REFINERY SCHEME'S MASS TARGETING AND BOTTOM SECTION

SYNTHESIS FOR HEAVY OIL

A Thesis

by

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ABSTRACT

This work evaluates the introduction of heavy oil in a refinery as a first step. The first step will yield an increase in the production of the bottom products (vacuum residue, gas oil and diesel). It will also reduce the production of the light products (gases, LPG and naphtha) from ADU/VDU for oil with API above 20. However, we showed that if the heavy oil is below 20 API, the vacuum residue will be the only increasing product. This also reflects on the unit capital cost. The power and steam required by the refinery should also increase as crude oil becomes heavier due to the high amount of steam used in the delayed coker unit. Nevertheless, the fuel for the fire heaters does not show the expected change as compared to the model.

The report goes to a further step by replacing bottom product processes with gasification and syngas routes. This step results to reduce the total production of fuel. Therefore, the fuel gasification paths MTG, DME (direct and indirect), and FT are more valuable than other gasification paths. All fuel paths showed a similar amount of fuel production, yielding extra production around 100,000 lb/hr compared to the base case. Moreover, the direct path of DME provided the lowest estimated cost compared to other fuel gasification paths. The MTG path and indirect DME path have a similar cost.

The final step is to investigate two challenges related to the gasification cases: water balance and fuel demand. The investigation shows that more than 95% of used water can be recovered by recycling water (both direct and indirect recycling).

Furthermore, the study shows that MTG and DME-indirect paths demand less fuel when compared to the base case.

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NOMENCLATURE

DC	Delayed coker unit
BTX	Benzene, toluene, and the three xylene isomers
ALK	Alkylation
API	American Petroleum Institute gravity
AGO	Atmospheric gas oil
ADU	Atmospheric distillation unit
ATK	Aviation turbine kerosene
CCR	Catalytic reforming
DC	Delayed coker unit
DME	Dimethyl ether
FT	Fischer Tropsch
FCC	Fluidized catalytic cracking
HGO	Heavy gas oil
HSR	Heavy naphtha
VGO	Heavy vacuum gas oil
H-C	Hydrocarbone
HC	Hydrocracking
H/C	Hydrogen to carbon ratio
ISO	Isomerization
LGO	Light gas oil (diesel)

LSR	Light naphtha
LP	Low pressure
MP	Medium pressure
MeOH	Methanol
MTG	Methanol to Gasoline
ΔT	Tempertuer different
VDU	Vacuum distillation unit
VR	Vacuum residue

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CHAPTER I

INTRODUCTION AND LITERATURE REVIEW

1.1 Introduction

The modern refinery, the petroleum distillation, came on stream in 1862 when Benjamin Jr.'s research showed that 50% of the new Pennsylvania Rock Oil can be distillates into first-rate burning oil, kerosene and paraffin oil and, while the other 40% can be used for another purpose such as lubrication. In the nineteenth century, mainly in the United States and Russia oil production have led to many developments in oil extraction. Then, in the twentieth century, different regions in the world started to produce oil, such as Indonesia, Mexico and Middle Eastern countries. Since then, the oil industry has grown by more technological research and developments in improving their processes and reducing the energy used (Speight, 2011).

The Petroleum refinery industry has different terms in classifying oil types. The crude oil can be solely segregated based on its chemical and physical characteristics, such as components, gravity and sulfur content. The standards and most common systems classification follow the following:

1. Conventional crude: It is the oil produced using conventional method and mainly produced from oil wells. . Also, its refinery is expected to be normal one, hence, it does not have hard operating conditions such as sulfur content. It consists of a mixture of hydrocarbon compounds: gaseous, liquid, and solid.

2. Opportunity crude (challenging crude or unconventional crude): It is the same as the conventional crude, but they are very difficult to process due to: high total acid number, high sulfur content and/or low API gravity.
3. Heavy oil: It is a class of opportunity crude with higher viscosity and lower API gravity (range of 10 to 15) than conventional petroleum. This type requires high thermal reservoir to process. Its sulfur content is expected to be higher than 2 wt%.
4. Tar sand (oil sand): This is also considered as a type of unconventional crude. It is a heavy black viscous-oil mixed with clay, sand and water. Therefore, it usually has an extremely high viscosity. The composition and construction of the sand play a vital role in the recovery process.
5. Biomass (bio-feedstocks): Usually refers to a mixture of carbon based organic compounds, carbon base, compounds which are used as fuel or feed to hydrocarbon production processes. There are four main resources of it: In contrast to the previously discussed types, this class originates from agricultural crops, wood, municipal and industrial wastes and landfill waste.

Historical evidence shows that human beings used oil in heating and lighting 4000 years ago (Speight, 2011). Nowadays, it still serves as the as the primary source of energy with many forecasting further demands to support the growing economy in many countries. According to the U.S. Energy Information Administration (EIA), The US demand for oil will continue increasing linearly till 2040 as Figure 1 shows their forecasting production and demand (Annual Energy Outlook 2015, 2015).

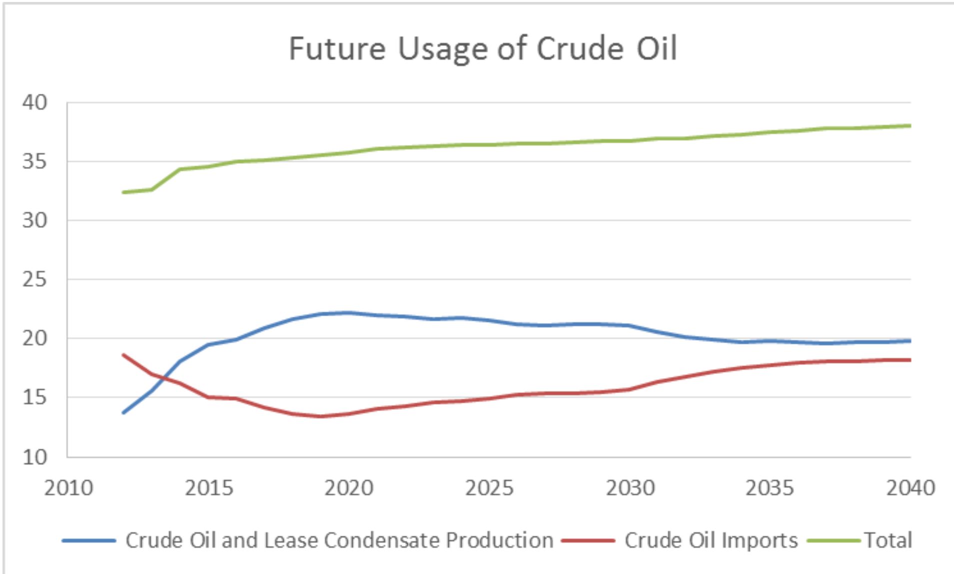

Figure 1: Future Usage of Crude Oil for the US **(Annual Energy Outlook 2015, 2015)**

Current crude oil differs in composition and properties when compared with the crude produced 50 years ago as we observed a decrease in API gravity and an increase in sulfur content in general (Speight, 2011). Such fluctuation in properties compelled the refinery industry to upgrade their processes in order to produce the expected energy and liquid fuel. Consequently, an overview of the refinery processes and with the expected changes is critical. There are three pathways applied in the industry, which utilize heavy crude oil and supply it to the commercial market:

- Oil field upgrading: the changes are done in the oil field itself within either within the borehole, the production equipment at the surface or the subsurface. . Such techniques will split the heavy oil into refined products and coke. The coke is collected in the open pit mines where it can be used later as tar sands. On the other hand, refinery products are distilled and recombined to produce clean synthetic crude (Speight, 2011).

- Refinery upgrading: the modifications are done in the refinery complex only, especially bottom section. These upgrades require sufficient amount of energy (gas and electric resources) (Speight, 2011) which should be considered before marching into such upgrades.
- Feed upgrading (mixing): the changes are done at the feed level where the opportunity crude, such as heavy oil, gets mixed with conventional crude to dilute its severe conditions. However, the ungrudging bound to refinery complex limits such as sulfur contact (Speight, 2011).

The report focus on the effects of the crude oil’s API change at the refinery schemes at mass and utility levels only without any consideration to other upgrading. It is a key stage to study for either the existing refinery or the new ones to aid in assisting the refinery limitations. In addition, this research will address the implementations of gasification synthesis at the bottom section of the refinery. This step paves the way for a new approach where we integrate alternative energy systems for a better solution. The new approach changes the refinery gases, refinery heavy products, biomass and coal from to power, liquid fuel and chemical products.

First, a literature review of the future refinery processes will be presented. Also part of the literature review will be about the new concept and gasification synthesis. After that, the problem statement and evaluation methodologies of the report will be presented in detailed. Finally, results and discussions of the potential scenarios will be addressed.

1.2 Literature Review

1.2.1 Refinery Processes

It will not be scientifically accurate to overlook at the refinery chemistry without describing the refinery processes itself. The chemistry is part of the refinery's conversion section, which will get affected by upgrading. The conversion section's purpose is to enhance the liquid fuels. There are two types of techniques under the conversion section:

- Hydrogen addition technologies:

Includes all processes that utilize the techniques of reacting hydrogen with the oil heavy products. Hydrogen addition technologies aim to improve the final yield of the light products via enhancing the H/C. Since, it is a catalyst base process; then, it is expected to yield more products, but unfortunately requires a large investment since it needs hydrogen. Its price range from $2.8 to 3.8 per kg-$H_2$ depend on the process used for production (Parks, 2011) (Genovese, 2009). In fact, it is a sensitive process since the feed's impurities disturb the catalyst. Therefore, most refineries prefer alternative technology to hydrogen addition. Nevertheless, the hydrogen addition technology applicable showed good potential for small application (Gary & Handwerk, 2007).

- Carbon rejection technologies:

This technology heats the feed under a high pressure or with a catalyst until it reaches its thermal fraction point which will fracture the feed into lower molecular size products. However, it produces more heavy products than the feed which is undesirable disadvantage. Moreover, those techniques have the advantages to be used in processes since not all of them catalyst base, hence easy to operate and control, and not limited to one type of feed (Gary & Handwerk, 2007).

We may conclude that the development of refinery processes is derived mainly on the demand for products, feedstock changing, feedback type, environmental regulations and Technological development. Those driving factors could help increasing the refinery flexibility, which, lead to different operational scenarios. This indicated that the modern refinery operation depends on the type of feedstocks and configuration. There are four categories presented in a modern refinery: separation, treatment, conversion and formulation and blending (Figure 2) (Fahim, Alsahhaf , & Elkilani , 2010).

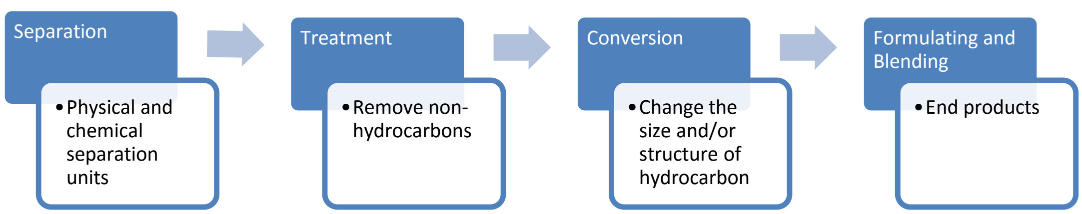

Figure 2: Refinery Unit Types

The following are the major refinery processes that effects that hydrocarbon production:

1.2.1.1 Atmospheric distillation unit and vacuum distillation unit (ADU/VDU)

The main purpose of the distillation section is to separate crude oil into different components. In the atmospheric distillation, the separation bases on the boiling points for liquid components while the separation occurs in vacuum distillation at very low pressure flashing points. The two units can accommodate different types of feedstocks (light and heavy oil) with operation adjusting. Therefore, it is the most flexible unit in a refinery. Likewise, a new path is investigated to the oil via applying catalytic distillation or reactive distillation. The reactive distillation is a distillation column that consists of two sections: the first section uses a distillation to separate the light components from crude. The second section employs chemical reactors to break the heavy components. Under such break up energy saving could be achieved via avoiding further heating to the second section (Fahim, Alsahhaf , & Elkilani , 2010). Yet, another common path is to utilize the membranes for separation since it has low energy usage and low capital cost. However, it requires to introduce a new membrane with a high selectivity. For the next 20 years, the ADU/VDU will remain as the first fractionation stage for the crude utilization. In fact, most scientific and industrial scholars anticipate improvement in such techniques, especially in the heat recovery and distillation units (Speight, 2011).

1.2.1.2 Thermal cracking

The thermal cracking is a carbon rejection process where it cracks or thermally decomposes heavy oil streams into lower molecular weight, lower boiling components, but heavier products are produced. Frequently, the feed is the bottom product of the vacuum column (Fahim, Alsahhaf , & Elkilani , 2010). The main use of it is to prepare the heavy product to be utilized more in the down processes, such as catalytic cracking. There are three widely thermal cracking methods that are commonly used in modern refinery as the following:

- Mild cracking (visbreaking): the unit partially heats the vacuum residues. It is main goal is achieved via reducing the vacuum residues's viscosity. Therefore, this step is highly associated with the production of some light products.
- Delayed coking: the unit totally cracks the vacuum residues to produce maximum lighter products. However, it results in more coke production,
- Flexi Coking: the unit cracks the vacuum residues similarly to the delayed coking, but it employs a steam to gasify most of the coke.

The main characteristics and the different between above mentioned three processes are illustrated in (Figure 3). Those processes are widely considered as key components of the future refinery as they take care of heavy product from ADU/VDU. Moreover, they are relatively simple. In the future, they will become more efficient where most of the changes will be done in the heater internals and introduce catalysts in the heater (Speight, 2011).

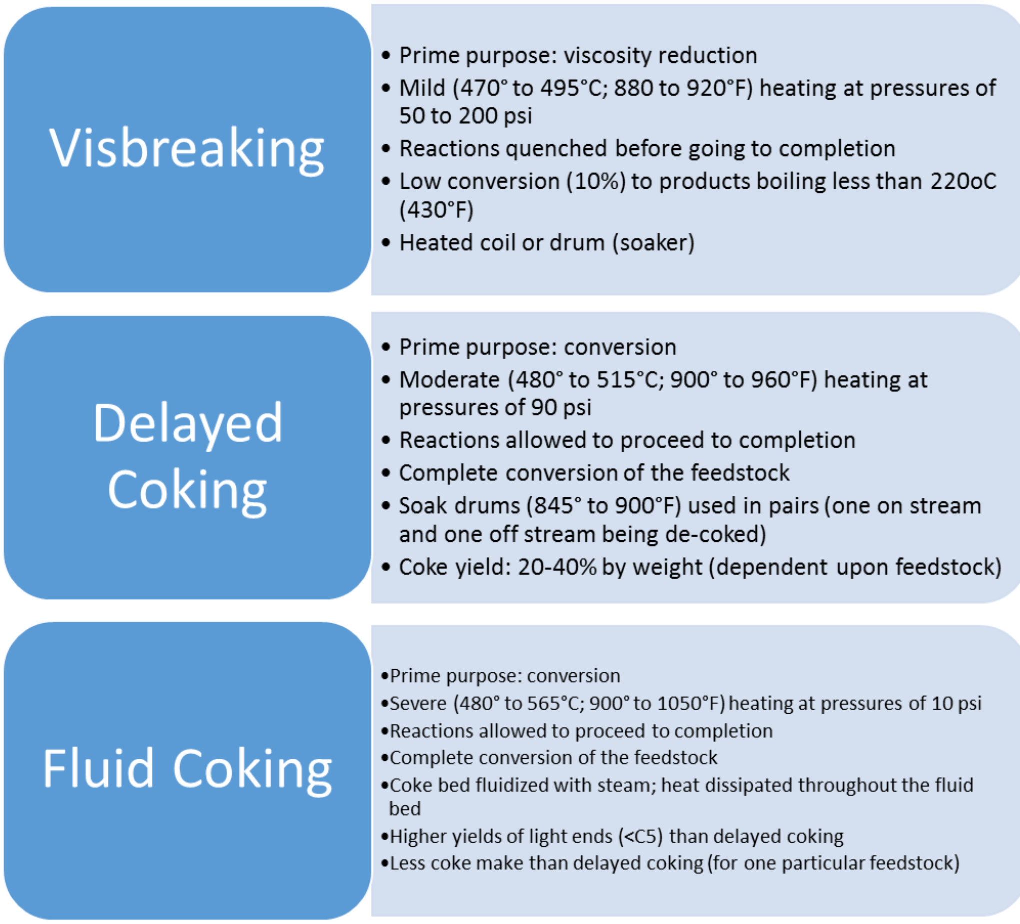

Figure 3: The Main Characteristics of the Thermal Cracking Types **(Speight, 2011)**

1.2.1.3 Catalytic hydrocracking

Hydrocracking is a thermal and hydrogenation process. It is a modern process. It is considered a major refinery upgrading in the last century since it enhances the low quality of the products to more valuable products such as gasoline and jet fuel through using hydrogen. Its primary advantage is that it accommodate different types of feeds. Therefore, the feedstocks are mostly middle distillates, residual fuel oils, and reduced crude. There are two series, set of chemical reactions occurring in the process: First reactions set involve heavy hydrocarbons. They are cracked to lighter unsaturated hydrocarbons by the catalytic cracking. Those reactions are endothermal. Thus, the feed is cooled as the reactions happen. The Zeolites catalyst is used (Gary & Handwerk, 2007). On the other hand, the second reactions set mainly focus on the unsaturated hydrocarbons. Those are usually saturated by using hydrogen. These reactions are exothermal which could cause excessive heating of the system. Therefore, hydrogen is used as a reactor temperature controller (Gary & Handwerk, 2007).

It should be considered as part of the utopian refinery since it is a source of high quality naphtha, jet fuel, diesel, and lube base oil. Moreover, the hydrocracking will an important part between the integration of refining and petrochemical plants because of its flexibility (Speight, 2011).

1.2.1.4 Catalytic cracking

The Catalytic cracking is a carbon rejection process that uses the catalyst and heat. It is exactly considered a thermal cracking except it applies catalyst while requiring less heat input for the cracking. The inequality between the thermal cracking and Catalytic cracking are elaborated in Figure 4. It edges over thermal cracking as it yields higher octane gasoline (Gary & Handwerk, 2007). The fluid catalytic cracking, fluidized bed reactor, is the commonly used process in refinery complex (Speight, 2011).

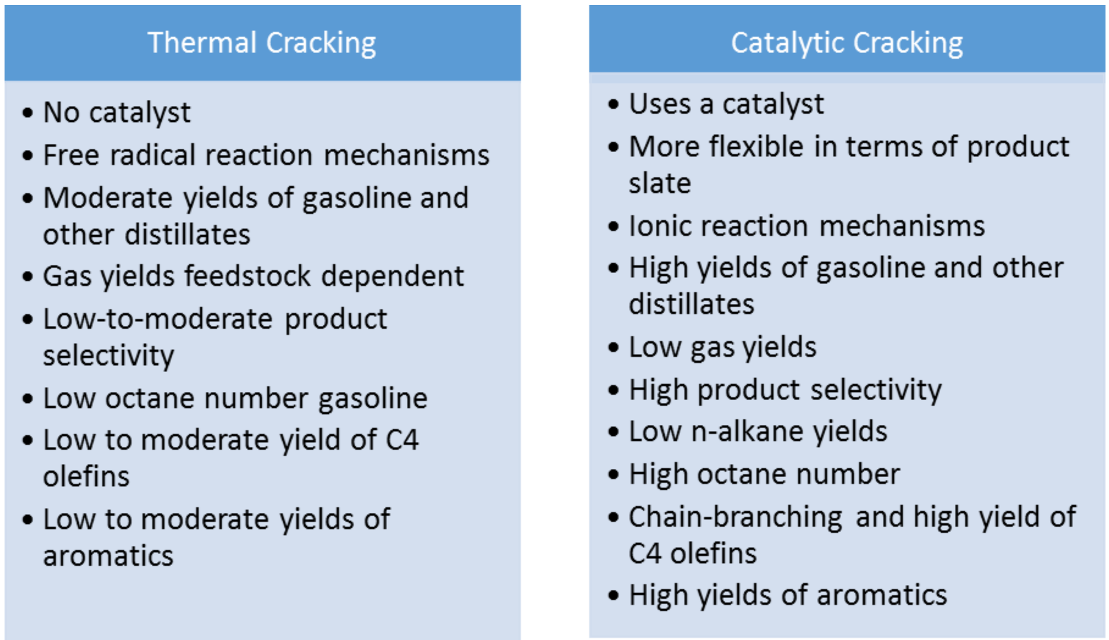

Figure 4: The Differences between the Thermal Cracking and Catalytic Cracking **(Speight, 2011)**

1.2.1.5 CCR reforming

It is a molecular change process where it used to enhance products, low octane gasolines and naphtha, to high octane aromatics. Four major reactions happen in the process: dehydrogenation, which is the main reaction (naphthenes to aromatics), dehydrocyclization (paraffins to aromatics), isomerization and hydrocracking. This process is the solely process in refinery complex that produce hydrogen as a byproduct (Fahim, Alsahhaf , & Elkilani , 2010).

