TECHNO-ECONOMIC ASSESSMENT OF FISCHER-TROPSCH AND DIRECT METHANE TO METHANOL PROCESSES IN MODULAR GTL TECHNOLOGIES

by

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In 2014, about 3.5% of the global gas production was flared, of which 0.289 TCF were in the US alone. This increase of natural gas flaring in the US has been exacerbated by the drilling and fracking activities in the shale gas plays. Improper flaring of natural gas leads to emissions of methane and other organic volatile compounds, sulfur oxides (SO_X) and carbon dioxide (CO_2). In fact, by 2020 the total gas volume flared is projected to be up to 60% greater than that in 2000, which is problematic. Thus, there is a great and pressing need for curbing or eliminating the flared natural gas and fugitive methane from remote reservoirs in order to protect the environment and avoid global warming.

This study aims at investigating the potential use of the Fischer-Tropsch (F-T) synthesis process, in a microchannel reactor (MCR), and the Direct Methane to Methanol (DMTM) process in a compact plant footprint for curbing or eliminating natural gas flaring. The two processes were modeled using the process simulator Aspen HYSYS v7.2 and their operational and economic performances were evaluated in terms of the products yield, net present value (NPV), payback period (PBP) and internal rate of return (IRR). In addition, the effects of tailgas and methane recycle ratios on these process performances are investigated.

The simulation results showed that the unit cost of the DMTM process was very sensitive to the methane recycle ratio, however, that of the F-T in MCR was less sensitive to the tail gas recycle ratios. In order to maintain an IRR > 10%, which is the minimal acceptable value, the tail gas recycle ratio for the F-T in MCR had to be greater than 8 and 30%, at CO conversions of 80% and 72%, respectively, whereas for the DMTM process, a minimum methane recycle ratio of 60% was required to achieve any profitability. In addition, the DMTM process appeared to have significantly higher net energy requirements per product yield when compared with those of the F-T in MCR process; however, both processes had higher energy requirements than those of conventional GTL technologies.

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PREFACE

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1.0 INTRODUCTION

At the end of 2014, the total worldwide proved crude oil reserve was 1700.1 Bbbl (billion barrel), where the US share was 49.3 Bbbl, representing 2.9%; and the total worldwide proved natural gas reserve was 6,606.4 TCF (trillion cubic feet), where the US share was 345.0 TCF, representing 5.2% [1], as shown in Figure 1-1. Natural gas can be produced worldwide from conventional and unconventional gas reservoirs, coal beds, and shale gas formation and from oil reservoirs as an associated gas. Table 1-1 shows the US natural gas activities over 5 years [2].

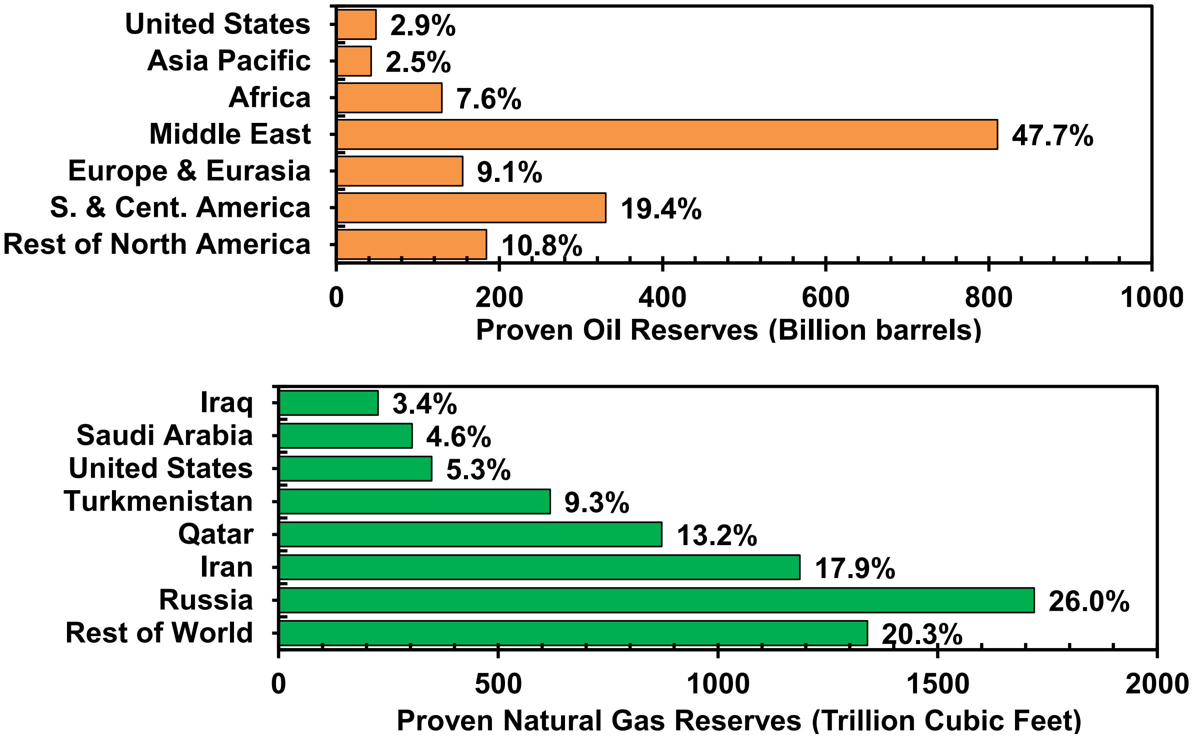


Figure 1-1: Countries with major proven oil and natural gas reserves [1]

Table 1-1: US natural gas activities over 5 years [2]

Annual-Brillion Cubic Feet	2010	2011	2012	2013	2014	2015
Gross Production	26,816	28,479	29,542	29,523	31,346	32,965
From Gas Wells	13,247	12,291	12,504	10,760	10,384	0
From Oil Wells	5,835	5,908	4,966	5,405	5,922	0
From Shale Gas Wells	5,817	8,501	10,533	11,933	13,754	0
From Coalbed Wells	1,917	1,779	1,539	1,426	1,285	0
Vented and Flared	166	209	213	260	289	0
Nonhydrocarbon Gases Removed	837	868	769	368	401	0
Dry Production	21,316	22,902	24,033	24,206	25,728	27,096

If the gas is produced from remote or stranded areas, lacking adequate gas transportation infrastructure, huge amounts of this gas are flared as a common practice in oil and gas industries in order to protect the plants downstream equipment from over-pressurization or explosion. Actually, gas flares can be found worldwide practically in all onshore and offshore gas and oil fields as well as processing facilities.

In 2012, a survey of global gas flaring found a total of 7467 flare sites worldwide, of which 6802 were upstream flares (production facilities), 628 were downstream sites (predominantly refineries) and 37 flares were found at Liquefied Natural Gas (LNG) terminals [3]. The flare sites were mostly in the US (2399), Russia (1053), Canada (332), Nigeria (325) and China (309). The estimated flared gas volume was 5.05 TCF. The upstream flaring was estimated at 4.57 TCF, downstream flaring at 0.38 TCF and flaring at LNG liquefaction plants was estimated at 0.11 TCF [3]. This flared gas volume represented about 3.5% of total worldwide natural gas consumption and 19.8% of US natural gas consumption [2]. If used to fuel vehicles in the US, it could power 74-million automobiles based on an average of 25 miles/gallon of gasoline [4] and 13,476 miles/year [5].

