price and also cleaner fuel which can be a solution for Indonesia oil depletion.

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Appendix

Appendix 1. Bare Module Cost

Code	**Equipment**	**Unit Ammount**	**FOB/unit ammount ($)**	**Total Module Factor**	**Bare Modul Cost ($)**
Air Separation Unit					
X-100	Filter Bag	1	131,238.50	2.03	266,414.16
K-100	Air Compressor	1	1,512,187.08	3.24	4,899,486.14
K-101	Air Compressor	1	2,073,705.75	3.24	6,718,806.63
K-102	Air Compressor	1	1,720,750.87	3.24	5,575,232.81
E-101	Intercooler	1	22,651.17	3.27	74,069.33
E-102	Intercooler	1	22,650.20	3.27	74,066.16
E-103	Intercooler	1	22,651.33	3.27	74,069.84
X-101	Pressure Swing Adsorption	1	33,445.29	4.20	140,470.24
Methane Preparation					
K-103	Methane Compressor	1	711,188.86	3.24	2,304,251.92
K-104	Methane Compressor	1	764,207.11	3.24	2,476,031.02
K-105	Methane Compressor	1	783,850.73	3.24	2,539,676.36
E-105	Intercooler	1	22,653.39	3.27	74,076.60
E-106	Intercooler	1	22,649.35	3.27	74,063.38
Syn Gas Reformer					
ERV-101	Partial Oxidation Reactor	1	3,131,893.82	2.24	7,015,442.17
E-104	Heat Exchanger	1	22,650.70	3.27	74,067.80
Fischer-Tropsch Process					
CRV-100	FT Reactor	1	2,799,570.95	2.24	6,271,038.92
E-107A	Heat Exchanger	1	22,650.98	3.27	74,068.70
E-107B	Heat Exchanger	1	22,649.25	3.27	74,063.06
V-101	Separator	1	34,319.99	4.20	144,143.97
T-100	Crude Oil Tank	1	64,572.30	1.41	91,046.94
Utilities					
P-100	Water Pump	1	5,789.66	3.47	20,090.13
P-101	Water Pump	1	5,440.37	3.47	18,878.07
P-102	Water Pump	1	5,789.66	3.47	20,090.13
K-106	Steam Turbine	1	1,214,083.41	3.24	3,933,630.26
AC-100	Air Cooler	1	105,964.57	2.46	260,672.85
T-102	Water Tank	1	50,700.48	1.41	71,487.68

Total Bare Modul Cost					**$ 43,359,435.27**

Appendix 2. Total Electricity Needed (Utility)

Code	Equipment	Thermal Energy (kW)	Motor Efficiency		Electricity
Needed					
K-100	Air Compressor	2,280.00	0.95	(Sieder, 2003)	2,400.00
K-101	Air Compressor	3,384.00	0.95	(Sieder, 2003)	3,562.11
K-102	Air Compressor	2,680.00	0.95	(Sieder, 2003)	2,821.05
K-103	Methane Compressor	887.90	0.95	(Sieder, 2003)	934.63
K-104	Methane Compressor	971.60	0.95	(Sieder, 2003)	1,022.74
K-105	Methane Compressor	1,003.00	0.95	(Sieder, 2003)	1,055.79
P-100	Pump	32.18	0.95	(Sieder, 2003)	33.87
P-101	Pump	21.45	0.95	(Sieder, 2003)	22.58
P-102	Pump	3.33	0.95	(Sieder, 2003)	3.51
Produced					
K-106	Steam Turbine	5,719.00	0.65	(Sieder, 2003)	3,764.15
			Total Electricity Needed		**8,092.12**

Appendix 3. Production Cost Calculation

Component	Unit	Quantity per Year	Unit Price ($)	Cost per Year ($)
Direct Production Cost				
Raw Material				
Natural Gas (Methane)	MMBTU	3,414,927.60	6.5	22,197,029.40
Utility				
Electricity	kWh	2,330,530.00	0.14	326,274.20
Cooling Water	ton	3,000.00	0.15	450.00
Maintenance	0.04 Fixed Capital			3,049,993.83
Supply	0.006 Fixed Capital			457,499.07
Labor + Supervisor + Laboratory	1.35 labor Cost			540,000.00
Fixed Charges				
Tax	0.03 Fixed Capital			769,727.08
Insurance				
Rent				
Interest				
Plant Overhead				
Plant Overhead	0.72 Labor + 0.024 Fixed Capital			2,117,996.30
Total				**29,458,969.88**

Appendix 4. Economic Feasibility and Scale Up Projection (Oil Price, CCF, and Payback Period)

	Production Capacity (Barrel per Day)						
	1,681	**3,278**	**4,917**	**6,556**	**8,195**	**9,834**	**11,473**
CCF	0.33	0.33	0.33	0.33	0.33	0.33	0.33
TCI ($)	97,474,516.89	147,743,740.07	188,435,985.90	223,937,634.42	256,019,528.84	285,615,545.55	313,292,440.20
Total Production Cost ($)	29,458,969.88	58,917,939.76	88,376,909.64	117,835,879.52	147,294,849.40	176,753,819.27	206,212,789.15
Revenue ($)	61,917,984.00	108,116,605.20	151,126,092.94	192,407,111.78	232,549,352.50	271,863,795.94	310,539,171.74
Cost per barrel ($)	125.93	109.94	102.45	97.83	94.59	92.15	90.22
Brent Crude Oil (April 14th, 2012) ($)	121.83	121.83	121.83	121.83	121.83	121.83	121.83
	Production Capacity (Barrel per Day)						
	1,681	**3,278**	**4,917**	**6,556**	**8,195**	**9,834**	**11,473**
CCF	0.31	0.41	0.48	0.54	0.59	0.64	0.68
TCI ($)	97,474,516.89	147,743,740.07	188,435,985.90	223,937,634.42	256,019,528.84	285,615,545.55	313,292,440.20
Total Production Cost ($)	29,458,969.88	58,917,939.76	88,376,909.64	117,835,879.52	147,294,849.40	176,753,819.27	206,212,789.15
Revenue ($)	59,903,811.00	119,807,622.00	179,711,433.00	239,615,244.00	299,519,055.00	359,422,866.00	419,326,677.00
Cost per barrel ($) (Based on Brent Crude Oil April 14th, 2012)	121.83	121.83	121.83	121.83	121.83	121.83	121.83
CCF Minimum	0.33	0.33	0.33	0.33	0.33	0.33	0.33
	Production Capacity (Barrel per Day)						
	1,681	**3,278**	**4,917**	**6,556**	**8,195**	**9,834**	**11,473**
CCF	0.31	0.41	0.48	0.54	0.59	0.64	0.68
TCI ($)	97,474,516.89	147,743,740.07	188,435,985.90	223,937,634.42	256,019,528.84	285,615,545.55	313,292,440.20
Total Production Cost ($)	29,458,969.88	58,917,939.76	88,376,909.64	117,835,879.52	147,294,849.40	176,753,819.27	206,212,789.15
Revenue ($)	59,903,811.00	119,807,622.00	179,711,433.00	239,615,244.00	299,519,055.00	359,422,866.00	419,326,677.00
Cost per barrel ($) (Based on Brent Crude Oil April 14th, 2012)	121.83	121.83	121.83	121.83	121.83	121.83	121.83
Payback Period	15.33	8.11	6.27	5.27	4.66	4.18	3.86