1.2.2 Integrated Alternative Energy Systems

It is a new concept that can substitute the alternative energy sources concept, the conventional approach, as a potential future solution for the global energy system. The concept focused to produce a flexible system with multi-feeds on and products out. The conventional approach tried to replace crude oil system by coal and biomass system, new products. Despite the fact that the results of such technique are quite beneficial, yet; it is a partial solution since the energy system contemporaneously reaches a stage where crude oil is being delivered to different energy services and feedstocks. The transportation and infrastructure of the oil industry are another parameters that drive the conventional approach equation. Addressing these challenges, the new concept was created. The new concept does not exclude the old system, rather it integrates its fundamentals and accumulate on it via introducing new feedstocks and processes. This is achieved by building in new advanced technologies to the existing energy systems to produce more various combinations of energy and chemicals. The syngas is pointed as the key solution since it's capable to deal with coal, heavy oil and biomass to produce multi-fuel and product strategies. The future refinery is expected to have the gasification section being able to handle natural gas, heavy product, biomass and coal. The syngas will be used to produce electricity, hydrogen, and oil products through Fisher-Tropsch (FT) synthesis and power generations (Figure 5). (Szklo, & Schaeffe, 2015).

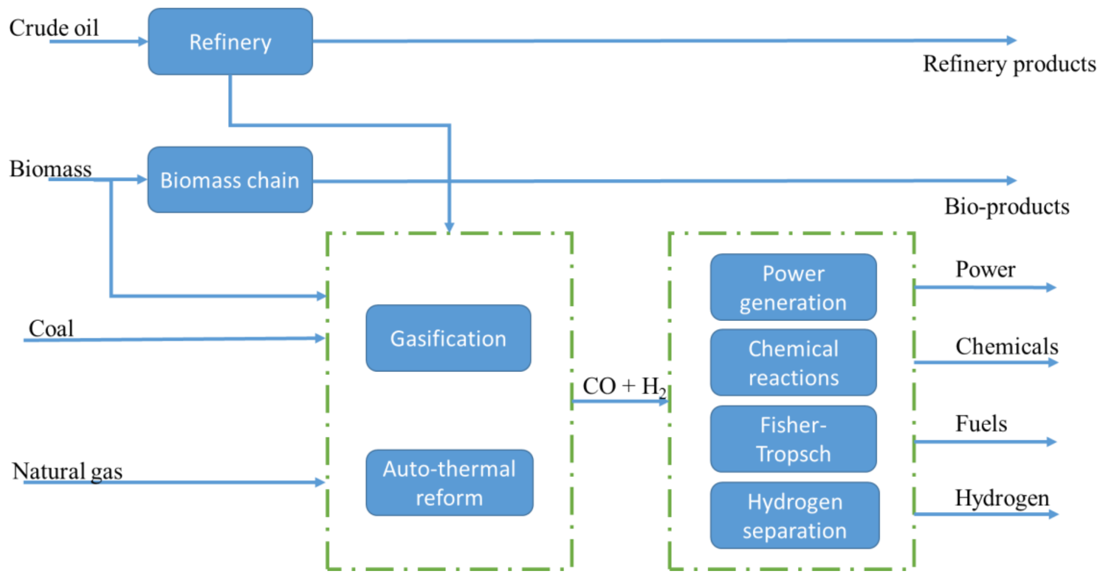

Figure 5: Integrated Alternative Energy Systems

1.2.3 Gasification and Syngas

Syngas is a gas mixture of hydrogen and carbon monoxide, which are natural products from gasifier any hydrocarbon material such as coal at a high pressure and temperature, 30 bar and 1500 °K. The presence of oxygen is the primary difference between gasification and combustion. When a feed is gasified, the oxygen and air is injected gradually and carefully in small amount which allow part of the feed to burn completely. Therefore, the required heat and pressure will be supplied which lead to break down the feed into synthesis gas (Higman & Burgt, 2003). This process consists of six main reactions described briefly in Table 1:

Table 1: Gasification Main Reactions

Reaction Process	Chemical Formula	Change in Enthalpy
Gasification with Oxygen	$C + \frac{1}{2} O_2 \rightarrow CO$	–3,922 Btu/lb C
Combustion with Oxygen	$C + O_2 \rightarrow CO_2$	–14,111 Btu/lb C
Gasification with Carbon Dioxide	$C + CO_2 \rightarrow 2\ CO$	6,267 Btu/lb C
Gasification with Steam	$C + H_2O \rightarrow CO + H_2$	4,750 Btu/lb C
Gasification with Hydrogen	$C + 2\ H_2 \rightarrow CH_4$	–2,672 Btu/lb C
Water Gas Shift	$CO + H_2O \rightarrow CO_2 + H_2$	–650 Btu/lb CO

There are four types of configurations for the gasification. Those configurations describe how oxygen should be introduced to hydrocarbon feed. Different configurations will produce different characteristics (Breault, 2015). A detailed explanation for each configuration is described in Table 2.

Table 2: The Four Basic Configurations for the Gasification

Bed Type	Hydro Dynamically
Moving Bed	Gas flow through a porous medium where the solids move down while gas moves up - a counter current fashion
Fluidized Bed	Bubbles of excess gas induced and promote mixing.
Entrained Flow	Both solids and gas move in a co-current manner, either up flow or down flow.
Transport Flow	Both gas and solids are fed to the reactor that is circulating fluidized beds.

The syngas is considered as a base for C1 chemistry. C1 chemistry deals with building products from materials that have one carbon atom per molecule. In fact, the C1 chemistry provides a variety of technologies that produce chemicals, clean fuel and power from the syngas (Keim, 1986). According to GTC World Gasification Database, there are currently 272 operated plants and 74 plants under construction. In 2014, they produced 115 MWth of synthesis gas. In fact, it is forecasted to reach more than 290 MWth by 2019 (What is Gasification?, 2015).

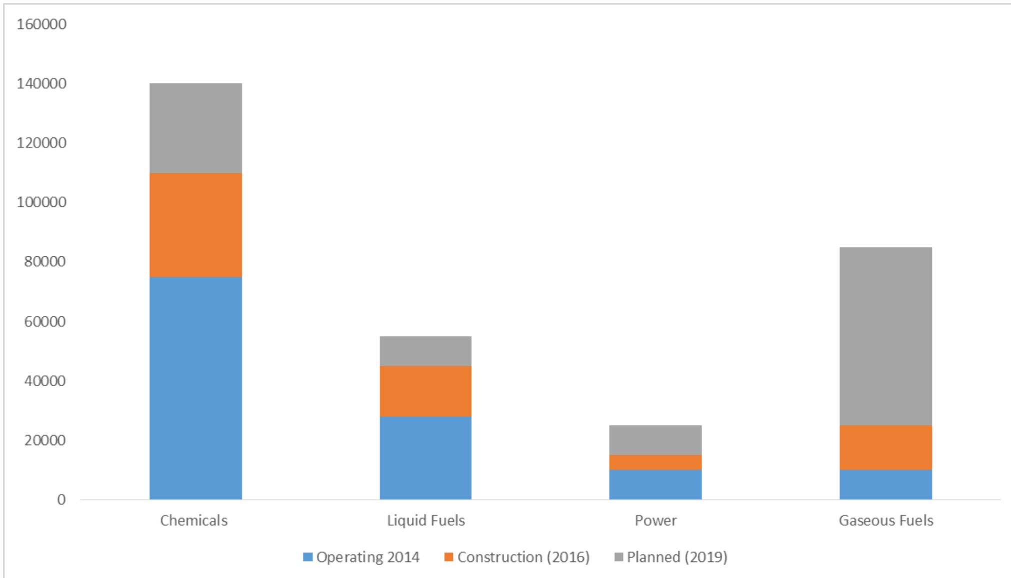

Figure 6: Gasification Plants (Operated, Construction and Planned) **(What is Gasification?, 2015)**

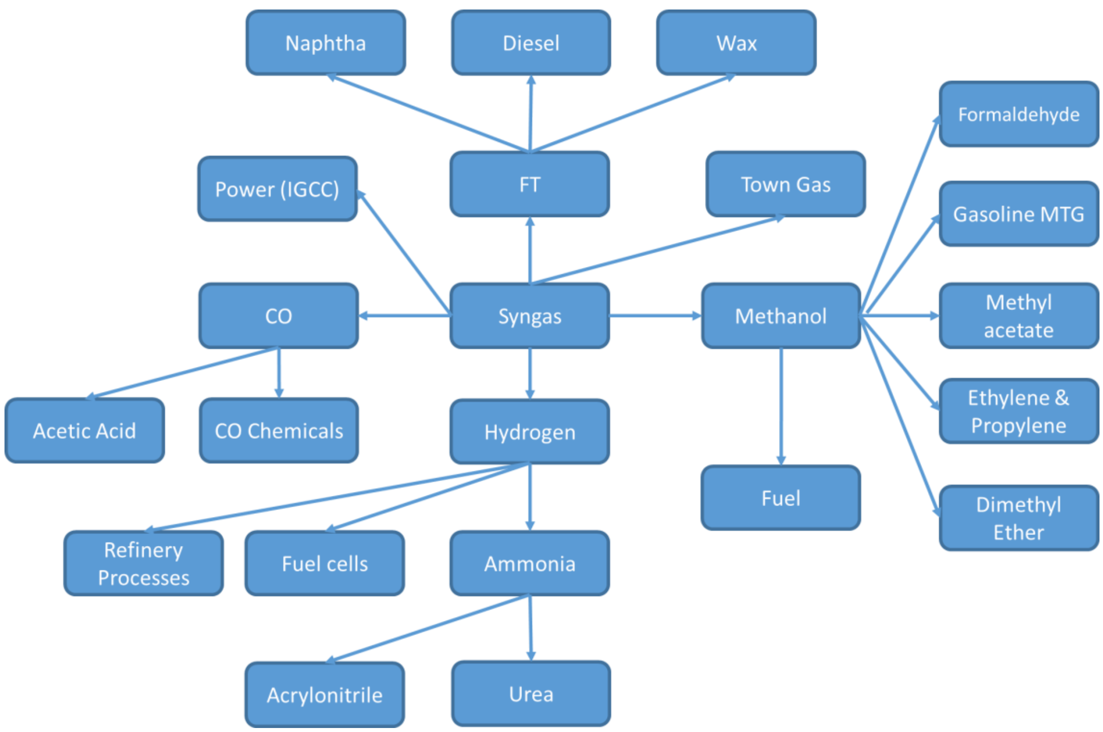

Figure 7: Coal to Syngas Routes **(Chemicals | netl.doe.gov, 2015)**

There are several applications for syngas. Figure 7 shows the different application for syngas routes. The following are the major syngas applications:

1.2.3.1 Power generation

Syngas can be fed to modify gas turbines to produce electricity and heat. Integrated gasification combine cycle is widely used within power generation. The US government pushed adaption for this technology since it due to its low carbon footprint (Higman & Burgt, 2003). Hence, the application of using syngas in power generation falls under the clean power initiative of 2015 (Clean Coal Power Initiative, 2015).

1.2.3.2 Hydrogen

Pure hydrogen is a very practical product for refineries since they are one of the largest H_2 consumers. In like manner, Hydrogen reacts with nitrogen in the air to produce ammonia in the Haber-Bosch process (Higman & Burgt, 2003). Due to the simplicity of the process, it became quite popular and highly adapted. For example, 27% of global ammonia production came from coal gasification in 2008 (Fertilizer Commercial Technologies, 2015).

1.2.3.3 Chemicals

Syngas plays a vital role in for the manufacturing of important industrial products such as methanol. Most the produced chemicals are being utilized as feeds to other processes to extract more products. It has been and still the most important application of syngas (Figure 6) (Higman & Burgt, 2003).

1.2.3.4 Liquid fuel

Using the Fischer-Tropsch process or MTG, the syngas can be transferred to liquid fuel, chiefly gasoline and diesel fuel. These processes are a catalytic process (Higman & Burgt, 2003).

1.2.4 Syngas to BTX

Syngas is an extremely flexible product since it can be used to produce major products such as methanol, dimethyl ether and propylene. These products have a wide range of potential applications. This report examines the production of Benzene, Toluene, and Xylene which all often referred as (BTX) from syngas. BTX is used to produce polymers and numerous consumer petrochemicals (Figure 8) (Pellegrino & Energetics, 2000).

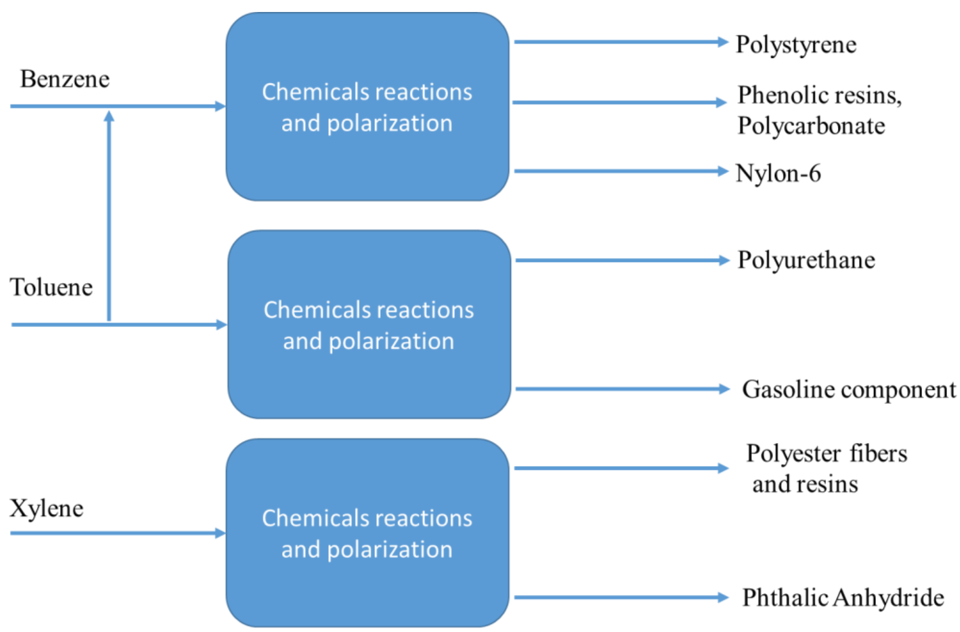

Figure 8: The BTX Chain **(Pellegrino & Energetics, 2000)**

There are three major paths were investigated to produce the BTX from syngas:

1.2.4.1. Synthesis of aromatics from syngas via dimethyl ether (DEM) and methanol

The conversion will take place in two reactors. First reactor converts the syngas to DME and methanol using the Cu/Zn/Al2O3 catalyst. The second reactor employs the modified HZSM-5 zeolite to convert DME and methanol to aromatics The conversion rate of this system ranges between 70% to 83%CO (Zhang, Tan, Yang, Xie, & Han, 2013).

Here, we can easily argue that the conversion methanol to aromatic (BTX) is quite superior as the methanol conversion approaches maximum at a very high rate close to 100%. It will be similar to the previous system, two reactors in parallel. The catalyst consists from ZSM5 zeolite (85 to 99 %), M1 (0.1 to 15 % of Ag, Zn and/or Ga) and M2 (0 to 5% of Mo, Cu, La, P, Ce and/or Co) (United States of America Patent No. US 20140018592 A1, 2014).

1.2.4.2. Direct conversion of syngas to aromatics using Fisher Tropsch and zeolite

Syngas can convert to aromatics using Fisher Tropsch (FT) and zeolite catalysts. The Fisher Tropsch catalyst uses to change the CO and H_2 to hydrocarbon, then zeolite catalyst is used to obtain high selectivity of aromatics. This procedure could be done via using either by a single reactor with a combination of the two catalysts or by two reactors in series. It is worth mentioning that if the reactors were connected in series, then, the first reactor should have the FT catalyst while the other should have the zeolite catalyst the expected conversion to around 49.5% (Corsaro, Wiltowski, Juchelkova, & Honus, 2014).

1.2.4.3. Conversion of methane to aromatics

This process benefit from a metal ion ZSM-5 zeolites catalyst. This approach was an industry investigation since methane is the main component of gases and solid fossil. Obviously, research or inventions of any direct conversion to chemicals will open a new door of industry applications. This process is shorter compared to other processes as it takes advantage of the low conversation of methane to BTX since maximum conversion is less than 8 % CO (Weckhuysen, Wang, Rosynek, & Lunsford, 1998).

CHAPTER II

PROBLEM STATEMENT

The crude oil will continue to be the primary source of energy that fuel the modern world economy. It is expected that produced oil from wells and mature fields will be heavier with time. A proposed solution, out of many, is to accommodate the new heavier oil in the refinery scheme since it is the paths for the integrated alternative energy systems.

This work will be examined refinery system and bottom products syntheses. It will be executed in three stages:

The first stage would study the effects as a function of changing the crude oil properties. Oil API will be decreasing at the moderate refinery schemes. The function will vary the API at four levels: extra light (API above 40), light oil (API between 30 and 40), medium oil (API between 20 and 30) and heavy oil (API blow 20). This stage will provide us with an insight about the mass profile of the refinery as well as the change in the utilities. The result will be presented in the following areas: ADU/VDU products, feed for refinery units, refinery final products, total utilities required and cost.

The second stage would address the refinery bottom section syntheses for heavy oil. Six scenarios will be generated and compared against each other based on the type of products and energy consummation: Delayed Coker (base case), Gasification to H-C (MTG, DME- direct and indirect), Gasification to Chemicals Production (BTX), and

Gasification to FT. This research is expected to contribute new approaches in utilizing the refinery gases while accounting the energy required by the refinery.

The third and final stage will investigate two major challenges related to the gasification cases. Here, we will only consider water balance and energy balance.

CHAPTER III

METHODOLOGY

3.1 First Stage Methodology

For the first stage, eight samples of crude oil were used to evaluate our proposed method. The oil samples were classified into four major groups based on their oil API. Hence, we ended up testing the following groups (below 20, from 20 to 30, from 30 to 40 and above 40). The nine feeds were taking from the Statoil ASA, a Norwegian multinational oil and gas company (Figure 9). In each group, two crude oil samples were compared against each other to evaluate the effect of decreasing crude API. Microsoft Excel was utilized to generate a robust sheet that allowed us to model and design the moderate refinery (Crude Oil Assays, 2014). Table 3 shows the crude oil samples, groups and properties.