Improper flaring of natural and associated gas leads to undesirable emissions of methane and other organic volatile compounds, including aromatic hydrocarbons (benzene, toluene, and xylene) and benzapyrene, which are known to be carcinogenic, in addition to sulfur oxides (SO_X) and carbon dioxide (CO_2). Methane, as the main component of the natural gas, has disastrous and detrimental impact on the global climate change, since it is a harmful Greenhouse Gas (GHG), with a global warming potential of 72 times that of CO_2 over a 20 year period, as reported by the Intergovernmental Panel on Climate Change (IPCC) [6]. The flaring data in 2010 indicted that the amount of flared associated gas from oil drilling sites, coupled with fossil fuel combustion and cement production led to tripling CO_2 emissions (1300 ± 110 Gt CO_2) when compared with the last recorded value in 1750-1970, which was (420 ± 35 Gt CO_2) [7]. As such, 2,400 Million tons of CO_2 were emitted annually, which accounted of about 1.2% of the worldwide emissions. That may seem to be insignificant, but in perspective, it is more than half of the Certified Emissions Reductions, which was issued under the rules and mechanisms of the Kyoto Protocol as of June 2011 [8]. In 2014, it was reported that between 3.9 - 4.9 TCF of natural gas, which is equivalent to 3.5% of the global gas production [9], was flared, of which 289 BCF was flared in the US alone. This increase in gas flaring in the US has been aggravated over the past decade by the increased drilling and fracking activities in the shale gas plays, and the absence of a local natural gas market, as can be observed in Figure 1-2 [10], which shows the significant number of flares present in the Bakken Shale Play. Howarth et al. [11] mentioned that methane emission are usually 40% higher for unconventional shale compared to conventional gas fields.

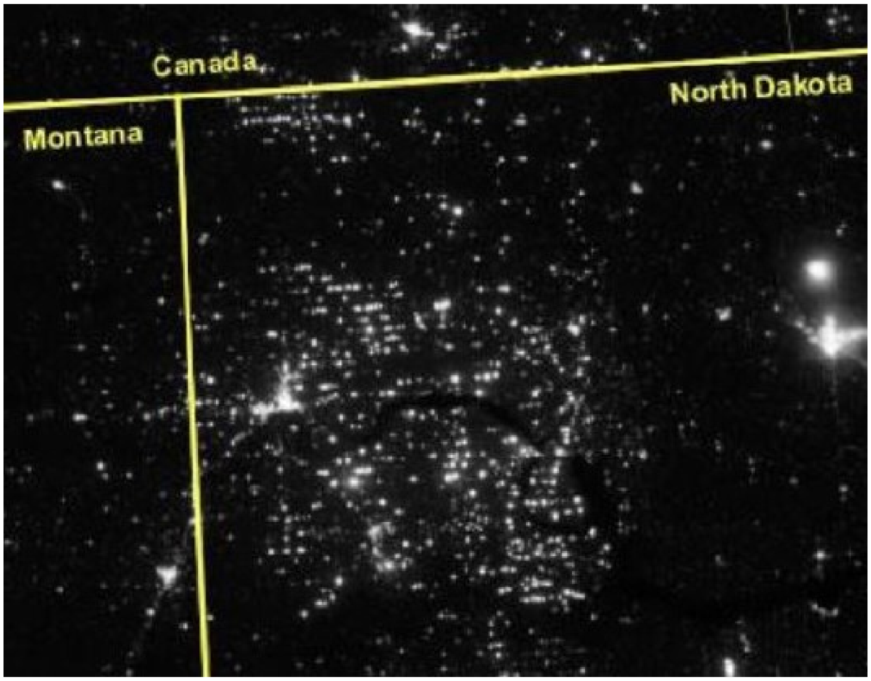


Figure 1-2: Natural gas flares in the Bakken Play, as seen from space [10]

In order to eliminate or curb the harmful impact of the flared natural gas and fugitive methane from remote and small gas or oil reservoirs, unit operations of small footprint, which are capable of converting this gas into liquids are required, since gas transportation cost could be up to five times as costly as liquid transportation [12, 13]. This could be achieved through process intensification and modular technologies, which refer to strategies that enable significant reduction of the physical dimensions of the conventional unit operation and allow simple geometric scaleup. Actually, process intensification has the potential to deliver major benefits to the petrochemical and chemical process industry by accelerating the response to market changes, simplifying scaleup and providing the basis for rapid development and deployment of new products [14]. Modular technology, on the other hand, enables on-site gas treatment and processing, reducing the environmental impact and safety risks inherent to pipeline transportation to a central facility. More details of process intensification and modular technologies are discussed in the following.

1.1 PROCESS INTENSIFICATION

Over the past two decades, significant attention has been focused on process intensification and modular technologies. The motivation of this approach lies in the fact that the main components of a given plant, such as reactors, heat exchangers, separators, etc., only represent about 20% of the overall capital cost, whereas 80% of the cost is incurred by installation and commissioning, which includes pipe-study, structural support, civil engineering, etc. [15]. This means that major reductions in the equipment size, coupled preferably with a degree of telescoping of equipment function, such as reactor/heat exchanger unit, or combined distillation/condenser/re-boiler, could result in significant cost savings by eliminating the support structure, expensive foundations and long pipe runs.

1.2 MODULAR TECHNOLOGY

Modularization is a construction method comprised of pre-built refining and processing facilities in small process units, with a designed capacity, which can be shipped to its destination on demand.

It has become an area of interest because of its simplicity in terms of scale-up and scale-down when a different capacity is necessary, by adding or removing modules [16]. Moreover, onshore and offshore remote natural gas fields require small, flexible units to be able to fit on the drilling, fracking and/or production pads and be eventually moved from one site to another when the field is depleted. The smaller the module, the cheaper it is to transport, and the faster the plant's commissioning, startup, and decommissioning will be [17].

The fact is that if the rate of natural gas flaring continues to increase at its current rate, by 2020 the total gas volume flared is projected to be up to 60% greater than that in the year 2000 [18]. Also, flaring natural gas is becoming more problematic due to increasing legislations and stringent regulations, in addition to worldwide extensive efforts to reducing GHG emissions [6]. Thus, there is a great and pressing need to curbing or eliminating the vented and flared natural gas from remote reservoirs in order to protect the environment and avoid global warming.

2.0 OBJECTIVE

This study aims at investigating the potential use of process intensification and modular strategies in technologies, which could be employed for curbing or eliminating the flared natural gas and fugitive methane from remote and small gas and/or oil reservoir. These technologies are the Fischer-Tropsch (F-T) synthesis process, in a microchannel reactor (MCR), and the Direct Methane to Methanol (DMTM) process in a compact plant footprint.

The main objective is to investigate the overall and economic performances and the potential use of these two processes in eliminating the deleterious impact on the environment and global climate change of this flared natural gas and fugitive methane by converting it into clean liquid hydrocarbons for cheaper and easier transportation to the end-user.

In order to achieve this objective, the two processes are modeled using the process simulator Aspen HYSYS v7.2. The operational performance of each process is assessed based on the mass and energy requirements; and the economic performance is evaluated in terms of the net present value (NPV), payback period (PBP) and internal rate of return (IRR). In addition, the effects of tailgas recycle ratios of the tail gas in the F-T synthesis process in MCR and methane in the DMTM process on the operational and economic performances are investigated.

3.0 BACKGROUND

3.1 FISCHER-TROPSCH SYNTHESIS PROCESS

The F-T synthesis process was originally developed by Franz Fischer and Hans Tropsch in the 1920's in an effort to produce liquid fuels, based on the 1902 discovery by Sabatier and Sanders [19]**,** that methane can be produced from H_2/CO mixtures over a nickel catalyst. It was a breakthrough in the conversion of coal to liquid fuel, such as kerosene and naphtha, and has been subsequently heavily investigated and developed commercially [20]. The F-T synthesis provides a pathway for converting carbon containing natural resources, such as natural gas, coal, heavy residue, biomass, municipal waste, etc., into liquid fuels and high value chemicals.

In this process, the CO and H_2 (syngas) react in the presence of a catalyst, conventionally iron or cobalt, to produce synthetic hydrocarbon products, primarily linear alkanes and alkenes. The overall F-T process involves three main steps: (1) syngas generation, (2) F-T catalytic reactions and (3) product upgrading as shown in Figure 3-1.

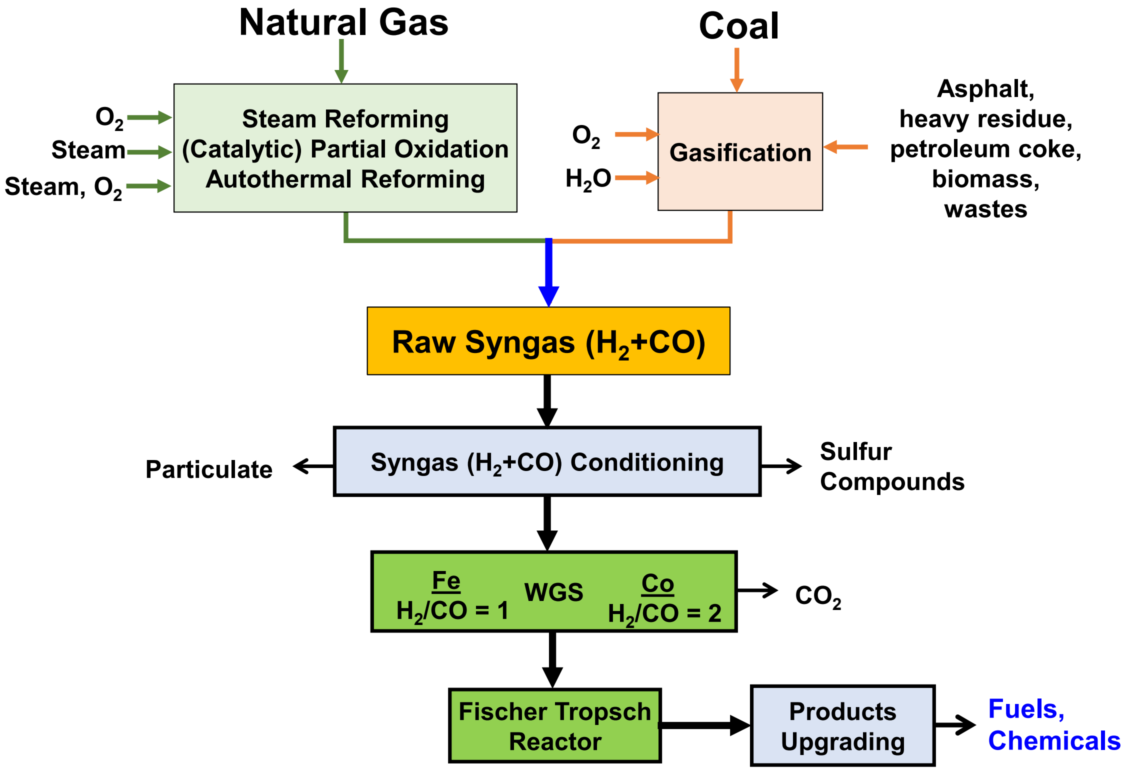


Figure 3-1: Overview of the F-T process [20]

3.1.1 Syngas generation

Syngas generation involves converting the carbonaceous feedstock into a H_2/CO mixture via reactions with steam and oxygen or optionally air. Solid feedstocks, such as coal and biomass, are converted in a gasifier, of which various types have been already in industrial applications [20]. Natural gas, on the other hand, is converted to syngas in a reformer using either partial oxidation (POX), steam methane reforming (SMR) or dry reforming (DR), as shown in Equations (3-1) to (3-3) [20].

Partial oxidation (POX): $CH_4 + 0.5\,O_2 \rightleftharpoons CO + 2\,H_2$ ΔH_{298} = -35.6 kJ/mol (3-1)

Steam methane reforming (SMR): $CH_4 + H_2O \rightleftharpoons CO + 3\,H_2$ ΔH_{298} = 205.6 kJ/mol (3-2)

Dry reforming (DR): $CH_4 + CO_2 \rightleftharpoons 2\,CO + 2\,H_2$ ΔH_{298} = 247.3 kJ/mol (3-3)

A fourth alternative route would be the combination of Equations (3-1) and (3-2), when the partial oxidation and the steam methane reforming occur simultaneously, which is known as Auto-thermal reforming (ATR).

These different routes have been extensively studied and many references defined the conditions and applicability for following each route. For instance, ATR and POX are not recommended in offshore operations due to the need for an air separation unit (ASU), thus increasing the plant total cost and footprint [21]. Also, safety issues are very important in operating in an oxidative atmosphere containing hydrogen and/or hydrocarbon due to the risk of explosion [22].

The produced syngas often requires a conditioning step to remove the contaminants, such as nitrogen, oxygen and sulfur, as well CO_2. The units accountable for these steps typically employ solid sorbents or physical and/or chemical solvent [23].

Process intensification efforts for F-T process have been primarily focused on syngas generation through SMR route. In the SMR reactor, Equation (3-2), coupled with Equation and (3-4) may occur simultaneously.

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad \Delta H_{298} = -41 \text{ kJ/mol} \tag{3-4}$$

Since Equation (3-2) is reversible and highly endothermic, authors have reported high reaction temperatures, between 850 °C and 950 °C [24, 25]. Equation (3-4) is known as the water-gas-shift (WGS) reaction, which is usually undesirable, except when H_2/CO ratios are low.

Typical catalysts for SMR are made of noble metals, such as Ruthenium (Ru) and Rhodium (Rh) because of the high activity of Equation (3-2) and low activity of Equation (3-4) [26]. Nonetheless, commercial applications employ supported Nickel because of its lower price, compared with that of the other noble metals [27]. Conventional supports include γ-Al_2O_3, $CaAl_2O_4$ and MgO. Figure 3-2 summarizes the relative activity of typical SMR catalyst.