Table 3: Statoil Crude Oils with Their API (SG)

API Below 20	
GRANE 2014 12 20.31 (0.932)	PEREGRINO 2012 13.42 (0.976)
API 20 to 30	
GIRASSOL200906 29.81 (0.877)	PAZFLOR 2012 02 25.27 (0.903)
API 30 to 40	
OSEBERG 2013 08 38.48 (0.832)	KISSANJE201106 30.81 (0.872)
API Above 40	
Ormen Lange 2010 05 60.36 (0.738)	GUDRUN BLEND 2014 11 50.73 (0.776)

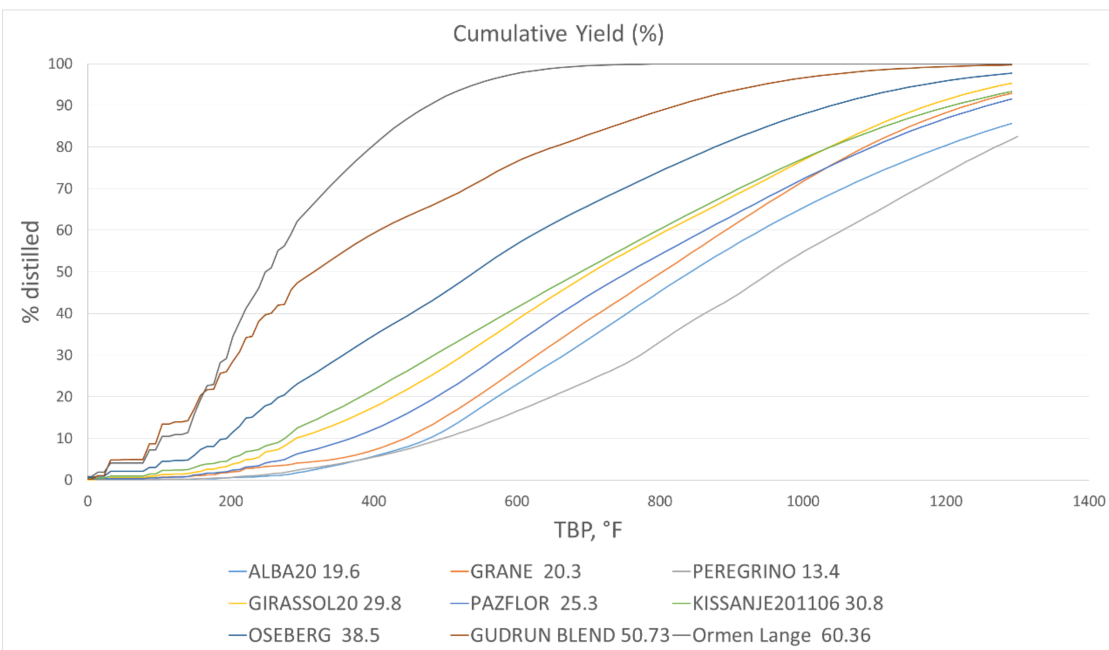

Figure 9: The Yield Distribution of the Nine Crude Oils

Refinery complex is heavily dependent on the nature of crude oil itself and the specifications of the expected products. The refinery processes can be divided into four sections: separation unit (Atmospheric Distillation Unit, Vacuum Distillation Unit and separation columns), chemical units (where chemical reaction takes place to produce valuable products like Delayed Coker and Catalytic reforming Unit), cleaning units (intended to remove unwanted materials like sulfur, salts and metals if any) and production blending units (Speight, 2011). Since we are trying to study the mass balance, the following units are considered the major contributor units:

3.1.1 Crude Distillation Unit

Distillation unit separates the crude oil into products by fractionating based on the products boiling point. In most refineries, the unit consists of two distillation column: atmospheric distillation column (from 30 to 42 trays) and vacuum distillation column. The combination of the previous mentioned two columns is not only considered most optimized as it is also cost effective. The crude is heated to 300 °F prior entering the Desalter. Then, the crude oil is heated again to reach 750°F before it enters to the atmospheric distillation column. The column fractions the crude oil into the gasses (butane and lighter), naphtha (LSR and HSR), kerosene, diesel (LGO), gas oil (HGO) and residue oil (the bottom product). The residue is fractioned more in the vacuum distillation column where the column operates at very low pressure 13.2 psi (high vacuum). The products come from the second column are gases, atmospheric vacuum gas oil (AGO), heavy vacuum gas oil (VGO) and vacuum residue (VR).

There are side columns (from four to 10 trays) which are used to strip the products from the light ends. High portion of heating crude oil is mostly done by using the columns'' products and pumparounds (Fahim, Alsahhaf , & Elkilani , 2010). It is worth noting that steam is used in the unit to strip out light ends from the bottom section of both columns (Fahim, Alsahhaf , & Elkilani , 2010). Moreover, steam is used in the unit to strip out light ends from the bottom section of both columns.

Using the true boiling point (TBP) date of crude oil, the products are defined. From product end boiling point (EP) cut, the volume percent is calculated as the cumulative volume present and then the vacuum residue is determined as the difference between the total and the sum of the calculated products. Table 4 shows the end point for the products (Fahim, Alsahhaf , & Elkilani , 2010):

Table 4: ADU/VDU Products Cut

Product	Initial boiling point (°F)	Endpoint (°F)	Processing use
Gasses	-	90	
LSR gasoline cut	90	180	Min. light gasoline
	90	190	Normal LSR cut
	80	220	Max. LSR cut
HSR gasoline (naphtha)	180	380	Max. reforming cut
	190	330	Max. jet fuel cut
	220	330	Min. reforming cut
Kerosine	330	520	Max. kerosine cut
	330	480	Max. jet-50 cut
	380	520	Max. gasoline
Light gas oil	420	610	Max. diesel fuel
	480	610	Max. jet fuel
	520	610	Max. kerosine
Heavy gas oil (HGO)	610	800	Catalytic cracker or hydrocracker feed
Vacuum gas oil	800	1050	Deasphalter or catalytic cracker feed
	800	950	Catalytic cracker or hydrocracker feed
Vacuum Residue	950-1050	---	delayed coker or visbreaker

The atmospheric vacuum gas oil and the heavy vacuum gas oil are both could be theoretically and practically considered as a single product since both are sent simultaneously to the same unit like Catalytic cracker or hydrocracker feed (Gary & Handwerk, 2007). For this report, the used cuts for ADU/VDU products are reported in Table 5.

Table 5: The Used Cut for ADU/VDU Products

Cut#	Product	Endpoint (°F)
1	Off gas	50
2	Light straight run naphtha	158
3	Naphtha	356
4	Kerosene	464
5	Light diesel (LGO)	554
6	Heavy diesel (HGO)	644
7	Atm. Gas oil (AGO)	698
8	Vacuum gas oil (VGO)	734
9	Vac. Distillate	1022
9	Vac. Residue	--

The investment cost and operation cost estimation for ADU/VDU is divided into three sections: Desalter section cost, ADU section and VDU section (Gary & Handwerk, 2007). Tables 6-8 illustrate the investment cost and operation cost data with their conditions when applied in our model:

Table 6: The Investment Cost and Operation Cost Data for Desalting Section

Crude oil desalting unit investment cost		
Flowrate <100 MBPD	$0.5142 * F^{0.5046}$	Costs include: 1. Conventional electrostatic desalting unit 2. Water and Caustic injection systems 3. Water preheating and cooling
flowrate>100 MBPD	$0.2726 * F^{0.6432}$	
Operation cost data (per bbl feed):		
Power, kWh	0.01–0.02	
Water injection, gal	1-3	
Demulsifier chemical, lb	0.005–0.01	
Caustic, lb	0.001–0.003	

Table 7: The Investment Cost and Operation Cost Data for ADU Section

ADU investment cost		
Flowrate <100 MBPD	$11.459 * F^{0.4296}$	Costs include: 1. The side cuts and their strippers 2. The battery limits facilities 3. Heat exchange required to cool top products and side cuts to ambient temperature 4. Control system
flowrate>100 MBPD	$3.7885 * F^{0.6624}$	
Operation cost data (per bbl feed):		
Steam (300 psig), lb	10.0	The database at : LHV basis and 0.8 heaters efficiency
Power, kWh	0.9	
Cooling water, gal	150	
Fuel, MMBtu (KJ)	0.05	

Table 8: The Investment Cost and Operation Cost Data for VDU Section

VDU investment cost		
Flowrate <100 MBPD	$12.478 * F^{0.3884}$	Costs include: 1. Facilities for single vacuum gas oil column 2.The flash zone contains a three-stage jet system operate at 30 to 40 mmHg 3. Coolers and exchangers to reduce VGO temperature to ambient
flowrate>100 MBPD	$3.0131 * F^{0.6958}$	
Operation cost data (per bbl feed):		
Steam (300 psig), lb	10.0	The database at: LHV basis and 0.8 heaters efficiency
Power, kWh	0.3	
Cooling water, gal	150	
Fuel, MMBtu (KJ)	0.03	

3.1.2 Thermal Cracking (Delayed Coking)

The thermal cracking is a carbon rejection process that specifically applied to the bottom product in vacuum column. The unit cracks the vacuum residues (VR) under high temperature, 750–860 °F, to lighter products (olefinic and aromatic) and side products (coke)

The process includes: furnace to supply heat, which is required for thermal cracking, two coke drums where coking takes place, a fractionator and stripping section. Moreover, it contains exchangers for heating the feed as well as coolers to cool down the products. The expected products are only gas, naphtha, gas oil and coke.

The final products are estimated using correlations that involve calculating the weight percent of vacuum residue Conradson Carbon Residue (wt% CCR) and Specific gravity (SG) (Fahim, Alsahhaf , & Elkilani , 2010). Table 9 displays the correlations and Table 10 shows the expected compensation for gas out.

Table 9: Delayed Coking Product Correlations

Product correlations
$\boldsymbol{CCR(wt\%) = 0.0058 * Exp(7.8499 * SG)}$
$\boldsymbol{Gas\ (wt\%) = 7.8 + 0.144 * CCR}$
$\boldsymbol{Naphtha\ (wt\%) = 11.29 + 0.343 * CCR}$
$\boldsymbol{Gas\ oil\ (wt\%) = 100 - Gas - Naphtha - coke}$
$\boldsymbol{Coke\ (wt\%) = 1.6 * CCR}$
$\boldsymbol{Naphtha\ SG = 53.604 + 0.3404 * API}$
$\boldsymbol{Gas\ oil\ SG = 10.356 + 0.9131 * API}$
$\boldsymbol{Coke\ SG = \frac{SG\ VR - SG_{GAS} * Gas\ (wt\%) - SG_{Naphtha} * Naphtha(wt) - SG_{gasoil} * gas\ oil(wt\%)}{Coke\ (wt\%)}}$

Table 10: The Compensation for Gas out From Delayed Coking

Dry Gas	Mole%	wt%	Dry Gas	Mole%	wt%
Methane	51.4	37.35	Butenm	2.4	5.90
Ethene	1.5	1.91	i-Butane	1.0	2.55
Ethane	15.9	21.67	n-Butane	2.6	6.63
Propene	3.1	5.93	H2	13.7	1.24
Propane	8.2	16.43	CO2	0.2	0.40

The cost estimated is heavily dependent on the flowrate and the amount of coke product (Gary & Handwerk, 2007). Table 11 lists the cost estimated and operation cost.

Table 11: The Investment Cost and Operation Cost Data for Delayed Coking

Investment cost estimated:	
$\boldsymbol{Cost\ (30\ \frac{bbl\ fresh\ feed}{ton\ coke}) = 4.5239 * F + 31.383}$ $\boldsymbol{Cost\ (10\ \frac{bbl\ fresh\ feed}{ton\ coke}) = 5.458 * F + 46.249}$ $\boldsymbol{Ton = 2000\ lb}$ Linear interpolation is used to calculate for different given point of bbl fresh feed per ton coke	
Costs included: 1. Naphtha, light gas oil, and heavy gas oil are the products from Coker fractionator. 2. Systems for hydraulic decoking, coke dewatering, coke crushing, coke separation, coke storage where it covers three days product and coke drums 3. Utility systems: blowdown condensation and wastewater purification 4. Heat exchange and cooler use to cool products.	
Operation cost data (per bbl feed):	
Steam, lb/t coke	30
Power, kWh/t coke	70
Cooling water, gal/bbl feed [30°FΔT]	0.14
Fuel, MMBtu/bbl feed	700

3.1.3 Catalytic Hydrocracking

Hydrocracking is a hydrogen addition process. With the presence of hydrogen and a catalyst, the process alters high molecular weight feed stocks, gas oil from the ADU/VDU, the FCC, and the delayed coking, to lower molecular weight products, gasoline, jet fuels, and diesel. The process consists mainly of a reactor, hydrogen separator to recycle the unreacted hydrogen, fractionator, heating and cooling system.

The products lean on the hydrogen severity, mode of operation, such as naphtha maximum gasoline mode, maximum ATK mode (jet fuel) or maximum diesel fuel. In the same way, the nature of feed firmly affects the products like gas oil from ADU/VDU can produce Naphtha, jet fuel, and/or diesel. The hydrogen presences between (1.5 – 4) wt% of feed. In this research, the gas oil from ADU/VDU is assumed to be sent directly to Hydrocracking unit. In practice, refineries have been using the ATK mode to accommodate high demands for jet fuel. In order to reach maximum ATK, the hydrogen is assumed to be 3 wt% (Fahim, Alsahhaf , & Elkilani , 2010). We applied the catalytic hydrocracking product correlations as mention in Table 12.

Table 12: Catalytic Hydrocracking Product Correlations

Product correlations
$$H\ (wt\%) = \frac{lb\ Hydrogen}{lb\ of\ feed} * 100$$
$$VH = 0.6621 * H(wt\%) * SG$$
$$Gasoline\ (LV\%) = 0.03734\ APIf^2 + 1.57575\ APIf + 0.014923\ K - 1.36473\ VH - 0.16324\ \frac{VH}{K}$$
$$C4\ (LV\%) = 0.020359\ (LV\%\ Gasoline)^2 + 0.04888\ (LV\%\ Gasoline) + 0.108964 APIf$$
$$HN\ (LV\%) = -0.10322\ (LV\% Gasoline)^2 + 2.981215\ (LV\% Gasoline) - 0.07898\ APIf$$
ATK: Calculated by different between mass feed with hydrogen and the sum of the products (mass flow)
$$\text{Product wt\%} = 0.8672 * \text{Product LV\%} - 0.9969$$
where VH: hydrocracking severity

The cost estimation strongly relies on the flow rate and the amount of hydrogen used (Gary & Handwerk, 2007). Table 13 describes the cost estimated and operation cost equations:

Table 13: The Investment Cost and Operation Cost Data for Catalytic Hydrocracking

<table>
<tr><th colspan="2">Investment cost estimated:</th></tr>
<tr><td colspan="1">$$\boldsymbol{Cost\ (1000\frac{scf\ hydrogen\ consumption}{bbl\ feed})}$$ $$\boldsymbol{= 4.86 * F + 28.3}$$ $$\boldsymbol{Cost\ (2000\frac{scf\ hydrogen\ consumption}{bbl\ feed})}$$ $$\boldsymbol{= 6.58 * F + 36.7}$$ $$\boldsymbol{Cost\ (3000\frac{scf\ hydrogen\ consumption}{bbl\ feed})}$$ $$\boldsymbol{= 7.67 * F + 46.27}$$ Linear interpolation from three points is used to calculate at a given point of hydrogen consumption</td><td>**Costs included:**
1. Stabilization, Fractionation system, Complete reaction system
2. Hydrogen facilities, sulfide removal of hydrogen recycle and hydrogen recycle compressors
3. Heat exchange and cooler use to cool products
4. A central control system</td></tr>
<tr><td colspan="2">Operation cost data (per bbl feed):</td></tr>
<tr><td>Steam, lb</td><td>$$50 + 25 * (\frac{scf\ H\ consumption}{bbl\ feed * 1000} - 1)$$</td></tr>
<tr><td>Power, kWh</td><td>$$8 + 5 * (\frac{scf\ H\ consumption}{bbl\ feed * 1000} - 1)$$</td></tr>
<tr><td>Cooling water, gal circulation (30 F)</td><td>$$300 + 150 * (\frac{scf\ H\ consumption}{bbl\ feed * 1000} - 1)$$</td></tr>
<tr><td>Fuel (LHV), MMBtu</td><td>$$0.1 + 0.1 * (\frac{scf\ H\ consumption}{bbl\ feed * 1000} - 1)$$</td></tr>
<tr><td>Catalyst replacement, $</td><td>$$4E8 * (\frac{scf\ H\ consumption}{bbl\ feed})^2 - 4E5 * \frac{scf\ H\ consumption}{bbl\ feed} + 0.08$$</td></tr>
</table>

3.1.4 Fluidized Catalytic Cracking (FCC)

The Catalytic uses the catalyst and heat. The unit converts heavy products into more valuable products such as gasoline and lighter products. However, heavy carbon product (catalyst coke) is produced during the cracking. The process mainly consists of a reactor, fractionator and the catalyst regeneration system.

The correlations for the yield obtained from the regression of plant data using a zeolite catalyst (Fahim, Alsahhaf , & Elkilani , 2010). We found the following correlation for estimation the products (Table 14):

Table 14: FCC Product Correlations

Product correlations
$\boldsymbol{Coke\ wt\% = 0.05356 * CONV - 0.18598 * API + 5.966975}$
$\boldsymbol{LCO\ LV\% = 0.0047 * CONV2 - 0.8564 * CONV + 53.576}$
$\boldsymbol{Gases\ wt\% = 0.0552 * CONV + 0.597}$
$\boldsymbol{Gasoline\ LV\% = 0.7754 * CONV - 0.7778}$
$\boldsymbol{iC4\ LV\% = 0.0007 * CONV2 + 0.0047 * CONV + 1.40524}$
$\boldsymbol{nC4\ LV\% = 0.0002 * CONV2 + 0.019 * CONV + 0.0476}$
$\boldsymbol{C4 = LV\% = 0.0993 * CONV - 0.1556}$
$\boldsymbol{C3\ LV\% = 0.0436 * CONV - 0.8714}$
$\boldsymbol{C3 = LV\% = 0.0003 * CONV2 + 0.0633 * CONV + 0.0143}$
$\boldsymbol{HCO = 100 - CONV - (LCO\ LV\%)}$
$\boldsymbol{Gasoline\ API = 0.19028 * CONV + 0.02772 * (Gasoline\ LV\%) + 64.08}$
$\boldsymbol{LCO\ API = -0.34661 * CONV + 1.725715 * (Feed\ API)}$

The cost estimation is heavily dependent on the flowrate (Gary & Handwerk, 2007). Below is the estimate and projected operation cost (Table 15):

Table 15: The Investment Cost and Operation Cost Data for FCC

Investment cost estimated:	
$\boldsymbol{Cost = 55.178 * F^{0.345}}$	
Costs included: 1. Product fractionation, Gas compression, Complete reactor-regenerator section 2. Heat exchange and cooler use to cool products 3. A central control system	
Operation cost data (per bbl feed):	
Steam, lb	-30
Power, kWh	6
Cooling water, gal circulation (30 F)	500
Fuel (LHV), MMBtu	0.1
Catalyst replacement, $	0.25

3.1.5 Catalytic Reforming (CCR)

Refinery complex uses this process to boost the production of high octane gasoline from heavy naphtha. The operation alters the naphthenes completely to aromatics while altering the paraffins partially to aromatics. The final output includes methane, ethane, propane and butanes. The process consists of series of reactors. Usually, each reactor has its own individual heaters and column which is used to segregate products that are reformat products such as hydrogen & gases and in butane (Fahim, Alsahhaf , & Elkilani , 2010). Table 16 presents the yield correlation used in this study.

Table 16: CCR Product Correlations

Product correlations
$H_2(wt\%) = -12.1641 + 0.06134 * C_5(vol\%) + 0.099482 * RONR$
$C_1(wt\%) = 11.509 - 0.125 * C5\ vol\ \%$
$C_2(wt\%) = 16.496 - 0.1758 * C5\ vol\ \% \ (5.10)$
$C_3(wt\%) = 24.209 - 0.2565 * C5\ vol\ \% \ (5.11)$
$Total\ C_4(wt\%) = 27.024 - 0.2837 -* C5\ vol\ \% \ (5.12)$
$nC_4(wt\%) = 0.585 * total\ C_4(wt\%)$
$iC_4(wt\%) = 0.415 * total\ C_4(wt\%)$
$C_5(vol\%) = -0.03806 * RONR^2 - 6.292793 * RONR - 14.4337 * K$
$SCFB\ H2 = 0.0002\ 0.48848\ H2\ (wt\%)$

The cost estimated is heavily dependent on the flowrate (Gary & Handwerk, 2007). The following are the cost estimated and operation cost (Table 17):

Table 17: The Investment Cost and Operation Cost Data for CCR

<table>
<tr><th colspan="3">Investment cost estimated:</th></tr>
<tr><td>Continues catalyst regeneration</td><td colspan="2">$Cost\ = 8.172 * F^{0.5684}$</td></tr>
<tr><td>Fixed bad</td><td colspan="2">$Cost\ = 11.622 * F^{0.5697}$</td></tr>
<tr><td colspan="3">Costs included:
1. All battery limit facilities (reactors, heaters, cooling and heating system…)
2. Stabilizer system for the products
3. A central control system</td></tr>
<tr><td colspan="3">Operation cost data (per bbl feed):</td></tr>
<tr><td colspan="2">Steam, lb.</td><td>30</td></tr>
<tr><td colspan="2">Power, kWh</td><td>3</td></tr>
<tr><td colspan="2">Cooling water, gal</td><td>400</td></tr>
<tr><td colspan="2">Fuel gas (LHV), MMBtu</td><td>0.3</td></tr>
<tr><td colspan="2">Catalyst replacement</td><td>0.16</td></tr>
</table>

3.2 Second Stage Methodology

In this stage, the Peregrino crude oil, which is a heavy offshore crude with an API 13.7°, will be used to compare the five scenarios (Table 18). For evaluation, the excel sheet will be used. In the simulation, the Cumulative yield of the ADU/VDU bottom product will be defined. For five scenarios, a gasifier will be simulated as RGibbs for the gasification section to calculate the amount of syngas out. Moreover, the H/C ratio coming out from the gasifier will be calculated by the Analyzer Tool in Aspen. In the three scenarios, the sweeting block will be either the desulfurization or sulfur remover for syngas. Figure 10 is the block flow diagram for gasification scenarios.

Table 18: Cumulative Yield of the Peregrino Crude Oil Bottom Product

Cumulative Yield (%)	Boiling Point (°F)	Cumulative Yield (%)	Boiling Point (°F)
0.0	970	34.1	1140
2.2	980	36.1	1150
4.4	990	38.1	1160
6.5	1000	40.1	1170
8.5	1010	42.0	1180
10.5	1020	44.0	1190
12.5	1030	45.9	1200
14.4	1040	47.9	1210
16.4	1050	49.8	1220
18.3	1060	51.6	1230
20.2	1070	53.5	1240
22.2	1080	55.3	1250
24.2	1090	57.0	1260
26.1	1100	58.8	1270
28.1	1110	60.5	1280
30.1	1120	62.1	1290
32.1	1130	63.8	1300

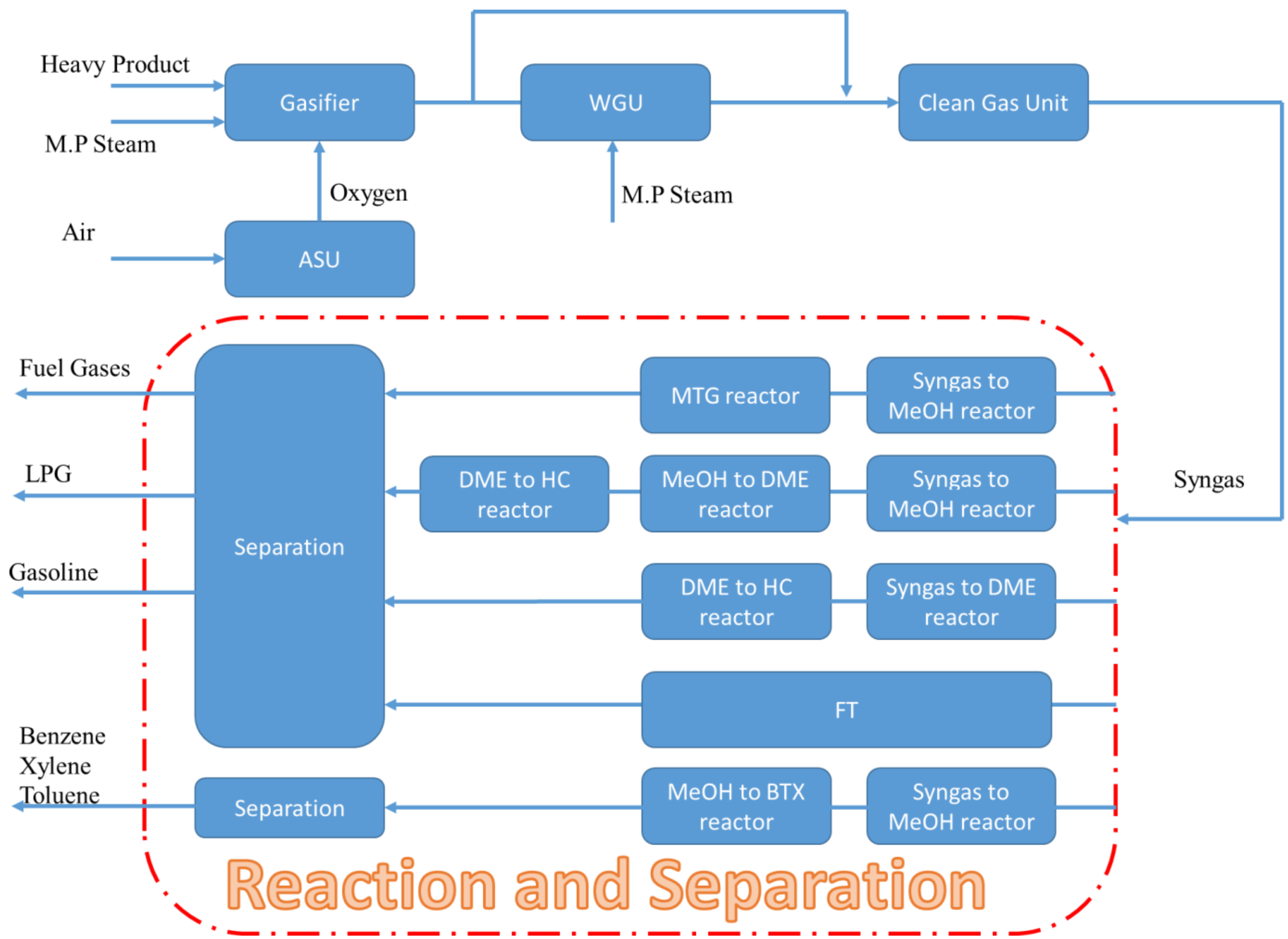

Figure 10: The Block Flow Diagram for Gasification to Power Scenario

CHAPTER IV

CASE DESCRIPTIONS, RESULTS AND DISCUSSION

4.1 The First Stage

4.1.1 The First Stage: Case Descriptions

The modern refinery scheme is arranged to evaluate the effects of varying the crude oil API on mass balance within the refinery. An advance integrative algorithm was designed with the crude oil feed (100,000 BPD), density, TBP and light components composition as user-defined input parameters. In refinery sachem, we sought obligated to take advantage of the atmospheric distillation unit, vacuum distillation unit, delayed coke, hydrocracking, Fluidized catalytic cracking, Catalytic reforming (CCR) and Isomerization. However, the delayed coker was not used for crude oil with high API (API=60.4).

Figure 11 illustrates the refinery scheme that was used for evaluations. The final products from the refinery sector are Gases, LPG, Gasoline, Kerosene Diesel, coke and hydrogen. At the end, the hydrogen will be used in hydrotreat refinery's products.

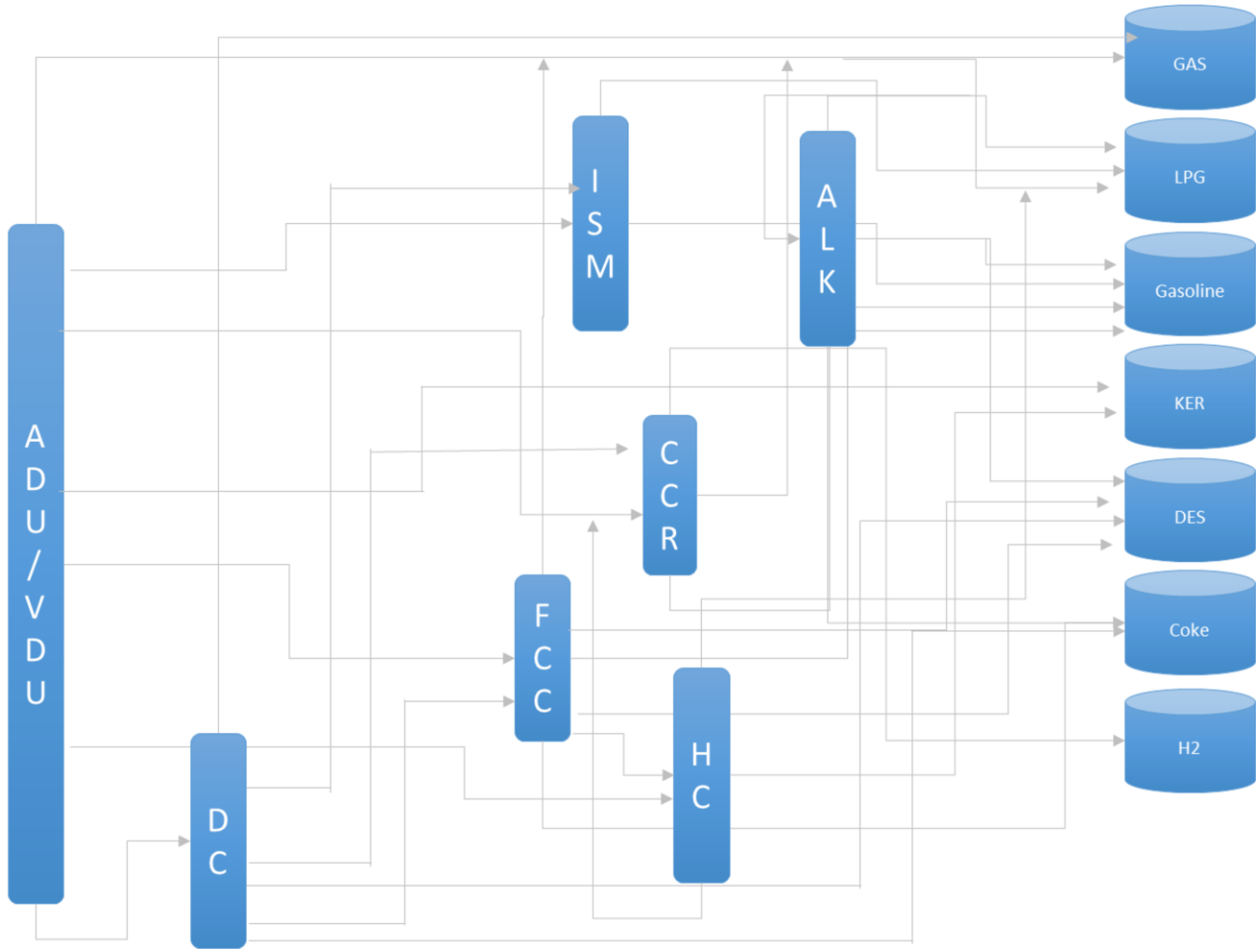

Figure 11: Refinery Scheme

Table 19: Refinery Desecration

Unit	Feeds	Products
ADU/VDU	Crude oil	Gasses, LPG, Naphtha, Gasoline, Kerosene, Diesel, Gas oil and Vacuum residue
Delayed Coker (DC)	Vacuum residue (ADU/VDU)	Gasses, LPG, Diesel, coke, Naphtha and Gas oil
Fluidized catalytic cracking (FCC)	Gas oil (ADU/VDU) Gas oil (DC)	Gasses, LPG, gasoline, Diesel, Light cycle oil, Heavy cycle oil and coke
Hydrocracking (HC)	Vacuum gas oil (ADU/VDU) Heavy cycle oil (FCC)	LPG, Gasoline, Naphtha, Kerosene and gasoline
Catalytic reforming (CCR)	Heavy naphtha (ADU/VDU) Heavy naphtha (DC) Heavy naphtha (HC)	Hydrogen, gases, LPG and gasoline
Isomerization (ISM)	Light naphtha (ADU/VDU) Light naphtha (DC)	LPG and gasoline
Alkylation	LPG (all refinery units)	LPG, gasoline, diesel and coke

Table 19 elaborate about the refinery scheme. The nine samples were sub-grouped according to their oil API. Then, the mass balance was evaluated for each sample. In this report, five major points of comparison had been utilized: the products of ADU/VDU, Feed to refinery units, refinery final products, refinery capital cost and utilities. The result of refinery balances and figures of the comparison is reported in the Appendix A.

4.1.2 The First Stage: Result

Table 20 and Table 21 are summery of a result for ADU/VDU products section and feed charge to refinery units section:

Table 20: ADU/VDU Products

ADU/VDU Products	API < 20			API 20 - 30		
API	-31.40			-15.20		
Butane and lighter	5.15E+03	1.89E+03	-63.3	8.25E+03	4.27E+03	-48.3
LSR	5.44E+03	5.91E+02	-89.1	1.27E+04	9.47E+03	-25.7
HSR	4.74E+04	3.79E+04	-20.1	1.33E+05	8.80E+04	-33.6
Kerosene	7.50E+04	4.95E+04	-34.0	1.11E+05	9.76E+04	-12.0
LGO	1.17E+05	6.17E+04	-47.1	1.20E+05	1.20E+05	0.4
HGO	1.35E+05	7.87E+04	-41.8	1.29E+05	1.35E+05	4.6
AGO	8.04E+04	4.39E+04	-45.4	7.47E+04	7.69E+04	3.0
VGO	3.74E+05	3.55E+05	-5.1	3.06E+05	3.16E+05	3.4
VR	4.82E+05	7.27E+05	50.9	3.76E+05	4.53E+05	20.3
	API 30 - 40			API > 40		
API	-19.93			-15.95		
Butane and lighter	2.87E+04	1.41E+04	-51.0	6.75E+04	8.19E+04	21.3
LSR	4.37E+04	2.14E+04	-51.1	1.25E+05	1.18E+05	-5.3
HSR	2.50E+05	1.56E+05	-37.6	6.01E+05	3.81E+05	-36.5
Kerosene	1.34E+05	1.22E+05	-9.3	1.79E+05	1.18E+05	-34.2
LGO	1.26E+05	1.10E+05	-12.4	8.59E+04	9.45E+04	10.0
HGO	1.20E+05	1.11E+05	-7.9	3.81E+04	9.04E+04	137.6
AGO	6.09E+04	6.61E+04	8.5	9.26E+03	4.20E+04	353.8
VGO	2.54E+05	2.97E+05	17.0	6.72E+03	1.65E+05	400.0
VR	2.01E+05	3.67E+05	82.4	0.00E+00	6.37E+04	--

Table 21: Feed Charge to Refinery Units

Units	API < 20			API 20 - 30		
API	-31.40			-15.20		
ADU	1.36E+06	1.43E+06	4.8	1.28E+06	1.32E+06	2.9
LSR	3.06E+04	4.72E+04	53.9	3.06E+04	3.23E+04	5.5
AKL	2.51E+04	2.10E+04	-16.5	2.62E+04	2.53E+04	-3.4
CCR	1.62E+05	1.91E+05	17.7	2.22E+05	1.89E+05	-14.7
FCC	3.08E+05	1.98E+05	-35.7	2.88E+05	3.04E+05	5.7
HC	4.15E+05	3.85E+05	-7.2	3.42E+05	3.55E+05	3.6
DC	4.82E+05	7.27E+05	50.9	3.76E+05	4.53E+05	20.3
Units	API 30 - 40			API > 40		
API	-19.93			-15.95		
ADU	1.22E+06	1.27E+06	4.7	1.08E+06	1.13E+06	5.3
LSR	5.30E+04	3.91E+04	-26.3	1.25E+05	1.21E+05	-2.9
AKL	2.42E+04	2.53E+04	4.3	4.78E+03	1.54E+04	221.7
CCR	3.13E+05	2.43E+05	-22.3	6.02E+05	4.16E+05	-31.0
FCC	2.28E+05	2.57E+05	12.8	4.73E+04	1.47E+05	210.9
HC	2.83E+05	3.30E+05	16.7	1.14E+04	1.83E+05	1506.3
DC	2.01E+05	3.67E+05	82.4	0.00E+00	6.37E+04	--

Table 22: Refinery Final Products

Refinery Products	API < 20			API 20 - 30		
API	-31.40			-15.20		
Gases	5.00E+04	7.57E+04	51.6	4.46E+04	4.91E+04	10.0
LPG	7.53E+04	7.44E+04	-1.3	7.36E+04	7.22E+04	-1.9
Gasoline	3.19E+05	2.97E+05	-6.9	3.50E+05	3.37E+05	-3.7
K/Jet fuel	3.60E+05	3.14E+05	-12.8	3.46E+05	3.41E+05	-1.5
Diesel	3.35E+05	2.32E+05	-30.7	3.20E+05	3.38E+05	5.7
Coke	1.17E+05	2.89E+05	147.8	6.82E+04	9.87E+04	44.8
H2	2.81E+04	3.37E+04	19.8	3.78E+04	3.25E+04	-14.1
Refinery Products	API 30 - 40			API > 40		
API	-19.93			-15.95		
Gases	3.70E+04	4.40E+04	18.9	3.23E+04	3.09E+04	-4.2
LPG	8.91E+04	7.71E+04	-13.6	1.30E+05	1.41E+05	8.4
Gasoline	4.02E+05	3.56E+05	-11.4	5.61E+05	4.90E+05	-12.7
K/Jet fuel	3.28E+05	3.48E+05	6.1	1.87E+05	2.44E+05	30.3
Diesel	2.48E+05	2.99E+05	20.4	9.36E+04	1.45E+05	55.3
Coke	3.61E+04	6.90E+04	91.1	1.98E+03	1.46E+04	639.8
H2	5.29E+04	4.14E+04	-21.7	1.01E+05	7.00E+04	-30.9

Table 23: Refinery Total Utilities

Utilities	API < 20			API 20 - 30		
API	-31.40			-15.20		
Power, kWh	1.52E+0	3.02E+0	98.7	1.05E+0	1.31E+0	25.3
Steam, lb/day	3.85E+0	5.72E+0	48.5	3.19E+0	3.41E+0	7.0
Cooling water, gal/day	6.60E+0	8.74E+0	-86.8	6.36E+0	6.42E+0	1.0
Fuel MMBtu/day	2.47E+0	2.93E+0	18.6	2.40E+0	2.43E+0	1.3
Utilities	API 30 - 40			API > 40		
API	-19.93			-15.95		
Power, kWh	7.50E+0	1.05E+0	40.3	2.92E+0	5.06E+0	73.2
Steam, lb/day	2.81E+0	3.26E+0	15.8	1.71E+0	2.24E+0	31.1
Cooling water, gal/day	6.17E+0	6.33E+0	2.6	5.05E+0	5.91E+0	17.0
Fuel MMBtu/day	2.35E+0	2.43E+0	3.4	2.41E+0	2.34E+0	-2.8

Table 22 to Table 24 are summery of a result for refinery final products section, refinery total utilities and refinery total cost.

Table 24: Refinery Total Cost

API	-31.40			-15.20		
	API < 20			API 20 - 30		
ADU	152	155	1.9	146	148	1.5
LSR	258	313	21.5	191	217	13.5
AKL	162	139	-14.2	158	161	1.9
CCR	221	216	-2.1	192	197	2.8
FCC	58	68	17.0	69	64	-6.8
HC	11	11	0.0	11	11	0.0
DC	23	22	-6.0	24	24	-1.7
	API 30 - 40			API > 40		
API	-19.93			-15.95		
ADU	138	145	5.1	100	127	26.4
LSR	122	183	49.9	0	72	--
AKL	147	152	3.8	86	127	47.1
CCR	162	185	14.6	43	116	172.6
FCC	82	73	-10.7	112	91	-18.3
HC	11	11	0.0	11	11	0.0
DC	23	24	2.1	14	19	35.3

4.1.3 The First Stage: Discussion

Simulating the comparison yielded different scenarios where products either increased or decreased based on the insitu oil API. Table 25 summarize our findings.

Table 25: ADU/VDU Products Analysis

ADU/VDU Products	API Below 20	API 20 to 30	API 30 to 40	API Above 40
API	**-31.40**	**-15.20**	**-19.93**	**-15.95**
Butane and lighter	↓	↓	↓	↑
LSR	↓	↓	↓	↓
HSR	↓	↓	↓	↓
Gasoline	↓	↓	↓	↓
LGO	↓	↑	↓	↑
HGO	↓	↑	↓	↑
AGO	↓	↑	↑	↑
VGO	↓	↑	↑	↑
VR	↑	↑	↑	↑

The ADU/VDU presented the first change of API declining. In our findings, the gases, light and heavy naphtha, and kerosene will decrease as the API decrease; which was observed in our experiment. On the other hand, the gases for API showed an increase in gasses due to the presence of dissolved gas in insitu crude oil. Moreover, atmospheric gas oil (AGO), vacuum gas oil (VGO) and vacuum residue (VR) will increase. We observed that that heavy oil with API below 20 will result to more vacuum residue and less all other products. Such observation will later play a key role in our interpretation and analysis. In fact, the light gas oil (LGO) and heavy gas oil (HGO) also depend in the TBP of crude oil with API.

Table 26: Feed Charge to Refinery Units Analysis

Feed Change to Refinery Unties	API Below 20	API 20 to 30	API 30 to 40	API Above 40
API	**-31.40**	**-15.20**	**-19.93**	**-15.95**
ADU	--	--	--	--
LSR	--	--	--	--
AKL	↓	--	--	↑
CCR	--	↓	↓	↓
FCC	↓	↑	↑	↑
HC	↓	↑	↑	↑
DC	↑	↑	↑	↑

In Table 26, the delayed coker shows an increasing of the feed in all cases. It is expected to see an increase because we also would observe an increase in heavy products of the feed. Similar analogy and conclusion could take place for the Fluidized catalytic cracking (FCC) and Hydrocracking (HC). FCC and HC handle the heavy products in the refinery. They feed decline at the API blow 20 because the heavy products from ADU/VDU would be accumulated in the delayed coke. Catalytic reforming (CCR) recovers the middle products and expected to fall off.