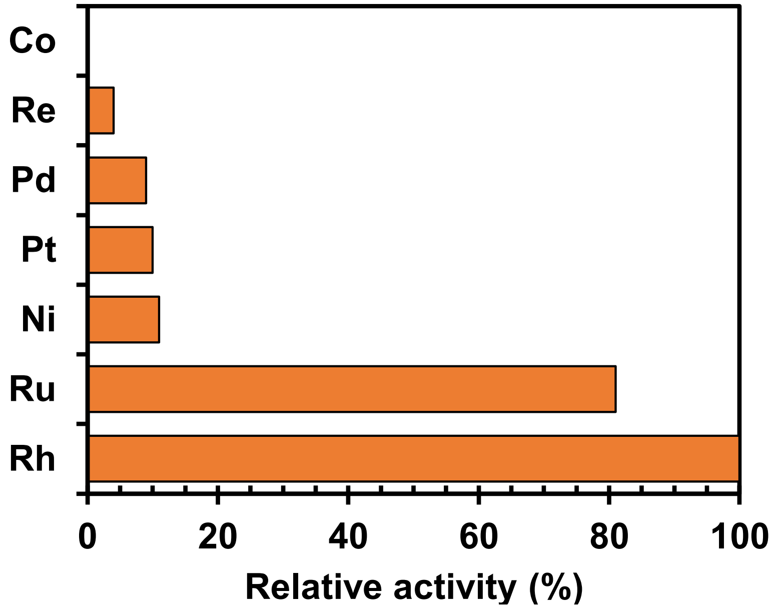


Figure 3-2: Relative activity of transition metals on SMR [26]

3.1.2 Fischer Tropsch Synthesis

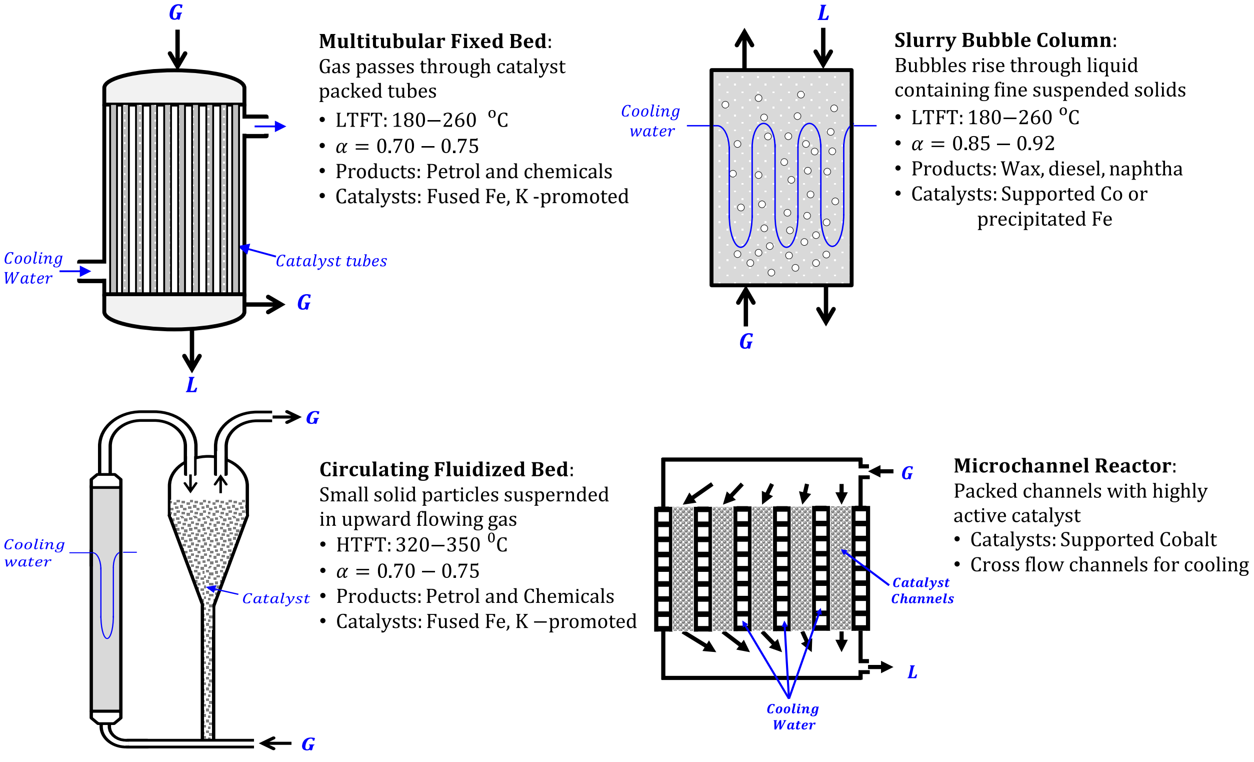
In F-T synthesis, CO and H_2 (syngas) react in the presence of a catalyst, conventionally iron or cobalt, to produce synthetic hydrocarbon products, primarily linear alkanes and alkenes. The main reactions occurring during the F-T process are paraffins, olefins, as shown in Equation (3-5) and the WGS reaction shown above.

$$F-T: \qquad CO+2H_2 \rightarrow -CH_2- \ +H_2O \tag{3-5}$$

Although many metals have been identified to catalyze F-T reactions, only iron (Fe) and cobalt (Co) have been used in industrial applications. Iron catalyst is cheap and has a high water-gas-shift (WGS) activity, however, it is prone to severe attrition in slurry reactors and the water produced during the reaction appeared to decrease its activity. Cobalt-based catalyst, on the other hand, has higher activity than iron catalyst since it is not strongly inhibited by water. It is more resistant to attrition and, as such, has a longer life in the reactor than iron catalyst. Cobalt-based catalyst, however, is more expensive than iron catalyst and has no WGS activity. During Cobalt catalyzed F-T reaction, the oxygen from CO dissociation is converted to H_2O, as shown in Equation (3-5). Conversely, iron catalyst has a high affinity for the WGS reaction as shown in Equation (3-4). resulting in the conversion of a significant portion of oxygen from CO dissociation into CO_2 [20]. Thus, the extent of the WGS reaction has to be closely considered as it affects the H_2/CO ratio in the F-T process.

Depending on the reaction temperature, the F-T process is referred to as low temperature F-T (LTFT) or high temperature F-T (HTFT). The temperature of the LTFT ranges from 180 to 260 °C and the syncrude produced is mainly wax consisting mostly of long chain hydrocarbons, while the temperature of the HTFT process is between 290 and 360 °C and the products are mostly short chain hydrocarbons and gases. Therefore, the final products of the LTFT process consist mostly of diesel, while gasoline production has been the focus of the HTFT [28]. The LTFT syncrude product is easy to upgrade by a hydroprocessing and a fractionation step to obtain naphtha and middle distillate, whereas the HTFT syncrude requires more complex refinery facilities [20, 29, 30]. It should be noted that recent R&D and commercial efforts have been focused on the LTFT due to the current drive for using more diesel engines than gasoline engines, the excellent quality of sulfur-free F-T diesel, and perhaps the mild conditions of the process.