Table 27: Refinery Final Products Analysis

Refinery Products	API Below 20	API 20 to 30	API 30 to 40	API Above 40
API	-31.40	-15.20	-19.93	-15.95
Gases	↑	↑	↑	↑
LPG	↓	↓	↓	--
Gasoline	↓	↓	↓	↓
K/Jet fuel	↓	↓	--	↓
Deasil	↑	--	↑	↑
Coke	↑	↑	↑	↑
H2	--	↓	↓	↓

In Table 27, the growing of the gas products in all cases took place due to the catalytic and thermal cracking of the heavier products from ADU/VDU and other units. In contrast, the LPG, gasoline, kerosene and jet fuel have been reduced since their main supplier (ADU/VDU) is also being reduced and the thermal and catalyst cracking don't substitute the change. All those declines products are accumulated in the coke. Hydrogen is a plus byproduct form implementing CCR. Therefore, the reductions in the table are mainly caused by the feed entering into the unit

Table 28: Refinery Total Utilities Analysis

Utilities	API Below 20	API 20 to 30	API 30 to 40	API Above 40
API	-31.40	-15.20	-19.93	-15.95
Power, kWh	↑	↑	↑	↑
Steam, lb/day	↑	↑	↑	--
Cooling water, gal/day	↓	--	--	↑
Fuel MMBtu/day	↑	--	--	--

In Table 28, the power and the steam required by the refinery increased as the crude oil became heavier. However, the heating fuel does not show the expected response. We interpreted this to be caused by the fact that middle cracking units and heavy cracking units are fuel heavy user. Consequently, the cumulative of the heavier products did not get affected nor required more fuel.

Table 29: Refinery Total Cost

Feed Change to Refinery Unties	API Below 20	API 20 to 30	API 30 to 40	API Above 40
API	-31.40	-15.20	-19.93	-15.95
ADU	↑	↑	↑	↑
LSR	--	--	--	--
AKL	↓	↓	↑	↑
CCR	↑	↓	↓	↓
FCC	↓	↑	↑	↑
HC	↓	↑	↑	↑
DC	↑	↑	↑	↑

In Table 29, the value of refinery units would be directly correlated to the feed charging out of the unit. The relatively heavy units (FCC, HC and DC) cost ascended due to the feed. However, the FCC and HC cost have been optimized and lowered as more coke accumulated in the DC. On the other hand, the ADU/VDU cost kept growing as it handled heavy oil.

4.2 The Second Stage

4.2.1 The Second Stage: Case Descriptions

The Peregrino (API is 13.7° and S.G. is 0.9743) was used in this stage. First, we had to remove the Delayed Coker unit from refinery balance and accounts the difference to make the base case for the evaluation. Tables 30-32 summarized the final products, capital cost and utilities changes:

Table 30: Products Reduction from the Base Case

(lb/hr)	Base case	DC out	Change
Gases	7.57E+04	1.02E+04	6.55E+04
LPG	7.16E+04	3.58E+04	3.57E+04
Gasoline	3.00E+05	1.43E+05	1.58E+05
K/Jet fuel	3.14E+05	3.09E+05	4.91E+03
Diesel	2.32E+05	8.14E+04	1.50E+05
Coke	2.89E+05	5.22E+03	2.84E+05
H2	3.37E+04	1.63E+04	1.74E+04

Table 31: Capital Cost Different From the Base Case

(MM$)	Base case	DC out	Change
ADU/VDU System	155	155	-
DC	313	-	313.3
FCC	139	119	19.2
HC	216	213	3.4
CCR	68	46	22.0
ISO	11	11	-
ALK	21	16	4.8
Total			362.7

Table 32: Utilities Different From the Base Case

	Base case	DC out	Change
Power, kWh	3.01E+06	5.34E+05	2.47E+06
Steam, lb/day	5.69E+06	3.01E+06	2.69E+06
Cooling water, gal/day	8.70E+06	3.54E+06	5.16E+06
Fuel MMBtu/day	2.71E+04	1.65E+04	1.06E+04

The removed flow, feed to delayed coker, will be gasified to yield syngas where can be utilized to produce hydrocarbon and/or power. Five cases were evaluated: Methanol to Gasoline (MTG), DME to hydrocarbon (indirect-through MeOH), DME to hydrocarbon (direct method), Fischer Tropsch (FT), and BTX production. The hydrocarbon production cases all share the following common processes: gasification and water gas shifting, syngas cleaning, reaction process and separation process. In brief, the heavy oil, steam and oxygen are injected into the gasified system to produce syngas. The gasified operates at 352 psig and 2192 °F (Cheng & Kung, 1994). Aspen-plus was used to calculate the conversion of the heavy products to syngas. The H2/CO ratio for syngas is 1.00. In all hydrocarbon cases expect DME to Gasoline (direct method), part of the syngas is sent to water gas shifting unit (WGS) in order to increase the H2/CO from

1.00 to 2.00. The DME to Gasoline (direct method) uses a ratio of 1 in the reaction section. Next, the syngas is cleaned (to remove H2S and CO2) in amine system. Prior to the final product, the clean syngas was sent to the reaction section to yield hydrocarbon. Finally, the separation unit outputted the final products. Tables 33-37 contain the mass balance for MTG, DME to H-C (indirect and direct), FT and MeOH to BTX.

4.2.2 The Second Stage: Result

The block flow diagram of five cases is illustrated in Figure 12.

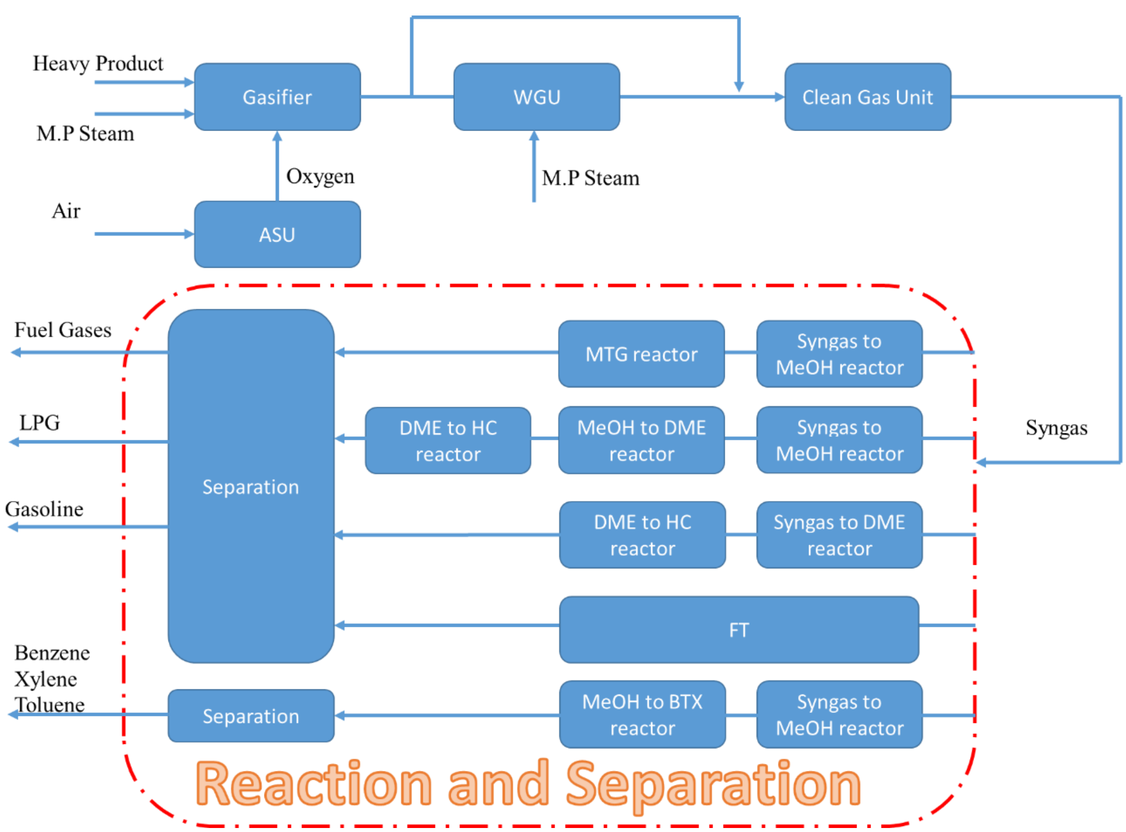

Figure 12: The Block Flow Diagram for Gasification Cases

Table 33: MTG Case Streams Summary

Mass Flow lb/hr	Heavy Product	Feed to Gasifier	Out from Gasifier	WGS out	Clean syngas
WATER	0	446878	28628	0	0
H2	0	0	108797	146069	145895
CO2	0	0	28770	848771	68651
CO	0	0	1529195	1007376	998929
CH4	0	0	24715	24715	24715
O2	0	522472	0	0	0
N2	0	0	0	0	0
S	0	0	0	0	0
H2S	0	0	0	0	0
COS	0	0	0	0	0
DIMET-01	0	0	0	0	0
PC2414F	750755	750755	0	0	0
Mass Flow lb/hr	**Syngas to MeOH out**	**Feed to MTG**		**MTG out**	
WATER	0	220101	water	858571	
H2	3477	3477	H2	3477	
CO2	68651	68651	CO2	68651	
CO	2003	2003	CO	2003	
CH4	24715	24715	Light gas	34748	
O2	0	0	Propane	25083	
N2	0	0	Propylene	5017	
H2S	0	0	Isobutane	35116	
COS	0	0	n-Butane	25083	
CH3OH	1139344	1139344	Butenes	5017	
C2H6O	562	562	C5+ gasoline	396308	
C3H8O	220	220			
Total Flow	1238972	1459072	Total Flow	1459072	

(Phillips, Tarud, Biddy, & Dutta, 2011)

Table 34: DME to H-C (Indirect) Case Streams Summary

Mass Flow lb/hr	Heavy Product	Feed to Gasifier	Out from Gasifier	WGS out	Clean syngas
WATER	0	446878	28628	0	0
H2	0	0	108797	146069	145895
CO2	0	0	28770	848771	68651
CO	0	0	1529195	1007376	998929
CH4	0	0	24715	24715	24715
O2	0	522472	0	0	0
N2	0	0	0	0	0
H2S	0	0	0	0	0
COS	0	0	0	0	0
DIMET-01	0	0	0	0	0
PC2414F	750755	750755	0	0	0
Total Flow lb/hr	750755	1720105	1720105	2026932	1238190
	Syngas to MeOH out	MeOH to DME		DME to H-C	
WATER	0	330636	WATER	427776	
H2	3477	3477	H2	0	
CO2	68651	68651	CO2	68651	
CO	2003	2003	CO	2003	
CH4	24715	24715	CH4	24715	
O2	0	0	Lightgas	2650	
N2	0	0	Propane	13250	
H2S	0	0	Propylene	132	
COS	0	0	Isobutane	105998	
CH3OH	1139344	0	n-Butane	1987	
C2H6O	562	0	Butenes	2650	
C3H8O	220	0	C5+ gasoline	536614	
DME	0	809489			
Total Flow lb/hr	1238972	1238972	Total	1246567	

(Phillips, Tarud, Biddy, & Dutta, 2011)

Table 35: DME to H-C (Direct) Case Streams Summary

Mass Flow lb/hr	Heavy Product	Feed to Gasifier	Out from Gasifier	Clean syngas
WATER	0	446878	28628	28628
H2	0	0	108797	108667
CO2	0	0	28770	2327
CO	0	0	1529195	1516373
CH4	0	0	24715	24715
O2	0	522472	0	0
H2S	0	0	0	0
COS	0	0	0	0
DIMET-01	0	0	0	0
PC2414F	750755	750755	0	0
Total Flow	750755	1720105	1720105	1680710
	syngas to DEM out			DME to H-C
WATER	41235		WATER	184508
H2	4325		H2	60551
CO2	689235		CO2	250631
CO	90982		CO	103980
CH4	24715		CH4	0
O2	0		O2	0
DME	750104		S	0
MEOH	81944		DME	0
			MEOH	81944
			Lightgas	2456
			Propane	12278
			Propylene	123
			Isobutane	98222
			n-Butane	1842
			Butenes	2456
			C5+ gasoline	497247
Total Flow	1682541		Total	1296236

(Ogawa, Inoue, Shikada, & Ohno, Direct Dimethyl Ether Synthesis, 2003) (Tan, et al., 2015)

Table 36: FT Case Streams Summary

Mass Flow lb/hr	Heavy Product	Feed to Gasifier	Out from Gasifier	WGS out	Clean syngas
WATER	0	446878	28628	0	0
H2	0	0	108797	146069	145895
CO2	0	0	28770	848771	68651
CO	0	0	1529195	1007376	998929
CH4	0	0	24715	24715	24715
O2	0	522472	0	0	0
DIMET-01	0	0	0	0	0
PC2414F	750755	750755	0	0	0
Total Flow	750755	1720105	1720105	2026932	1238190
	FT Products				
WATER	629325				
H2	450.				
CO2	68650.				
CO	19978				
carbon no.		carbon no.			
C1	3300	C16	14223		
C2	5769	C17	13426		
C3	7692	C18	13321		
C4	9531	C19	13118		
C5	10824	C20	12817		
C6	12027	C21	12419		
C7	12936	C22	11922		
C8	13551	C23	11328		
C9	13873	C24	10636		
C10	14398	C25	11076		
C11	14726	C26	10237		
C12	14859	C27	9300		
C13	14796	C28	9643		
C14	14537	C29	8559		
C15	14083	C30	156394		
Total	895309				

(Gabriel, et al., 2014)

Table 37: MeOH to BTX Case Streams Summary

Mass Flow lb/hr	Heavy Product	Feed to Gasifier	Out from Gasifier	WGS out	Clean syngas
WATER	0	446878	28628	0	0
H2	0	0	108797	146069	145895
CO2	0	0	28770	848771	68651
CO	0	0	1529195	1007376	998929
CH4	0	0	24715	24715	24715
O2	0	522472	0	0	0
N2	0	0	0	0	0
H2S	0	0	0	0	0
COS	0	0	0	0	0
DIMET-01	0	0	0	0	0
PC2414F	750755	750755	0	0	0
Total Flow lb/hr	750755	1720105	1720105	2026932	1238190
	Syngas to MeOH out			MeOH to BTX	
WATER	0		H2	3477	
H2	3477		CO2	68651	
CO2	68651		water	641167	
CO	2003		C1 and C2	44674	
CH4	24715		C3	44906	
O2	0		C4	49896	
N2	0		C5+	9979	
H2S	0		Benzene	24948	
COS	0		Toluene	144698	
CH3OH	1139344		Xylene	154677	
C2H6O	562		C9+	49896	
C3H8O	220				
Total Flow	1238190		Total Flow	1236969	

4.2.3 The Second Stage: Discussion

Due to the reduction of the liquid fuel from removing the Delayed Coker, the cases that produce the liquid fuel are more verbal to be focused on. The four processes (MTG, DME to H-C (indirect), DME to H-C (direct) and FT) show a similar amount of liquid fuel production. Moreover, liquid production is more from the base case by 120,000 to 130,000 lb/hr (Table 38). This slight change is due the chemical distribution of the products (gases LPG, Gasoline and Diesel). Another important finding, it is expected to have an increasing on production of gasification cases from the base case by around 50% of coke production from base case since the other 50 % was used to improve the H2/CO ratio.

Table 38: Products Comparison between the Base Case and Liquid Fuel Cases

	Base case	MTG	DME to H-C (indirect)	DME to H-C (direct)	FT
	lb/hr				
Gases	65,504	42,110	2,409	2,232	9,069
LPG	35,747	52,637	112,743	104,472	22,991
Gasoline and Diesel	313,228	431,624	487,831	452,043	469,029
Coke	284,238	0	0	0	0
Gases		-36%	-96%	-97%	-86%
LPG		47%	215%	192%	-36%
Gasoline and Diesel		38%	56%	44%	50%

The capital costs were calculated based on different methods. Most of those methods were estimated using the six-tenths factor rule based other studies and reasonable assumptions (Swanson, Satrio, Brown, Platon, & Hsu, 2010) (Phillips, Tarud, Biddy, & Dutta, 2011) (Tan, et al., 2015). The Chemical Engineering plant cost index Table 39 was utilized to adjust prices to a common basis year (year 2015) since different sources were used.

Table 39: Chemical Engineering Index

Year	Index	Year	Index
2005	468.2	2011	585.7
2006	499.6	2012	584.6
2007	525.4	2013	567.3
2008	575.4	2014	576.1
2009	521.9	2015	560.7
2010	550.8		

The MTG case, DME to H-C (Indirect and direct), FT case and MeOH to BTX case are reported in Tables 40-44.

Table 40: MTG Capital Cost Calculation

MTG						
	Flow lb/hr	Base Cost (MM$)	Base flow lb/hr	Cost (MM$)	Year	Cost (MM$, 2015)
Gasifier	1.72E+06	28.2	2.97E+05	97	2007	103.03
ASU	5.22E+05	19.5	5.68E+04	92	2007	98.41
Amine system	2.03E+06	12.1	3.45E+05	42	2007	44.61
WGS	5.87E+05	0.136	9.38E+04	1.34	2007	1.42
Syngas to MeOH	1.24E+06	15.3	2.01E+05	55	2007	58.26
MTG and Products recovery	1.36E+06	21.6	8.44E+04	151	2007	161.42
			Total	438		467.15

Table 41: DME to H-C (Indirect) Capital Cost Calculation

DME to H-C (indirect)						
	Flow lb/hr	Base Cost (MM$)	Base flow lb/hr	Cost (MM$)	Year	Cost (MM$, 2015)
Gasifier	1.72E+06	28.2	2.97E+05	97	2007	103.03
ASU	5.22E+05	19.5	5.68E+04	92	2007	98.41
Amine system	2.03E+06	12.1	3.45E+05	42	2007	44.61
WGS	5.87E+05	0.136	9.38E+04	1.34	2007	1.42
syngas to MeOH	1.24E+06	15.3	2.01E+05	55	2007	58.26
MeOH to DME to H-C	1.14E+06	38.7	8.77E+04	233	2011	223.10
Products recovery	8.09E+05	5.4	4.66E+04	40	2011	38.22
			Total	559		567.04

Table 42: DME to H-C (Direct) Capital Cost Calculation

DME to H-C direct						
	Flow lb/hr	Base Cost (MM$)	Base flow lb/hr	Cost (MM$)	Year	Cost (MM$, 2015)
Gasifier	1.72E+06	28.2	2.97E+05	97	2007	103.03
ASU	5.22E+05	19.5	5.68E+04	92	2007	98.41
Amine system	1.72E+06	0.13	9.38E+04	2.83	2007	3.02
Syngas to DME	1.68E+06	15.3	2.01E+05	68	2007	62.06
DME to H-C	8.32E+05	21.6	8.44E+04	107	2007	98.41
			Total	366		364.92
Note: the price for the syngas to DME and DME to H-C were calculated bass in the MTG case with a reduction of 14% (Ogawa, Inoue, Shikada, Inokoshi, & Ohno, Direct Synthesis of Dimethyl Ether form Synthesis Gas, 2015)						

Table 43: FT Capital Cost Calculation

FT						
	Flow lb/hr	Base Cost (MM$)	Base flow lb/hr	Cost (MM$)	Year	Cost (MM$, 2015)
Gasifier	1.72E+06	28.2	2.97E+05	97	2007	103.03
ASU	5.22E+05	19.5	5.68E+04	92	2007	98.41
WGS	5.87E+05	0.1	9.38E+04	1.34	2007	1.42
Amine system	2.03E+06	29.3	5.04E+05	78	2007	82.80
Syngas to FT and Products recovery	1.24E+06	704.7	2.51E+06	430	2006	482.15
			Total	697		768

Table 44: MeOH to BTX Capital Cost Calculation

Syngas to BTX (Methanol path)						
Unit	Flow lb/hr	Base Cost (MM$)	Base flow lb/hr	Cost (MM$)	Year	Cost (MM$, 2015)
Gasifier	1.72E+06	28.2	2.97E+05	97	2007	103.03
ASU	5.22E+05	19.5	5.68E+04	92	2007	98.41
Amine system	2.03E+06	12.1	3.45E+05	42	2007	44.61
Syngas to MeOH	1.24E+06	15.3	2.01E+05	55	2007	58.26
BTX reactor	1.14E+06	35.6	8.44E+04	220	2011	210.94
Products recovery	1.14E+06	0.58	1.84E+04	11	2006	11.81
			Total	516		527.06
Note: the capital cost for the BTX reactor was assumed like the DME to the hydrocarbon system since both of the share the same single conversion (40% DME conversion).						

(Draghiciu, Isopescu, & Woinaroschy, 2009)

The five cases capital cost is reported in Tables 40-44. At first glance, the direct path of DME is the lowest estimated cost of the other cases that produce gasoline. The

MTG path and indirect DME path share similar cost required to produce one gallon of gasoline in a plant with 20 years as lifetime (Table 45). The most important finding is that the cost for MTG and DME paths are below the reduced price of the base case. In contrast, the Fischer Tropsch showed a higher price.

Table 45: Capital Cost Comparison between the Base Case and Liquid Fuel Cases

	Base case	MTG	DME to H-C (indirect)	DME to H-C (direct)	FT
Gasoline and Diesel (Gal/hr)	50,116	69,060	78,053	72,327	75,045
Cost(MM$)	434.4	467.1	567.0	364.9	767.8
Cost per gasoline ($/(Gal/hr))	8,667	6,764	7,265	4,753	10,231
Cost per gasoline ($/Gal) (20 years)	0.050	0.039	0.042	0.028	0.059

4.3 The Third Stage

4.3.1 The Third Stage: Case Descriptions (Water Balance)

After running the cases in the second dimension, two major challenges are required to be addressed before marching farther. First, the water balance around the system and heater fuel required by processes. . The water balance is very important to be evaluated since the gasification system requires a great amount of water to fix the hydrogen to carbon ratio. Second, for every one mole of hydrocarbon being produced we expect one mole of water to be produced (Figure 13).

$$nCO + (2n + 1)H_2 \rightarrow C_nH_{2n+2} + nH_2O$$

Figure 13: Syngas to Hydrocarbon General Reaction

4.3.2 The Third Stage: Result (Water Balance)

Table 46 present the water balance for major units for MTG, DME (direct and indirect) and FT cases.

Table 46: Water Balance for Liquid Fuel Cases

MTG			
IN		Out	
Gasifier	446878	MeOH to H-C	858571
WGS	306827		
MeOH to H-C	220101		
Total	973806		858571
DME to H-C (Indirect)			
IN		Out	
Gasifier	446878	MeOH to DME	330636
WGS	306827	DME to H-C	427776
Total	753705		758413
DME to H-C (direct)			
IN		Out	
Gasifier	446878	DME to H-C	240315
Total	446878		240315
DME to FT			
IN		Out	
Gasifier	446878	FT to H-C	629325
WGS	306827		
Total	753705		629325

4.3.3 The Third Stage: Discussion (Water Balance)

MTG, DME to H-C (direct) and FT are processes that require water. Despite the fact that produced water from the system can be totally recycled (Direct and Indirect recycle), we anticipate to provide clean water to those processes in our calculation.

MeOH to H-C and gasifier units adapt well with the direct recycle water. On the other hand, clean water is WGS. The DME to H-C (indirect) is water sufficient if the direct and indirect water recycle were applied, (Table 47).

Table 47: Water Recycling for Liquid Fuel Cases

	MTG	DME to H-C (Indirect)	DME to H-C (direct)	FT
Net required	115235	0	124380	206562
Out	0	4708	0	0
Total recycle	858571	753705	629326	240316
Direct recycle	666979	446878	446878	240316
Indirect recycle	306827	306827	182448	0

4.3.4 The Third Stage: Case Descriptions and Result (Pinch Analysis)

To review the fuel requirement, MTG and DME (indirect) were sublimated using Aspen plus. The base for the design is National Renewable Energy Laboratory reports (Phillips, Tarud, Biddy, & Dutta, 2011) (Tan, et al., 2015). The gasifier was simulated using RGIBBS reactor. On the other hand, WGS was modeled using RStoic reactor. The remained of the simulation process was designed base in the NREL reports. To simplify the simulation code, the heat exchangers in the NREL reports design as cooling and heating unit. The full design of two processes with their streams is summarized in

Appendix B. After simulating the code, the Aspen Energy Analyzer Tool was used to evaluate stream data and estimate the utility requirement to the processes

In Table 48 and 49, a summary of the hot and cold streams for both processes.

Table 48: Hot and Cold Streams for DME Case

DME			
Cold Streams	T in (°F)	Tout (°F)	Q (MMBtu/hr)
DME+R_To_DME2	74	177	1.00E+09
HC5_To_HC4-1	44	78	1.56E+07
MEHRP_To_MEOHTOR	30	249	7.89E+08
SG6_To_SYGAS	121	270	2.24E+08
To Reboiler@T1_TO_HC7	184	207	7.34E+07
To Reboiler@T2_TO_GS1	201	202	1.05E+06
WGS_heat	109	200	5.54E+07
AA_heat	149	1200	2.06E+09
Hot streams	T in (°F)	Tout (°F)	Q (MMBtu/hr)
DHC1_To_DHC2	177	121	5.57E+08
DHC3_To_DHC4	121	43	1.09E+09
M4_To_M4-2	199	27	3.59E+08
MTGP_To_HC1	249	121	3.36E+08
MOEH_To_MEOH4	270	39	1.44E+09
PRODUCT_To_SG3-1	1200	200	1.66E+09
To Condenser@T2_TO_OH2	147	107	5.99E+06
To	48	36	2.74E+06
FD3_heat	43	43	5.29E+07
SH1_heat	39	30	3.78E+07
DMETOHC_heat	177	176	8.07E+08
MEOH_heat	270	270	1.06E+09
DL_heat	27	26	7.19E-01
MEOHSYSN_heat	249	248	1.55E+08

Table 49: Hot and Cold Streams for MTG Case

MTG			
Cold Streams	T in (°F)	Tout (°F)	Q (MMBtu/hr)
MEOHPU_To_MEOHTOR	30	310	848679495.4
SG6_To_SYGAS	121	270	224254919.7
HC5_To_HC4-1	16	43	14755908.78
To Reboiler@T1_TO_HC7	162	201	111673161.6
To Reboiler@T2_TO_GS1	180	193	18355067.28
Hot streams	T in (°F)	Tout (°F)	Q (MMBtu/hr)
MOEH_To_MEOH4	270	39	1524759633
PRODUCT_To_SG44	1200	200	1655602264
MTGP_To_HC1	310	22	1003380902
M4_To_M4-2	199	27	359462235.4
To Condenser@T2_TO_OH2	112	80	33733452.96
To	77	39	14874743.63
SH1_heat	39	30	39601526.49
MEOH_heat	270	270	1028965484
DL_heat	27	26	0.719104986

The Composite curves and grand composite curves for both cases are in Figures 14- 17.

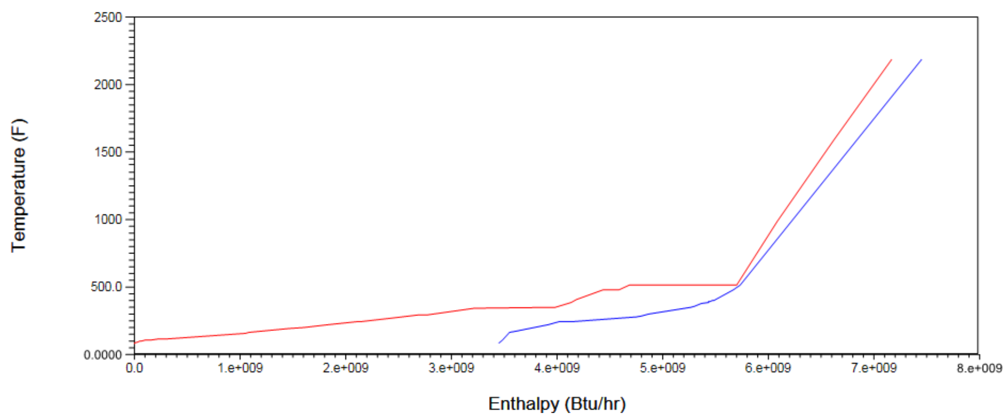

Figure 14: Composite Curves for DME Case

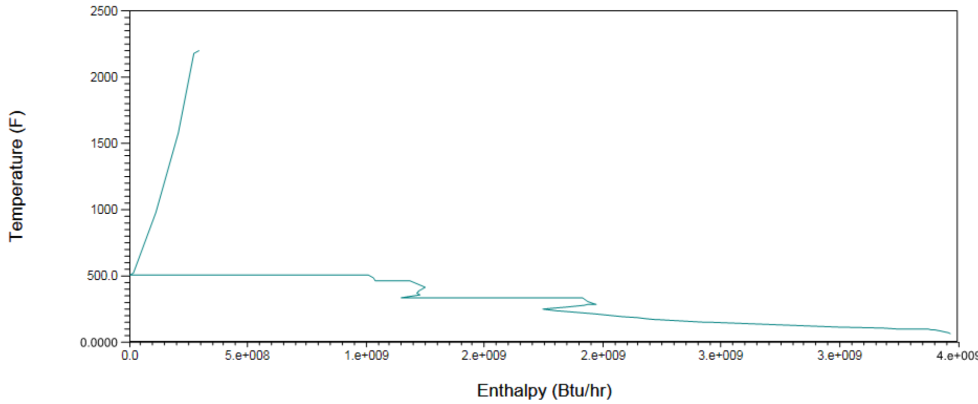

Figure 15: Grand Composite Curves for DME Case

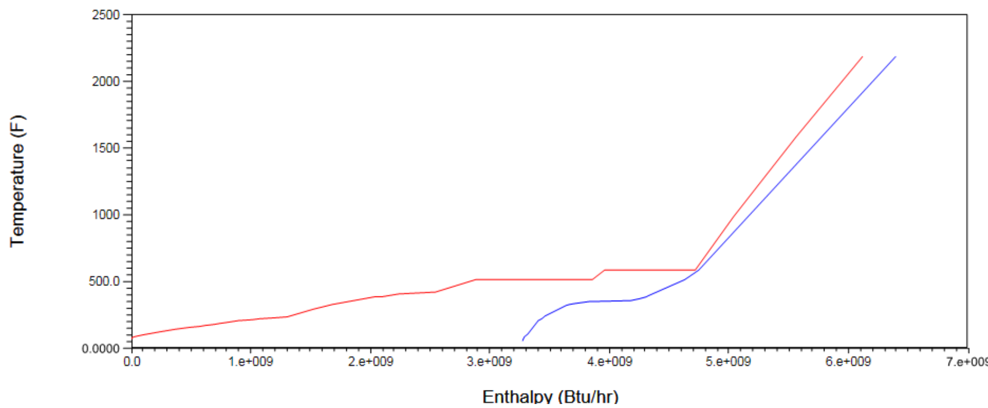

Figure 16: Composite Curves for MTG Case

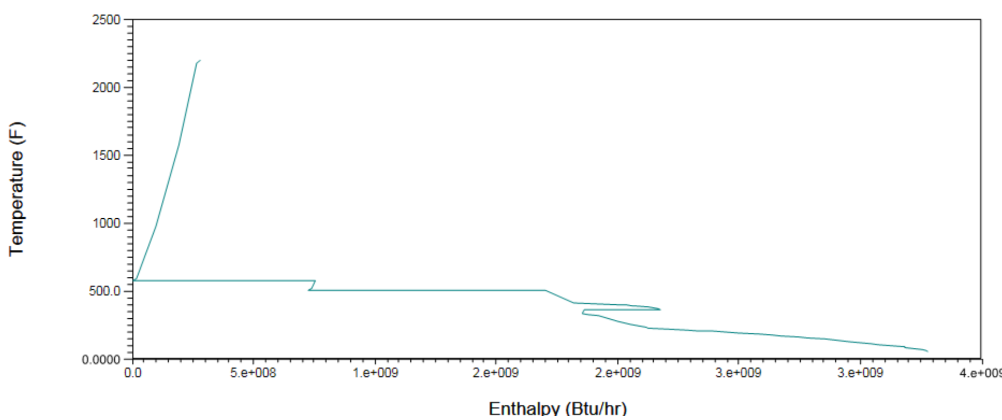

Figure 17: Grand Composite Curves for MTG Case

4.3.5 The Third Stage: Discussion (Pinch Analysis)

At first sight, the pinch point is observed at 500 °F for DME case and 520 °F. This gave us an indication that the system can be used to produce MP and LP steam. On top of everything, both problems consist only of two streams above the pinch. Hence, this alluded that hot utilities could be required only for the gasifier. Moreover, the total fuel required (total hot utility) for both cases is underneath the base case by 25% for MTG and 22% for DME (indirect) at ΔT=20°F, and it can reach to 32% for MTG and 20% for DME (indirect) at ΔT=10°F (Table 50).

Table 50: Fuel Demand Comparison

	ΔT(°F)	Q (MMBtu/h)	Change
Base case		442	0
MTG	20	329.9	-25%
	10	298.6	-32%
DME to G (Indirect)	20	343.4	-22%
	10	308.8	-30%

CHAPTER V

CONCLUSION, RECOMMENDATIONS AND FUTURE WORK

5.1 Conclusion

The application of introducing the heavy oil to the refinery will yield an increase in the production of the bottom products (vacuum residue, gas oil and diesel). It will also reduce the production of the light products (gases, LPG and naphtha) from ADU/VDU for oil with API above 20. However, we showed that if the heavy oil is below 20 API, the vacuum residue is will be the only increasing product. This also reflects on the unit capital cost. The operating cost of units handling heavy products will raise while the operating cost of units that process lighter products will be reduced

The power and steam required by the refinery should also increase as crude oil becomes heavier due to the high amount of steam used in the delayed coker unit. Nevertheless, the fuel for fire heater does not show the expected change as compared to the model. This unexpected result is mainly attributed to the fact that both the middle cracking units and heavy cracking units consume high levels of fuel. Therefore, any reduction of the feed to middle cracking units will accumulate in heavy cracking units which still demands more fuel to complete processing.

The cut of delayed coker unit reduces the production of fuel. Therefore, the fuel gasification paths (MTG, DME (direct and indirect) and FT) are more valuables than others gasification paths. All fuel paths showed the similar amount of fuel production yielding extra production around 100,000 lb/hr compared to the base case. This is due to the utilization of coke. This slight change is mainly caused by the chemical distribution of the products (gases LPG, Gasoline and Diesel). Therefore, it is expected to have a production increase from the base case by around 50% of the coke-out from the delayed coke, while the other 50 % is used to increment the H2/CO ratio.

The direct path of DME provided the lowest estimated cost compared to other fuel gasification paths. The MTG path and indirect DME path share similar cost. Among the most important finding of this work, the cost for MTG and DME paths came below the reduced price of the base case while the Fischer Tropsch showed higher price. The comparison base on the cost required to produce one gallon of gasoline for plant with 20 years lifetime projection.

Gasification paths are profoundly ran on the water, generally the gasifier and WGS. However, more than 95 % is recovered for the fuel gasification paths as a result of the recycling (both direct and indirect).

MTG and DME-indirect paths demand less fuel when compared to the base case. The pinch point is above 520 °F, hence, they serve as a good source for MP and LP steam.

5.2 Recommendations and Future Work

Initially, the fuel gasification paths showed a capable future based on our model. However, an important issue must be addressed in future work that involves the scenario when the gasification is more economic than base case (delayed coke) in refineries. As the increasing of hydrocarbon production is from gasified the heavy products, coke.

The block for syngas peroration (gasifier unit, air separation unit and water gas shifting unit) is one of the main area could potentially use to enhance the operational cost and capital cost of the gasification system. On the grounds that the block for syngas peroration has many parameters that could be varied (fuel in, steam to the gasifier, steam to WGS, oxygen to the gasifier, energy in or out and H/C ratio) (Figure 18). Moreover, the block benefits from fuel gas in providing energy to gasifier but it can be used as a hydrogen source to enhance the H2/CO ratio. Furthermore, the gasifier can be operated either exothermal or endodermal. It confides in the balance between water, heavy oil and oxygen.

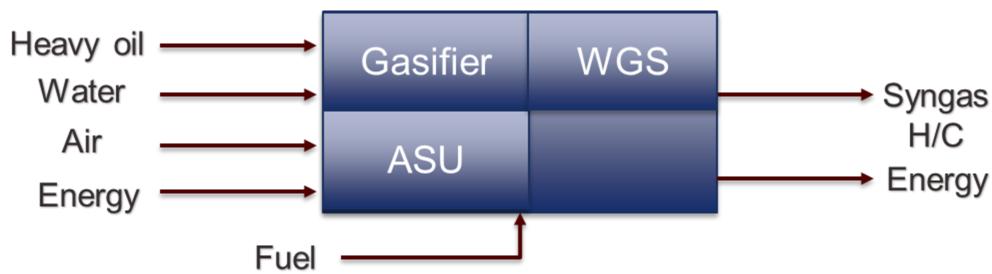

Figure 18: The Syngas Peroration Block

Due to the system extensive size, it requires at least two or three treys to accommodate the feed. Incorporating scenarios that involve different blended products is worth considering and investigating. In addition, the methanol and DME are considered a highly demanded chemical commodity which is easily sold in the market.

Heat integration, mass integration and power calculation techniques can be applied between the refinery and gasification section successfully. Yet, more detailed estimation are required for the capital cost as the top level calculations showed promising results.

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APPENDIX A

REFINERY BALANCE

Table 51: Refinery Balance for Gudrun Blend Well (API=50.7)

GUDRUN BLEND 2014 11		
	Mass flow	API
BPD	lb/day	
100000	27203804	50.7

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	8.7	8.7
Light straight run naphtha	158	20.3	20.3
Naphtha	356	54.7	54.7
Kerosene	464	64.7	64.7
Light diesel	554	72.4	72.4
Heavy diesel	644	79.6	79.6
Atm. Gas oil	698	82.8	82.8
Vacuum gas oil	950	95.2	95.2

Cost	MM$
ADU/VDU System	127
DC	72
FCC	127
HC	116
CCR	91
ISO	11
ALK	19

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	5.95E+03	6.75E+04	6.69E+04
Steam, lb/day		9.70E+04	1.98E+05	6.56E+04	-3.35E+05
Cooling water, gal/day		1.46E+07	2.98E+06	1.91E+02	5.58E+06
Fuel MMBtu/day		4.85E+03	5.95E+02	5.49E+02	1.12E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	1.46E+05	1.12E+05	1.41E+04	4.79E+03	5.06E+05
Steam, lb/day	8.38E+05	1.12E+06	0.00E+00	2.59E+05	2.24E+06
Cooling water, gal/day	5.03E+06	1.49E+07	1.13E+07	4.79E+06	5.91E+07
Fuel MMBtu/day	2.27E+03	1.12E+04	2.82E+03	0.00E+00	2.34E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	190569	0	0	0	0	50864	
C2	2351	298911	0	0	0	0	29501	
Gas	0	0	0	0	167268	0	2602	
total	2351	489480	0	0	167268	0	82967	742067
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	47934	450150	0	0	47298	0	8068	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	129176	0	8068	
IC4	0	0	337794	0	0	0	3470	
NC4	853013	305722	21506	0	58885	227086	8032	
C4=	0	0	0	0	0	0	0	
IC5	867518	0	0	0	0	0	0	
NC5	0	0	12284	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	1768465	755872	371583	0	235359	227086	27638	3386003
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	2828157	6834316	306389	0	1801204	0	0	11770067
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	2835184	0	0	0	0	3019830	0	5855014
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	2267647	0	26080	0	556656	0	640503	3490885
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	2619	0	160843	187411	0	350873
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	1677956	0	0	0	1695	0	1679650

Table 52: Refinery Balance for Ormen Lange Well (API=60.35)

Ormen Lange 201005		
	Mass flow	API
BPD	lb/day	
100000	25838915	60.35

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	7.2	7.2
Light straight run naphtha	158	19.5	19.5
Naphtha	356	73.6	73.6
Kerosene	464	88.8	88.8
Light diesel	554	95.8	95.8
Heavy diesel	644	98.8	98.8
Atm. Gas oil	698	99.5	99.5
Vacuum gas oil	950	100.0	100.0