Schematic of different reactors used for F-T process commercial applications are depicted in Figure 3-3. The circulating fluidized-bed reactor (CFBR) is used for the HTFT process, whereas multi-tubular fixed-bed reactors (FBRs) and slurry-bubble-column reactors (SBCRs) are used for the LTFT process. In addition, microchannel reactors (MCRs) for LTFT process in small-scale applications have been recently receiving increasing attention, even though; no commercial applications are yet available. This study focuses on the MCR technology for F-T synthesis.



Multitubular Fixed Bed:
Gas passes through catalyst packed tubes
- LTFT: 180−260 °C
- $\alpha = 0.70 - 0.75$
- Products: Petrol and chemicals
- Catalysts: Fused Fe, K -promoted

Slurry Bubble Column:
Bubbles rise through liquid containing fine suspended solids
- LTFT: 180−260 °C
- $\alpha = 0.85 - 0.92$
- Products: Wax, diesel, naphtha
- Catalysts: Supported Co or precipitated Fe

Circulating Fluidized Bed:
Small solid particles suspernded in upward flowing gas
- HTFT: 320−350 °C
- $\alpha = 0.70 - 0.75$
- Products: Petrol and Chemicals
- Catalysts: Fused Fe, K −promoted

Microchannel Reactor:
Packed channels with highly active catalyst
- Catalysts: Supported Cobalt
- Cross flow channels for cooling

Figure 3-3: Typical reactor designs found in the industry for F-T application [20]

Arzamendi et al. [22] conducted a 3-D, CFD heat transfer study in an MCR for the LTFT syntheses. Their simulated MCR consisted of a steel block with 80 square microchannels of 1 mm × 1 mm arranged in a cross-flow configuration, 40 vertical channels for the syngas and 40 horizontal channels for the cooling water as shown schematically in Figure 3-4. Their simulation variables were GHSV (5000-30,000 h^{-1}), pressure (5 and 35 atm), temperature (483-523 K), water feed (0.2-250 gram/minute), steam content (0.2-34.8%) and CO conversion (25%-95%). Their results showed that increasing CO conversion required decreasing the reactor pressure, which decreased the F-T process selectivity to middle distillates. Adjusting the cooling water flow rate in the range 0.25-250 gram/minute, however, allowed maintaining the process temperature at suitable values while avoiding the use of low pressures. Their results indicated that when processing up to 30,000 h^{-1} of syngas with $H_2/CO = 2$, the F-T process could be carried out at low-temperature (483-513 K), if an appropriates pressure value between 5 and 35 atm and a cooling water flow rate value between 0.25 and 250 g/minute were selected. They reported relatively high values of the overall heat transfer coefficients (20-320 Wm^{-2} K^{-1}) and claimed that the MCR exhibited good isothermicity under most of these simulated conditions.

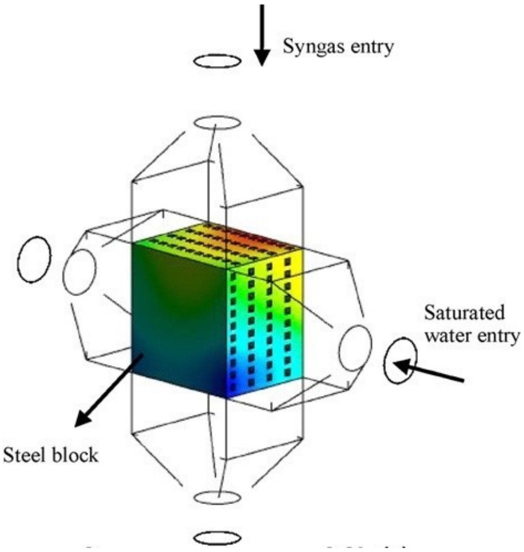


Figure 3-4: Structure of a micro-channel reactor [22]

From an economic perspective, a gas-to-liquid facility would create a demand for cheap or free natural gas, on one hand, while providing a supply of high quality liquid fuel and/or chemicals on the other hand [31]. In the case of shale gas as the feedstock, the syncrude produced could be transported by trucks to refineries instead of pipelines. Currently, modular units are being developed by companies, such as CompactGTL and Velocys, and manufactured by Texas-based, Ventech Engineers International LLC, to take advantage of using MCR technology for stranded gas monetization and flare reduction [32].

3.2 DIRECT METHANE-TO-METHANOL (DMTM)

Direct Methane-to-Methanol (DMTM) or partial oxidation of methane to methanol and formaldehyde (oxygenates) as given in Equation (3-6) could be a promising process for the effective use of flared and/or vented natural gas from remote and small gas and oil reservoirs. This equation shows that the reaction between methane and oxygen free radicals releases 126.2 kJ in order to produce one mole of methanol. It is a convenient alternative to conventional technologies, once the minimum selectivity to methanol and conversion are reached. Methanol is a precursor, which is used extensively in the production of many other chemicals, such as formaldehyde, acetic acid, chlorinated hydrocarbons and amines. It is also a versatile product with a variety of applications, such as biodiesel manufacture [33], removal of nitrogen compounds from water [34] and fuel cells development [35].

$$CH_4 + {}^{\cdot}O \rightarrow CH_3OH \qquad \Delta H_{298} = -126.2 \text{ kJ/mol} \qquad (3\text{-}6)$$

Recently, significant study has been focused on investigating the effect of reaction parameters and operating conditions on the DMTM process [36–38]. The temperature for optimum conversion was reported to be in the range of 400 - 500 ºC at pressures between 5 - 50 atm [36], while the effect of residence time on methanol selectivity was determined to be significant at values less than 100 s, while optimum yields were obtained at residence time of 200 s. Moreover, methanol selectivity was found to be optimum at 5% O_2 [36–38].

The main challenge of direct methane utilization is that it is a very stable, symmetrical molecule and is the least reactive hydrocarbon. However, the production of low molecular weight oxygenates could overcome the high stability of methane [39]. From the kinetic point of view, which is an important issue in reactor design, the gas-phase methane oxidation was studied and various mechanisms were proposed as summarized by Zhang et al. [40]. However, there remain significant challenges from both kinetics and thermodynamics of the gas-phase methane oxidation.

Despite its potential, the DMTM process still has to overcome some operational challenges before achieving commercialization. Recently, de Klerk [41] performed an engineering evaluation of this processes for small-scale GTL applications and reported the following:

1. The process is less efficient when operates with air as the oxidant fuel. It requires an air separation unit (ASU) to generate high purity oxygen. This significantly increases the cost and makes the process less attractive when compared with direct syngas generation by autothermal reforming, which also requires an ASU;
2. Since Natural Gas is only around 91-93% methane, the introduction of inert gases is unavoidable, which significantly affects the process design, gas loop capacity, recycle strategy and purge loss;

3. DMTM has a significant utility footprint, with required large compressor and cooling duties;
4. The overall carbon and thermal efficiencies of the DMTM process are around 35% and 28%, respectively, which are almost half of those of the indirect methanol synthesis process, 66% and 53%, respectively; and
5. The process chemistry is hard to control and requires operation within a very narrow and constrained range to avoid further methanol oxidation.