Cost	MM$
ADU/VDU System	100
DC	0
FCC	86
HC	43
CCR	112
ISO	11
ALK	14

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	3.57E+02	0.00E+00	2.18E+04
Steam, lb/day		9.70E+04	1.19E+04	0.00E+00	-1.09E+05
Cooling water, gal/day		1.46E+07	1.78E+05	0.00E+00	1.82E+06
Fuel MMBtu/day		4.85E+03	3.57E+01	0.00E+00	3.64E+02
	HC	CCR	ISO	ALK	Total
Power, kWh	5.97E+03	1.60E+05	1.37E+04	1.49E+03	2.92E+05
Steam, lb/day	3.42E+04	1.60E+06	0.00E+00	8.04E+04	1.71E+06
Cooling water, gal/day	2.05E+05	2.13E+07	1.10E+07	1.49E+06	5.05E+07
Fuel MMBtu/day	9.30E+01	1.60E+04	2.74E+03	0.00E+00	2.41E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	279929	0	0	0	0	0	
C2	2343	438470	0	0	0	0	0	
Gas	0	0	0	0	53794	0	0	
total	2343	718399	0	0	53794	0	0	774536
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	183383	660074	0	0	15440	0	0	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	42170	0	0	
IC4	0	0	535932	0	0	0	0	
NC4	489189	448048	6684	0	19223	14137	0	
C4=	0	0	0	0	0	0	0	
IC5	705298	0	0	0	0	0	0	
NC5	0	0	3818	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	1377870	1108123	546434	0	76833	14137	0	3123398
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	2913564	9882095	95229	0	583909	0	0	13474798
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	4305981	0	0	0	0	188003	0	4493984
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	2062153	0	8106	0	176973	0	0	2247232
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	814	0	46615	0	0	47429
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	2430122	0	0	0	0	0	2430122

Table 53: Refinery Balance for Oseberg Well (API=38.0)

OSEBERG 201308		
BPD	Mass flow lb/day	API
100000	29165103	38

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	3.1	3.1
Light straight run naphtha	158	7.4	7.4
Naphtha	356	29.9	29.9
Kerosene	464	41.3	41.3
Light diesel	554	51.5	51.5
Heavy diesel	644	61.0	61.0
Atm. Gas oil	698	65.8	65.8
Vacuum gas oil	950	84.9	84.9

Cost	MM$
ADU/VDU System	138
DC	122
FCC	147
HC	162
CCR	82
ISO	11
ALK	23

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	1.13E+04	2.13E+05	1.02E+05
					-
Steam, lb/day		9.70E+04	3.78E+05	2.07E+05	5.11E+05
Cooling water, gal/day		1.46E+07	5.67E+06	6.20E+02	8.51E+06
Fuel MMBtu/day		4.85E+03	1.13E+03	1.79E+03	1.70E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	2.28E+05	9.19E+04	6.63E+03	7.55E+03	7.50E+05
Steam, lb/day	1.32E+06	9.19E+05	0.00E+00	4.08E+05	2.81E+06
Cooling water, gal/day	7.90E+06	1.22E+07	5.30E+06	7.55E+06	6.17E+07
Fuel MMBtu/day	3.51E+03	9.19E+03	1.33E+03	0.00E+00	2.35E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	140137	0	0	0	0	160591	
C2	6835	220350	0	0	0	0	93144	
Gas	0	0	0	0	258748	0	8216	
total	6835	360487	0	0	258748	0	261951	888021
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	122511	332059	0	0	72216	0	25473	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	197230	0	25473	
IC4	0	0	120081	0	0	0	10957	
NC4	260868	225740	33902	0	89907	350585	25359	
C4=	0	0	10519	0	0	0	0	
IC5	217199	0	0	0	0	0	0	
NC5	0	0	19364	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	600578	557799	183866	0	359353	350585	87261	2139442
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	1237360	5166320	482995	0	2766640	0	0	9653315
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	3217616	0	0	0	0	4662150	0	7879767
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	3020660	0	41112	0	870010	0	2022245	5954027
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	4129	0	269983	591708	0	865819
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	1264955	0	0	0	5350	0	1270305

Table 54: Refinery Balance for Kissanje Well (API=31.0)

KISSANJE201106		
BPD	Mass flow lb/day	API
100000	30543619	31

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	1.5	1.5
Light straight run naphtha	158	3.6	3.6
Naphtha	356	17.7	17.7
Kerosene	464	28.0	28.0
Light diesel	554	36.9	36.9
Heavy diesel	644	45.7	45.7
Atm. Gas oil	698	50.8	50.8
Vacuum gas oil	950	73.3	73.3

Cost	MM$
ADU/VDU System	145
DC	183
FCC	152
HC	185
CCR	73
ISO	11
ALK	24

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	1.58E+04	4.76E+05	1.14E+05
Steam, lb/day		9.70E+04	5.27E+05	4.63E+05	-5.69E+05
Cooling water, gal/day		1.46E+07	7.90E+06	1.14E+03	9.48E+06
Fuel MMBtu/day		4.85E+03	1.58E+03	3.29E+03	1.90E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	2.69E+05	7.53E+04	5.18E+03	7.87E+03	1.05E+06
Steam, lb/day	1.56E+06	7.53E+05	0.00E+00	4.26E+05	3.26E+06
Cooling water, gal/day	9.37E+06	1.00E+07	4.14E+06	7.87E+06	6.33E+07
Fuel MMBtu/day	4.11E+03	7.53E+03	1.04E+03	0.00E+00	2.43E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	104960	0	0	0	0	301118	
C2	2365	165678	0	0	0	0	174651	
Gas	0	0	0	0	291931	0	15406	
total	2365	270638	0	0	291931	0	491174	1056108
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	73226	249931	0	0	80384	0	47763	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	219538	0	47763	
IC4	0	0	1160	0	0	0	20544	
NC4	94828	170167	35368	0	100076	409115	47549	
C4=	0	0	94839	0	0	0	0	
IC5	136777	0	0	0	0	0	0	
NC5	0	0	20202	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	304831	420098	151569	0	399998	409115	163619	1849231
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	912089	4035740	503885	0	3098319	0	0	8550033
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	2916957	0	0	0	0	5440495	0	8357452
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	2644816	0	42890	0	992346	0	3485766	7165818
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	4307	0	328976	1321528	0	1654811
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	984138	0	0	0	10032	0	994171

Table 55: Refinery Balance for Girassol Well (API=29.8)

GIRASSOL200906		
BPD	Mass flow lb/day	API
100000	30733263	29.8

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	0.9	0.9
Light straight run naphtha	158	2.1	2.1
Naphtha	356	14.1	14.1
Kerosene	464	23.5	23.5
Light diesel	554	33.2	33.2
Heavy diesel	644	43.5	43.5
Atm. Gas oil	698	49.3	49.3
Vacuum gas oil	950	72.4	72.4

Cost	MM$
ADU/VDU System	146
DC	191
FCC	158
HC	192
CCR	69
ISO	11
ALK	24

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	1.64E+04	4.57E+05	1.28E+05
Steam, lb/day		9.70E+04	5.48E+05	4.44E+05	-6.38E+05
Cooling water, gal/day		1.46E+07	8.22E+06	1.21E+03	1.06E+07
Fuel MMBtu/day		4.85E+03	1.64E+03	3.49E+03	2.13E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	2.78E+05	6.81E+04	4.09E+03	8.17E+03	1.05E+06
Steam, lb/day	1.62E+06	6.81E+05	0.00E+00	4.42E+05	3.19E+06
Cooling water, gal/day	9.70E+06	9.08E+06	3.27E+06	8.17E+06	6.36E+07
Fuel MMBtu/day	4.23E+03	6.81E+03	8.18E+02	0.00E+00	2.40E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	94855	0	0	0	0	305704	
C2	233	149877	0	0	0	0	177311	
Gas	0	0	0	0	326902	0	15640	
total	233	244732	0	0	326902	0	498655	1070522
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	14046	226154	0	0	90245	0	48490	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	246469	0	48490	
IC4	0	0	0	0	0	0	20857	
NC4	79064	154039	36720	0	112353	423838	48273	
C4=	0	0	120722	0	0	0	0	
IC5	75004	0	0	0	0	0	0	
NC5	0	0	20974	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	168114	380194	178417	0	449066	423838	166111	1765740
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	714817	3685954	523155	0	3474417	0	0	8398344
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	2662305	0	0	0	0	5636279	0	8298584
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	2877900	0	44531	0	1108899	0	3645025	7676354
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	4472	0	363226	1268102	0	1635801
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	897948	0	0	0	10185	0	908133

Table 56: Refinery Balance for Pazflor Well (API=25.2)

PAZFLOR 2012 02		
BPD	Mass flow lb/day	API
100000	31621585	25.2

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	0.5	0.5
Light straight run naphtha	158	1.4	1.4
Naphtha	356	9.3	9.3
Kerosene	464	17.6	17.6
Light diesel	554	27.4	27.4
Heavy diesel	644	38.1	38.1
Atm. Gas oil	698	44.1	44.1
Vacuum gas oil	950	67.9	67.9

Cost	MM$
ADU/VDU System	148
DC	217
FCC	161
HC	197
CCR	64
ISO	11
ALK	24

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	1.80E+04	7.12E+05	1.35E+05
Steam, lb/day		9.70E+04	6.00E+05	6.92E+05	-6.74E+05
Cooling water, gal/day		1.46E+07	9.01E+06	1.43E+03	1.12E+07
Fuel MMBtu/day		4.85E+03	1.80E+03	4.12E+03	2.25E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	2.88E+05	6.02E+04	4.41E+03	7.89E+03	1.31E+06
Steam, lb/day	1.67E+06	6.02E+05	0.00E+00	4.27E+05	3.41E+06
Cooling water, gal/day	1.00E+07	8.02E+06	3.53E+06	7.89E+06	6.42E+07
Fuel MMBtu/day	4.37E+03	6.02E+03	8.82E+02	0.00E+00	2.43E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	78286	0	0	0	0	383005	
C2	4901	124150	0	0	0	0	222147	
Gas	0	0	0	0	345505	0	19595	
total	4901	202436	0	0	345505	0	624747	1177588
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	36402	187517	0	0	95254	0	60752	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	260151	0	60752	
IC4	0	0	0	0	0	0	26131	
NC4	16005	127905	35454	0	118590	439658	60480	
C4=	0	0	157847	0	0	0	0	
IC5	29103	0	0	0	0	0	0	
NC5	0	0	20250	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	81509	315422	213551	0	473996	439658	208115	1732250
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	754431	3159792	505107	0	3669453	0	0	8088783
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	2343543	0	0	0	0	5829751	0	8173294
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	2890500	0	42994	0	1173262	0	4004386	8111143
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	4318	0	386700	1978329	0	2369346
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	767082	0	0	0	12761	0	779843

Table 57: Refinery Balance for Grane Well (API=20.3)

GRANE		
	Mass flow	API
BPD	lb/day	
100000	32655831	20.3

Oil Cuts	Temp (°F)	W fre (%)	V fre (%)
Off gas	90	0.5	0.5
Light straight run naphtha	158	1.1	1.1
Naphtha	356	5.4	5.4
Kerosene	464	11.7	11.7
Light diesel	554	21.2	21.2
Heavy diesel	644	31.9	31.9
Atm. Gas oil	698	38.2	38.2
Vacuum gas oil	950	66.4	66.4

Cost	MM$
ADU/VDU System	151
DC	225
FCC	162
HC	226
CCR	59
ISO	11
ALK	24

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	1.98E+04	8.67E+05	1.36E+05
Steam, lb/day		9.70E+04	6.60E+05	8.43E+05	-6.82E+05
Cooling water, gal/day		1.46E+07	9.91E+06	1.50E+03	1.14E+07
Fuel MMBtu/day		4.85E+03	1.98E+03	4.33E+03	2.27E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	3.42E+05	5.23E+04	4.33E+03	7.83E+03	1.52E+06
Steam, lb/day	1.99E+06	5.23E+05	0.00E+00	4.23E+05	3.85E+06
Cooling water, gal/day	1.19E+07	6.98E+06	3.46E+06	7.83E+06	6.60E+07
Fuel MMBtu/day	5.18E+03	5.23E+03	8.65E+02	0.00E+00	2.47E+04

	Units Products (lb/day)							
Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	64757	0	0	0	0	417783	
C2	35	103150	0	0	0	0	242318	
Gas	0	0	0	0	349648	0	21374	
total	35	167908	0	0	349648	0	681475	1199065
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	27817	155982	0	0	96422	0	66268	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	263340	0	66268	
IC4	0	0	0	0	0	0	28504	
NC4	36864	106578	35185	0	120044	513665	65971	
C4=	0	0	168979	0	0	0	0	
IC5	36049	0	0	0	0	0	0	
NC5	0	0	20097	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	100731	262560	224260	0	479805	513665	227012	1808033
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	715440	2731876	501276	0	3713992	0	0	7662584
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	1800246	0	0	0	0	6830821	0	8631067
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	2797398	0	42668	0	1187077	0	4006400	8033543
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	4285	0	390773	2408467	0	2803525
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	660606	0	0	0	13919	0	674525

Table 58: Refinery Balance for Peregrino Well (API=13.4)

PEREGRINO 2012 12			
		Mass flow	API
BPD		lb/day	
	100000	34207358	13.4

Oil Cuts	Temp (F)	W fre (%)	V fre (%)
Off gas	90	0.2	0.2
Light straight run naphtha	158	0.3	0.3
Naphtha	356	3.7	3.7
Kerosene	464	7.9	7.9
Light diesel	554	12.9	12.9
Heavy diesel	644	19.1	19.1
Atm. Gas oil	698	22.5	22.5
Vacuum gas oil	950	49.4	49.4

Cost	MM$
ADU/VDU System	155
DC	313
FCC	139
HC	216
CCR	68
ISO	11
ALK	22

Utilities	ADU	VDU	Desalter	DC	FCC
Power, kWh	2.00E+03	8.73E+04	2.35E+04	2.40E+06	8.69E+04
Steam, lb/day		9.70E+04	7.84E+05	2.34E+06	-4.35E+05
Cooling water, gal/day		1.46E+07	1.18E+07	2.26E+03	7.24E+06
Fuel MMBtu/day		4.85E+03	2.35E+03	6.50E+03	1.45E+03
	HC	CCR	ISO	ALK	Total
Power, kWh	3.28E+05	6.71E+04	1.26E+04	6.54E+03	3.02E+06
Steam, lb/day	1.92E+06	6.71E+05	0.00E+00	3.53E+05	5.72E+06
Cooling water, gal/day	1.15E+07	8.95E+06	1.01E+07	6.54E+06	8.74E+06
Fuel MMBtu/day	4.91E+03	6.71E+03	2.53E+03	0.00E+00	2.93E+04

Units Products (lb/day)

Gases	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	Total
C1	0	72929	0	0	0	0	859890	
C2	3	116783	0	0	0	0	498744	
Gas	0	0	0	0	224976	0	43993	
total	3	189712	0	0	224976	0	1402627	1817317
LPG	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
C3	1075	176844	0	0	61452	0	136395	
NC3	0	0	0	0	0	0	0	
C3=	0	0	0	0	167833	0	136395	
IC4	0	0	0	0	0	0	58668	
NC4	18890	121078	29366	0	76507	476480	135784	
C4=	0	0	157544	0	0	0	0	
IC5	13864	0	0	0	0	0	0	
NC5	0	0	16774	0	0	0	0	
LPG	0	0	0	0	0	0	0	
Total	33830	297922	203684	0	305792	476480	467241	1784949
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Gasoline	1101295	3236673	418384	0	2377037	0	0	7133388
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Kerosene	1187741	0	0	0	0	6336327	0	7524068
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Diesel	1479707	0	35612	0	769823	0	3284487	5569630
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
Coke	0	0	3576	0	264572	6679190	0	6947338
	ADUVDU	Reforming	ALK	CCR	FCC	HC	Delayed coker	
H2	0	779309	0	0	0	28649	0	807958

APPENDIX B

MTG AND DME (INDIRECT) STIMULATION CASES

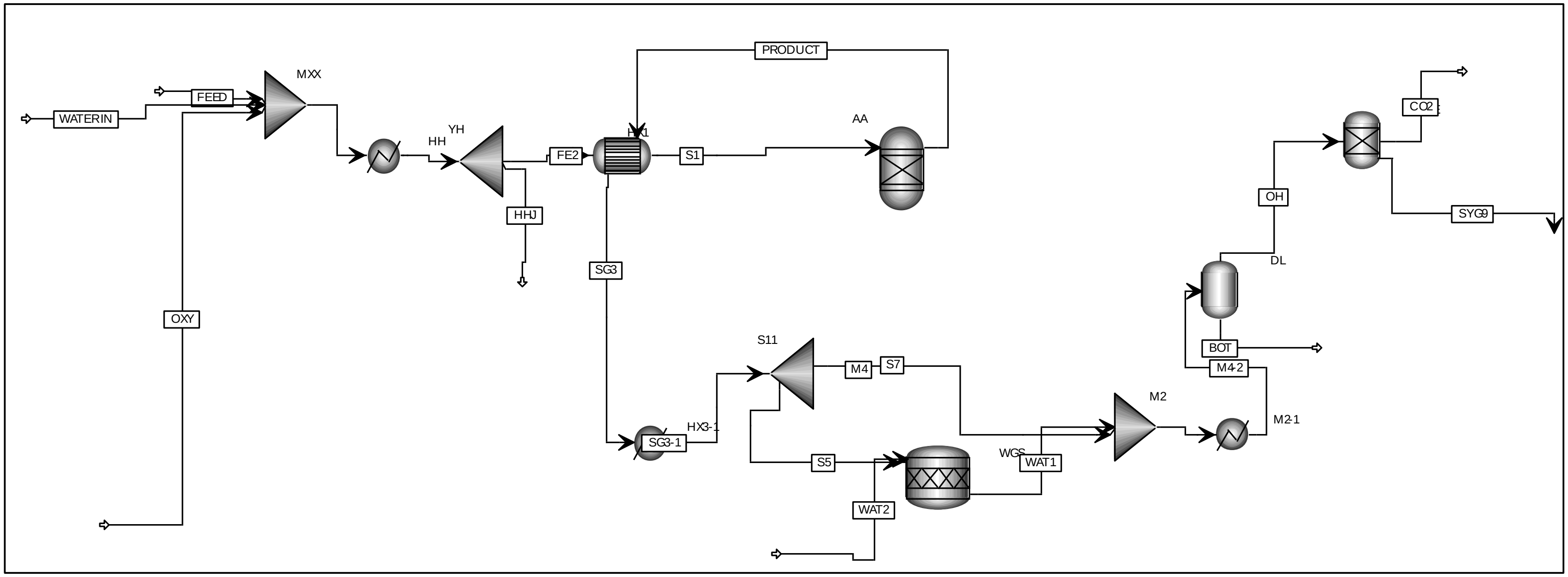

Figure 19: Methanol Simulation Sheet (Part I)

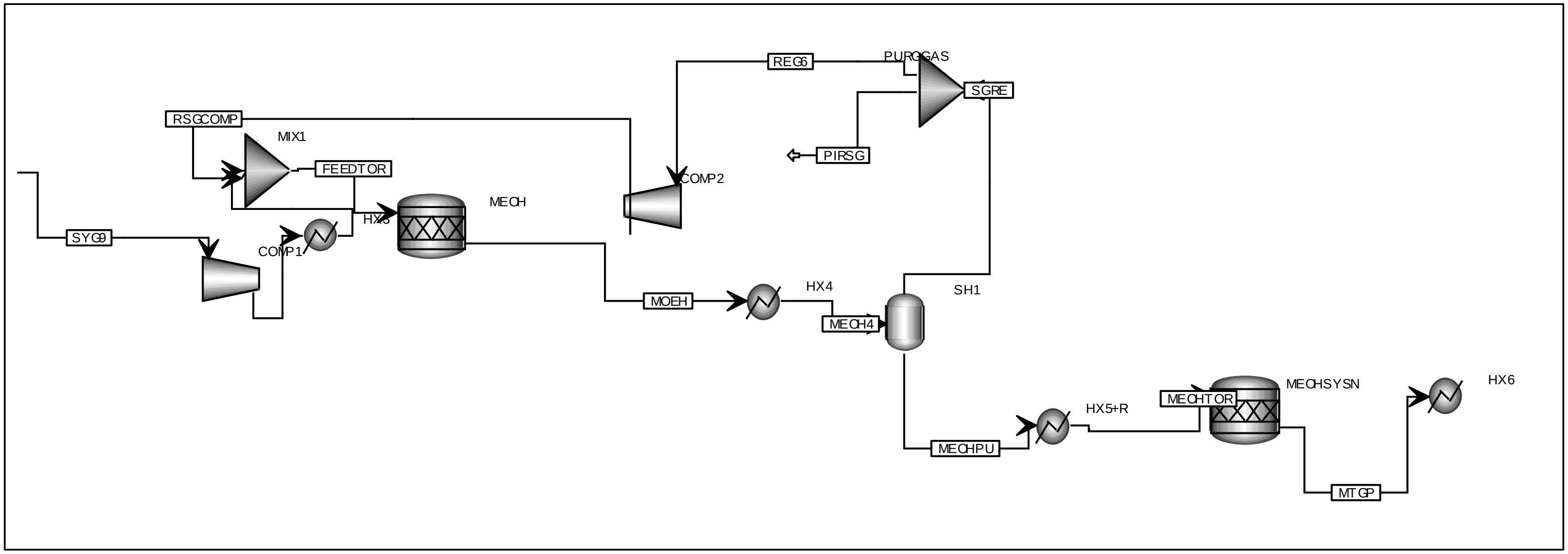

Figure 20: Methanol Simulation Sheet (Part II)