The modular unit proposed in this study is a tentative replicate of the GasTechno technology Mini-GTL plant [42], which has a small footprint of (90 ft x 70 ft) and can be scaled down to 5,000 standard cubic feet of methane per day (5 MSCF/day). The general schematic of the GasTechno process is shown in Figure 3-5.

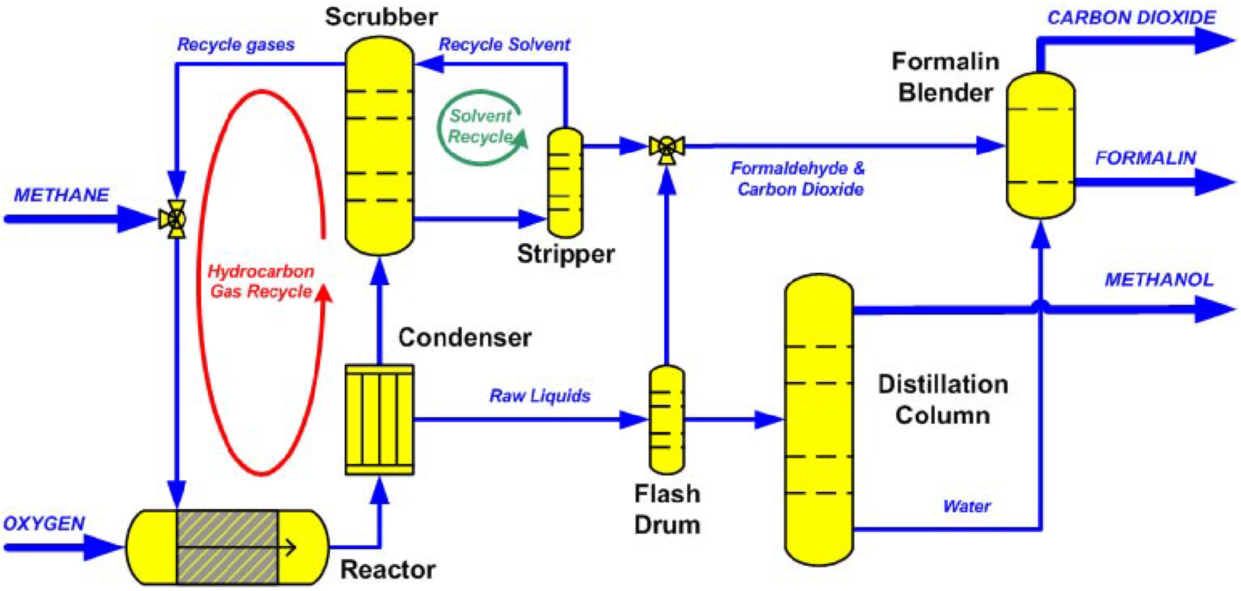


Figure 3-5: GasTechno DMTM process flowsheet (Taken from GasTechno [43])

As shown in Equation (3-6), the reaction between methane and oxygen free radicals is highly exothermic (-126.2 kJ/mole of methanol) and accordingly this heat of reaction has to be removed from the process in order to keep isothermal conditions in the DMTM reactor. In addition, since the reactions take place in the liquid-phase, the DMTM reactor should operate at high pressures, around 80 atmospheres. The patented DMTM reactor has a unique design with different temperature zones, as shown in Figure 3-6.

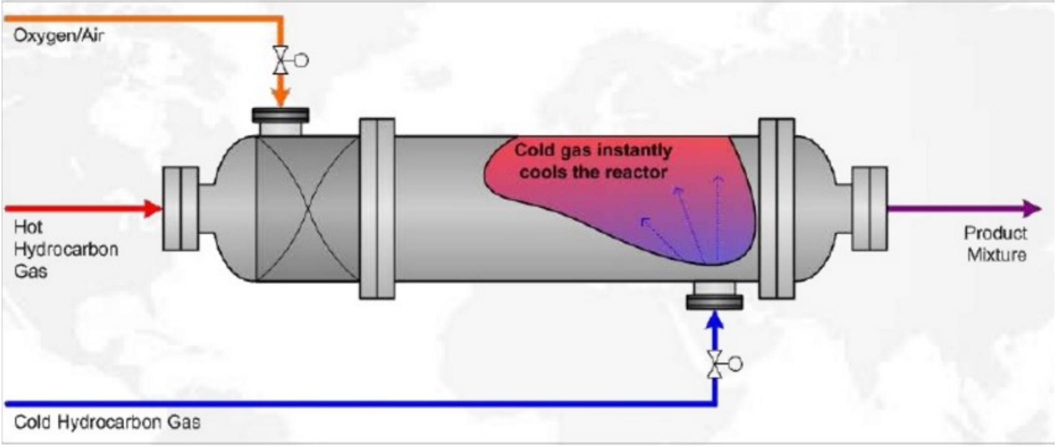


Figure 3-6: DMTM reactor design (Taken from Patent No. US 2010/0158760 A1 [44])

The process output contains two secondary oxygenate products, ethanol and formaldehyde according to Equations (3-7) and (3-8), respectively:

$$CH_4 + O_2 \rightarrow CH_2O + H_2O \qquad \Delta H_{298} = -275.9 \text{ kJ/mol} \qquad (3\text{-}7)$$

$$2CH_4 + O_2 \rightarrow C_2H_5OH + H_2O \qquad \Delta H_{298} = -160.6 \text{ kJ/mol} \qquad (3\text{-}8)$$

In order to enable selling each products, a separation unit, located onsite or offsite, is necessary after the DMTM reactor. Actually, membrane filtration technology was reported to be the most efficient route for separating these products, but a distillation column could also be an alternative.

4.0 RESEARCH APPROACH

4.1 ASSUMPTIONS AND FEED CONDITIONS

The two proposed, small-scale GTL processes were modeled using the process simulator Aspen HYSYS v7.2. The processes were built and simulated to obtain complete energy and mass balances for every stream. The following two assumptions were made:

1. The entire system was assumed to be under non-transient (steady state) conditions, even though it is known that external forces could boost the production of the associated gas, such as losing control of well production and overpressure in the equipment downstream units, as well as market fluctuations affecting the demand for hydrocarbons.
2. The conversion and selectivity values, under process conditions (temperature, pressure and flow rate), and products yield were taken from the literature data and those available by the manufacturer.

In the F-T process carried out in the MCR, the natural gas feed molar flow rate to the process was set at 2,367.79 kmol/h, which corresponds to around 47.5 MMscf/day. This value was based on the ranges provided by CompactGTL for Oilfields producing up to 50 MMscf/day of associated natural gas offshore or 5-150 MMscf/day onshore [45]. Moreover, the feed composition is given in Table 4-1, and was assumed to be identical to that of the associated gas from the Bakken field reported by Wocken et al. [46]. Thus, the feed gas for both processes is a relatively wet, nitrogenrich, sour natural gas containing 0.01 mol% H_2S, 5.21 mol % N_2, 0.57 mol% CO_2 and 0 mol% C_{10}^+. It should be mentioned that all H_2S was completely removed from this associated gas before entering the MCR in order to prevent the F-T catalyst poisoning, and therefore, all simulations were performed were without H_2S.

Table 4-1: Gas Composition used in this study (Bakken associated gas) [46]

Component	Mole %	Molar flow rate (kmol/h)
C_1	57.67	1,361.34
C_2	19.94	474.15
C_3	11.33	269.42
i-C_4	0.97	23.07
n-C_4	2.83	67.29
i-C_5	0.38	9.04
n-C_5	0.71	16.76
C_6	0.22	5.23
C_7	0.09	2.14
C_8	0.04	0.95
C_9	0.01	0.24
H_2O	0.02	0.48
N_2	5.21	123.89
CO_2	0.57	13.55
H_2S	0.01	0.24
	100	2,367.79

4.2 ECONOMIC ASSESSMENT METHODOLOGY

In this study, the following criteria were used for the F-T in MCR and DMTM processes in order to perform the economic analysis as follows:

1. The funds used by the company to acquire physical assets, such as property, industrial buildings or equipment, were included in the analysis.

2. A Fixed Capital Investment (FCI) of 238.5-Million 2016 USD was used for the F-T process [17, 47].
3. The FCI for the DMTM process was estimated from the value reported by GasTechno, who reported 56-Million USD for a 30 MMscf feed, for a process that includes an ASU [42, 43].
4. The cost of the ASU, which was determined to be about 19-Million USD using the economic assessment charts in Ebrahimi et al. [48], was not included in this analysis.
5. The feed in this study was 41.7-million SCF and the capital cost was linearly scaled from 30-million SCF. Therefore, the fixed capital cost for the DMTM process was determined to be 51.5-million USD.
6. The costs of heat and cooling utilities as well as electricity were taken from the literature in order to estimate the costs of running the processes as given in Table 4-2 [20, 49].
7. Both units were assumed to have a lifetime of 20 years, which is a typical, conservative value in the processes industry.
8. A constant depreciation rate of 10% of the FCI per year was considered in both processes, in order to account for the physical deterioration of the facilities and their decline in usefulness throughout the years, considering eventual accidents, corrosion, and wear and tear.
9. A tax rate of 30% was used.
10. The minimal interest rate was set at 10%.
11. An on-stream factor (annual average capacity factor) of 0.90411 was used, representing 330 days of operations per year, with one month of planned maintenance stop.
12. The catalyst of the SMR and the F-T process need to be regenerated periodically and it was assumed to cost 1% of the unit's CAPEX [23].

Three main metrics were then used to evaluate the economic feasibility of the process: (1) Payback Period (PBP), (2) Net Present Value (NPV) and (3) Internal Rate of Return (IRR).

Table 4-2: Main assumptions for economic evaluation [20, 49]

Plant economic life	20 years
Annual average capacity factor	0.90411
Water	0.02 USD/m^3
Electricity	0.26 USD/kWh
Natural Gas	0 USD/m^3
Synthetic crude oil	42 USD/bbl
Methanol	1.06 USD/gal
Ethanol	1.16 USD/gal
Formaldehyde	1.41 USD/gal

The PBP is the time required for the amount invested in an asset to be repaid by the net cash outflow generated by the asset. It is used to evaluate the risk associated with a proposed project. PBP is expressed in years and fractions of years.

$$Payback\ period\ (years) = \frac{FCI}{TPI - TOC} \tag{4-1}$$

Where:

FCI: the fixed capital investment ($)

TPI: the total plant investment ($/year)

TOC: the total operating cost ($/year)

The NPV is the difference between the present value of the cash inflow and the present value of the cash outflow. NPV is used in capital budgeting to analyze the profitability of a projected investment or project.

$$NPV = -TCI + (TPI - TOC) \cdot \frac{(1+i)^Y - 1}{i \cdot (1+i)^Y} + TAC \cdot (1+i)^{-Y} \tag{4-2}$$

Where:

TCI: the total capital investment ($)

i: the internal rate of return, which was set to 10% when calculating the NPV.

Y: the plant lifetime (years)

TAC: the total annualized cost, calculated using:

$$TAC\ (\$/year) = Depreciation + TOC \tag{4-3}$$

The IRR is a discount rate that makes the net present value (NPV) of all cash flows from a particular project equal to zero. It is also referred to as the maximum discount rate. It is a metric used in capital budgeting for measuring the profitability of potential investments.

Once the PBP, NPV and IRR were calculated, a sensitivity analysis was performed to determine how the fluctuations in both the FCI and the cash flow affect the IRR. In this study, both the FCI and the cash flow were varied from -30 to 30% in the simulated runs.

Because of its abundance and low demand, the price of stranded gas was assumed to be null ($0.00) for the base-case. However, a sensitivity analysis was performed and the effect of an eventual stranded gas market charging regular natural gas price on the cash flow are also discussed in the results section.

4.3 FISCHER-TROPSCH PROCESS SIMULATION

4.3.1 F-T Process Description

In this study, Figure 4-1 shows a simplified flow diagram of the proposed F-T process in a MCR; and a more detailed schematic of this process is given in Appendix A. The process was modeled in Aspen HYSYS v7.2. The simulated module consists of five (5) main units: (1) SMR; (2) Flash separation, (3) semi-permeable membrane separator; (4) F-T and (5) Three-phase separator. The recycle ratio of the tail gas, including the unreacted CO and H_2, was varied in order to determine the optimum recycle ratio, which enhances synthetic hydrocarbon product yield and increases the overall syngas conversion.

The methane molar flow rate of 1,361 kmol/h at temperature of 25 °C and 1.0 bar was fed to the unit to combine with 2,042 kmol/h of steam at 180 °C. This is because the steam to methane molar ratio was set at 1.5, which is the ratio required to achieve a maximum methane conversion of 91.3% with a CO selectivity of 88.6% in the SMR reactor [23, 24] . The combined stream was compressed to 4 bar and heated to 800 ºC, which is the optimum operating condition in the microchannel SMR [18, 36, 37]. It should be mentioned that the Peng-Robinson Equation of state (PR-EOS) was used in the Aspen simulation due to its reliability in predicting the phase behavior of organic fluids [52].