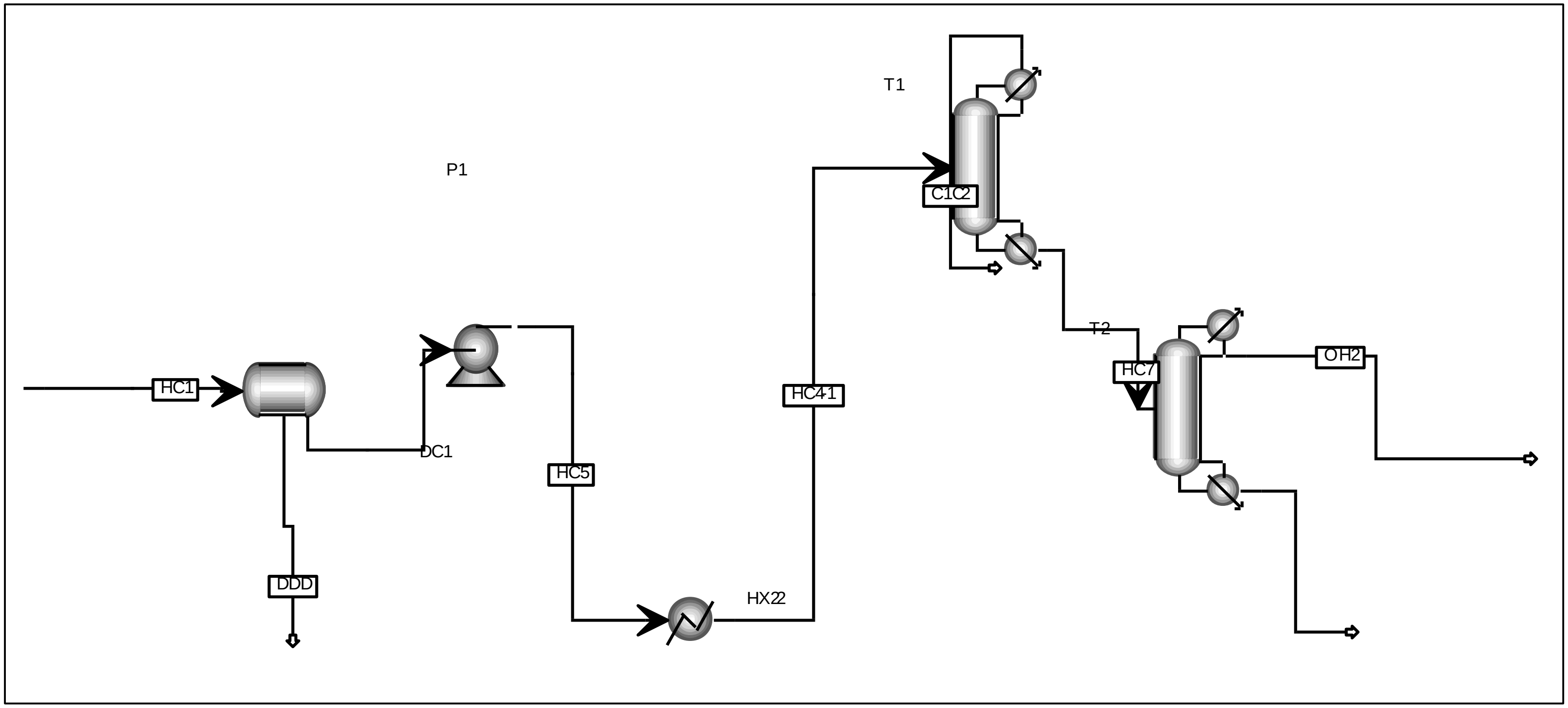

Figure 21: Methanol Simulation Sheet (Part III)

Table 59: Key Streams Data of Methanol Path (Part I)

Mass Flow lb/hr	Feed HV	Water in	Oxygen	Feed to Gasified	Out from Gasified
	FEED	WATERIN	OXY	S1	PRODUCT
WATER	0	446878	0	446878	28627
H2	0	0	0	0	109563
CO2	0	0	0	0	28751
CO	0	0	0	0	1528410
CH4	0	0	0	0	24759
O2	0	0	522472	522472	0
N2	0	0	0	0	0
S	0	0	0	0	0
H2S	0	0	0	0	0
COS	0	0	0	0	0
DIMET-01	0	0	0	0	0
C2	0	0	0	0	0
C3	0	0	0	0	0
IC4	0	0	0	0	0
C4H8	0	0	0	0	0
NC4	0	0	0	0	0
MEOH	0	0	0	0	0
C8	0	0	0	0	0
C3H6	0	0	0	0	0
1-OCT-01	0	0	0	0	0
BENZE-01	0	0	0	0	0
TOLUE-01	0	0	0	0	0
M-XYL-01	0	0	0	0	0
PC2414F	750755	0	0	750755	0
Total Flow lbmol/hr	489	24805	16328	41622	112701
Total Flow lb/hr	750755	446878	522472	1720110	1720110
Total Flow cuft/hr	11373	589409	361665	902712	9126380
Temperature F	300	446	300	341	2192
Pressure psia	367	367	367	367	353

Table 60: Key Streams Data of Methanol Path (Part II)

Mass Flow lb/hr	after WGS	Clean gas	MeOH out	H-C production
	M4	SYG9	MEOHPU	HC4-1
WATER	33752	3175	3174	589986
H2	146374	146374	55	55
CO2	832387	8324	4542	4542
CO	1016920	1016920	1044	1044
CH4	24759	24759	2758	10988
O2	0	0	0	0
N2	0	0	0	0
S	0	0	0	0
H2S	0	0	0	0
COS	0	0	0	0
DIMET-01	0	0	0	0
C2	0	0	0	914
C3	0	0	0	22851
IC4	0	0	0	31990
C4H8	0	0	0	4569
NC4	0	0	0	22850
MEOH	0	0	1110330	66620
C8	0	0	0	122235
C3H6	0	0	0	4569
1-OCT-01	0	0	0	130821
BENZE-01	0	0	0	0
TOLUE-01	0	0	0	29824
M-XYL-01	0	0	0	77852
PC2414F	0	0	0	0
Total Flow lbmol/hr	131246	110824	35168	40656
Total Flow lb/hr	2054200	1199560	1121900	1121710
Total Flow cuft/hr	3421590	1835860	22939	959935
Temperature F	391	80	86	590
Pressure psia	353	353	435	435

Table 61: Key Streams Data of Methanol Path (Part III)

Mass Flow lb/hr	Overhead 1 gases	Overhead 2 LPG	Gasoline
	C1C2	OH2	GS1
WATER	55	345	2336
H2	54	0	0
CO2	4279	195	0
CO	1042	1	0
CH4	10891	89	0
O2	0	0	0
N2	0	0	0
S	0	0	0
H2S	0	0	0
COS	0	0	0
DIMET-01	0	0	0
C2	828	85	1
C3	13098	8120	1633
IC4	5981	16826	9182
C4H8	623	2395	1551
NC4	2186	11794	8871
MEOH	207	11011	37391
C8	0	180	122055
C3H6	2865	1464	240
1-OCT-01	0	130	130691
BENZE-01	0	0	0
TOLUE-01	0	1	29823
M-XYL-01	0	0	77853
PC2414F	0	0	0
Total Flow lbmol/hr	1394	1132	4968
Total Flow lb/hr	42110	52637	421626
Total Flow cuft/hr	17941	1617	11863
Temperature F	102	175	379
Pressure psia	391	220	240

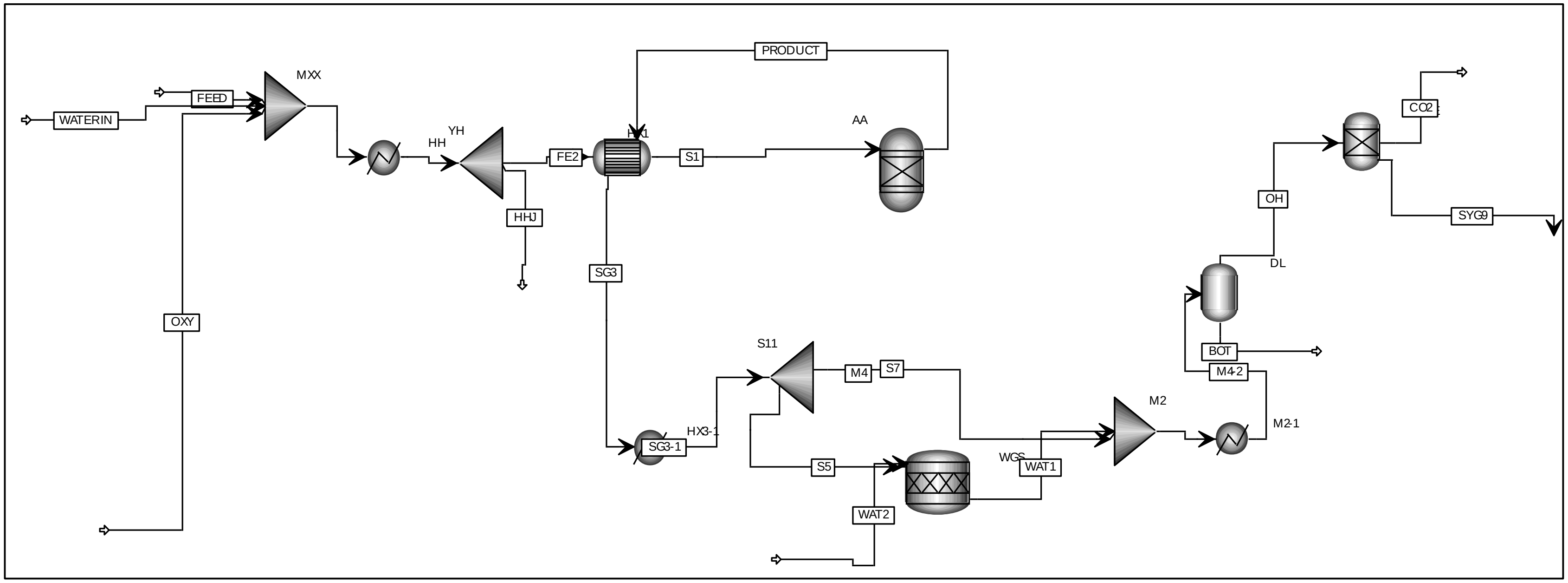

Figure 22: DME Simulation Sheet (Part I)

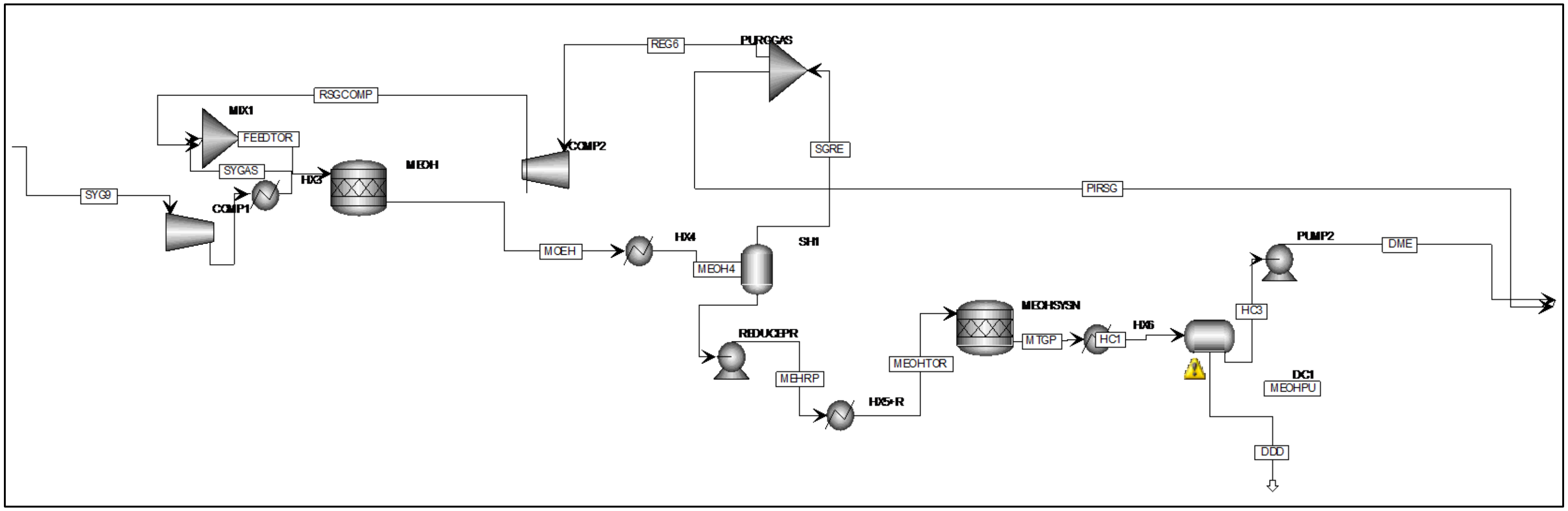

Figure 23: DME Simulation Sheet (Part II)

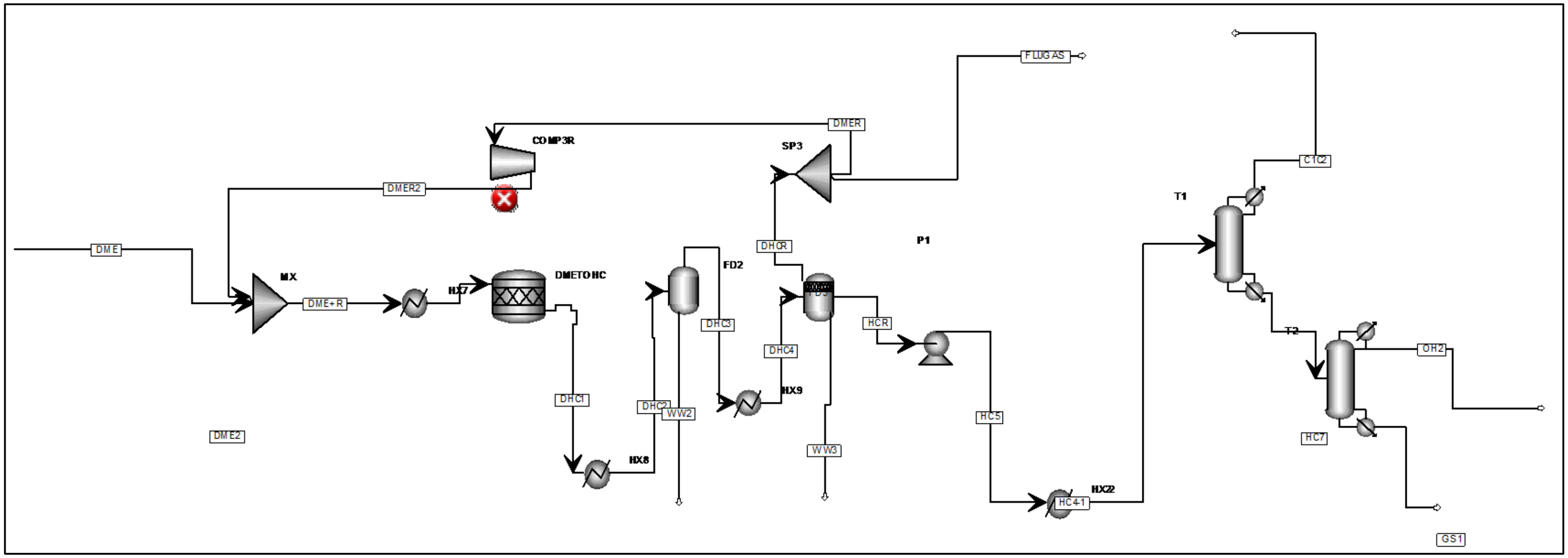

Figure 24: DME Simulation Sheet (Part III)

Table 62: Key Streams Data of Dme Path (Part I)

Mass Flow lb/hr	Feed HV	water in	Oxygen	Feed to Gasifier	Out from Gasifier
	FEED	WATERIN	OXY	S1	PRODUCT
WATER	0	446878	0	446878	28627
H2	0	0	0	0	109563
CO2	0	0	0	0	28751
CO	0	0	0	0	1528410
O2	0	0	522472	522472	0
COS	0	0	0	0	0
MEOH	0	0	0	0	0
DIMET-01	0	0	0	0	0
CH4	0	0	0	0	24759
C2	0	0	0	0	0
C2H4	0	0	0	0	0
C3	0	0	0	0	0
C3H6	0	0	0	0	0
IC4	0	0	0	0	0
NC4	0	0	0	0	0
C4H8	0	0	0	0	0
C5	0	0	0	0	0
C5H10	0	0	0	0	0
C6	0	0	0	0	0
C6H12	0	0	0	0	0
C7	0	0	0	0	0
C7H14	0	0	0	0	0
C8	0	0	0	0	0
C8H16	0	0	0	0	0
C9	0	0	0	0	0
C9H18	0	0	0	0	0
BENZE-01	0	0	0	0	0
TOLUE-01	0	0	0	0	0
M-XYL-01	0	0	0	0	0
PC2414F	7.51E+0	0	0	750755	0
Total Flow lbmol/hr	489.068	24805	16328	41622	112701
Total Flow lb/hr	7.51E+0	446878	522472	1720110	1720110
Temperature F	300	220	250	300	2192
Pressure psia	367.398	367	367	367	353

Table 63: Key Streams Data of DME Path (Part II)

Mass Flow lb/hr	after WGS	Clean gas	MeOH Production	DME
	M4	SYG9	MEOHPU	DME
WATER	33752	3175	3173	12467
H2	146374	146374	57	57
CO2	832387	8324	3941	3928
CO	1016920	1016920	1089	1089
O2	0	0	0	0
COS	0	0	0	0
MEOH	0	0	1089710	51408
DIMET-01	0	0	0	744148
CH4	24759	24759	2122	2121
C2	0	0	0	0
C2H4	0	0	0	0
C3	0	0	0	0
C3H6	0	0	0	0
IC4	0	0	0	0
NC4	0	0	0	0
C4H8	0	0	0	0
C5	0	0	0	0
C5H10	0	0	0	0
C6	0	0	0	0
C6H12	0	0	0	0
C7	0	0	0	0
C7H14	0	0	0	0
C8	0	0	0	0
C8H16	0	0	0	0
C9	0	0	0	0
C9H18	0	0	0	0
BENZE-01	0	0	0	0
TOLUE-01	0	0	0	0
M-XYL-01	0	0	0	0
PC2414F	0	0	0	0
Total Flow lbmol/hr	131246	110824	34474	18738
Total Flow lb/hr	2054200	1199560	1100100	815218
Temperature F	391	80	86	96
Pressure psia	353	353	435	129

Table 64: Key Streams Data of DME Path (Part III)

Mass Flow lb/hr	H-C production	Overhead 1 gases	Overhead 2 LPG	Gasoline
	HC4-1	C1C2	OH2	GS1
WATER	2248	0	529	1719
H2	0	0	0	0
CO2	502	502	0	0
CO	514	514	0	0
O2	0	0	0	0
MEOH	0	0	0	0
DIMET-01	6442	4794	541	1107
CH4	574	574	0	0
C2	129	129	0	0
C2H4	59	59	0	0
C3	1379	1378	0	1
C3H6	13	13	0	0
IC4	6828	253	1734	4840
NC4	108	0	26	82
C4H8	187	2	47	138
C5	9734	0	1427	8307
C5H10	6788	0	1077	5711
C6	20170	0	1454	18716
C6H12	4628	0	396	4233
C7	144375	0	2616	141759
C7H14	32069	0	917	31152
C8	185167	0	1	185166
C8H16	7226	0	0	7226
C9	3745	0	0	3745
C9H18	925	0	0	925
BENZE-01	0	0	0	0
TOLUE-01	0	0	0	0
M-XYL-01	0	0	0	0
PC2414F	0	0	0	0
Total Flow lbmol/hr	4501	212	164	4125
Total Flow lb/hr	433812	8220	10765	414828
Temperature F	173	96	224	396
Pressure psia	312	200	152	172