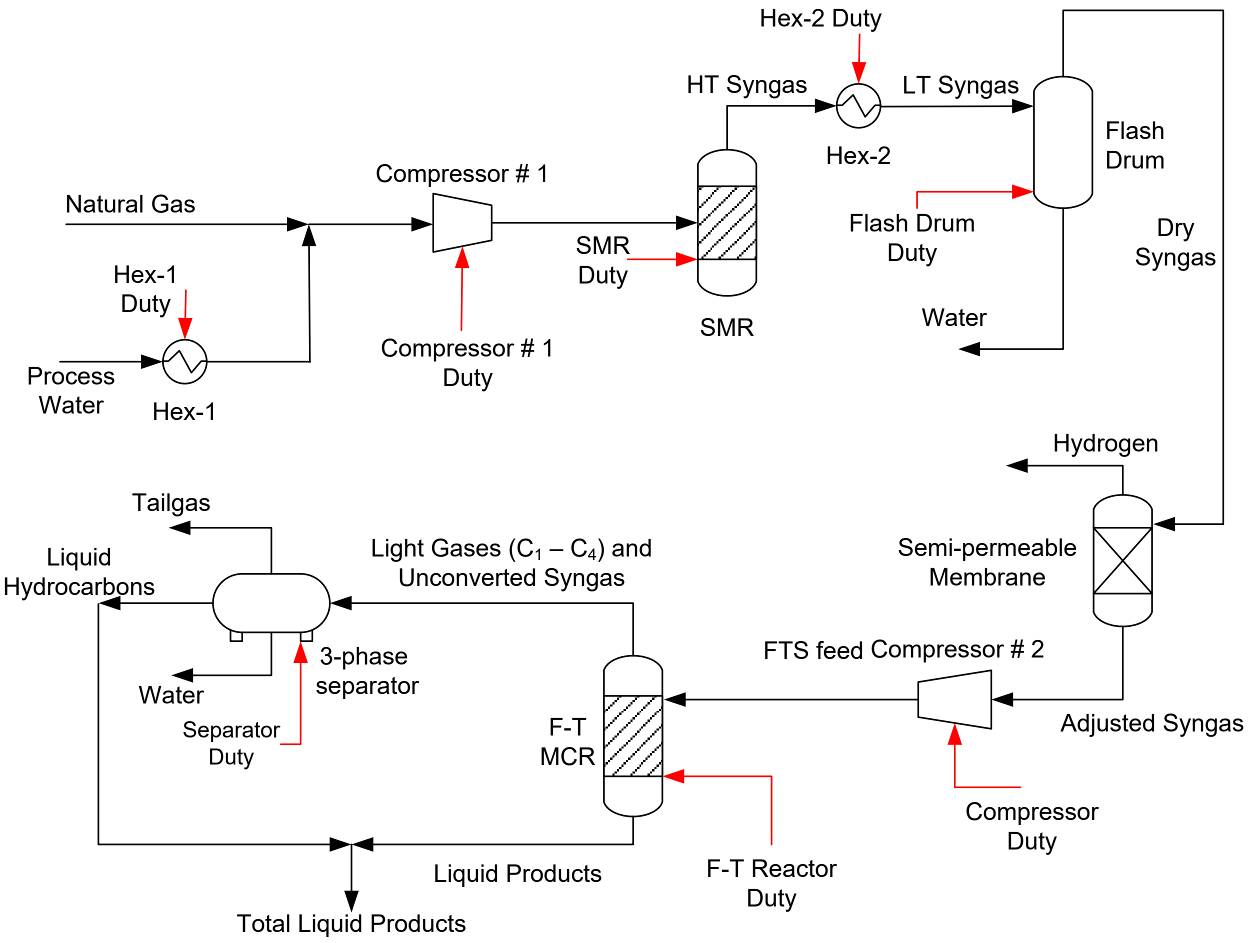


Figure 4-1: SMR followed by the F-T process flow diagram used in the simulation

The hot gaseous product stream from the microchannel SMR consists of a mixture of unconverted methane, water, hydrogen, carbon monoxide and carbon dioxide at 800 °C. The products are cooled down to 280 °C before entering the flash drum to remove all the unreacted water. This water could potentially be recycled and used as a cooling utility for the F-T reactor or the SMR.

4.3.2 Units Description and Assumptions

4.3.2.1 Natural Gas Reformer

The natural gas reformer (SMR) was represented using a conversion reactor unit in Aspen HYSYS since it uses the stoichiometry and conversion of the base reactant (CH_4) to determine the composition of the output stream. The reaction proceeds until either the fixed conversion value is reached or the limiting reactant is exhausted. Further details about the conversion reactor can be found elsewhere [53]. The SMR conditions were set to P = 20 atm, T = 860 ºC and steam/CH_4 ratio of 1.5/1. The equilibrium CH_4 conversion was set to 75% and the equilibrium CO selectivity was set to 73%. These values were obtained from the experimental data for steam methane reforming inside a MCR by Mazanec et al. [24], who performed their experiments in a Velocys microchannel reactor at 860 ºC and 20 atm over a highly active cobalt catalyst. Their results showed that the SMR equilibrium conversion and selectivity were reached within very short residence times (about 6.4 ms) inside a MCR coated with Ni/Al_2O_3 as can be observed in Figure 4-2. Arzamendi et al. [21] performed CFD simulations on SMR inside a MCR and used square channels of 7 mm × 7 mm, and 20 mm in length. Generally, the size of the microchannel SMR is about 1/25 (4%) of a conventional industrial reformer [24]. The coating of the adjacent channels is a Palladium-based catalyst, which facilitates the reaction of methane with oxygen, thus enhancing its conversion rate [21].

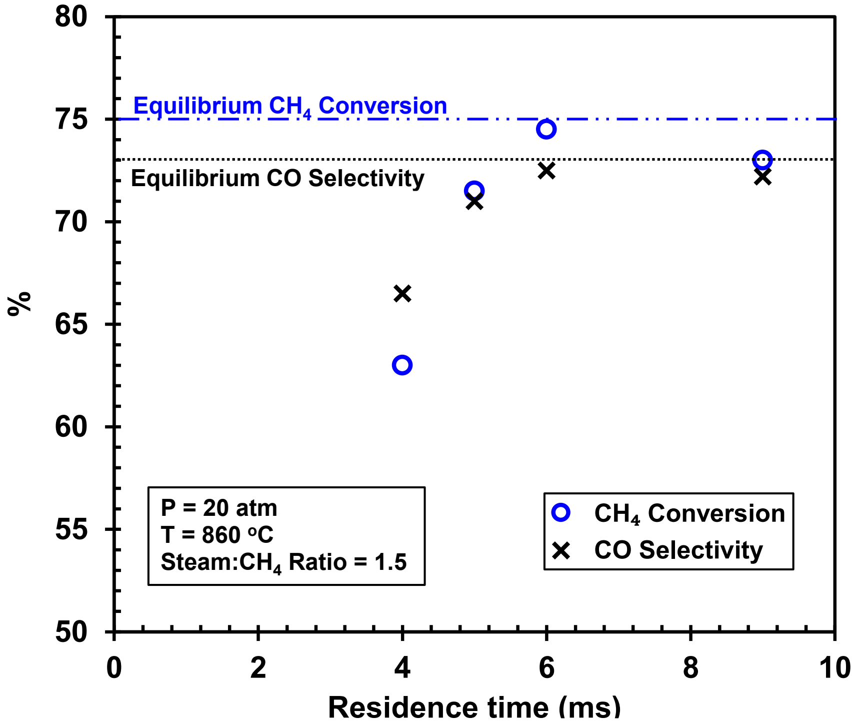


Figure 4-2: Experimental values of SMR performance in a microchannel reactor (Reproduced from Mazanec et al. [24]**)**

4.3.2.2 Flash Drum

A flash drum was placed after the SMR in order to separate the water from the stream prior to entering the semi-permeable membrane and the microchannel F-T reactor. In order to find the optimal temperature at which all the water could be separated in the flash drum, a case study was performed in which the water separation from the gaseous products stream was determined at various temperatures, as shown in Figure 4-3. The chosen operating temperature was subsequently set to 10 °C, allowing the removal of about 99% of the water.

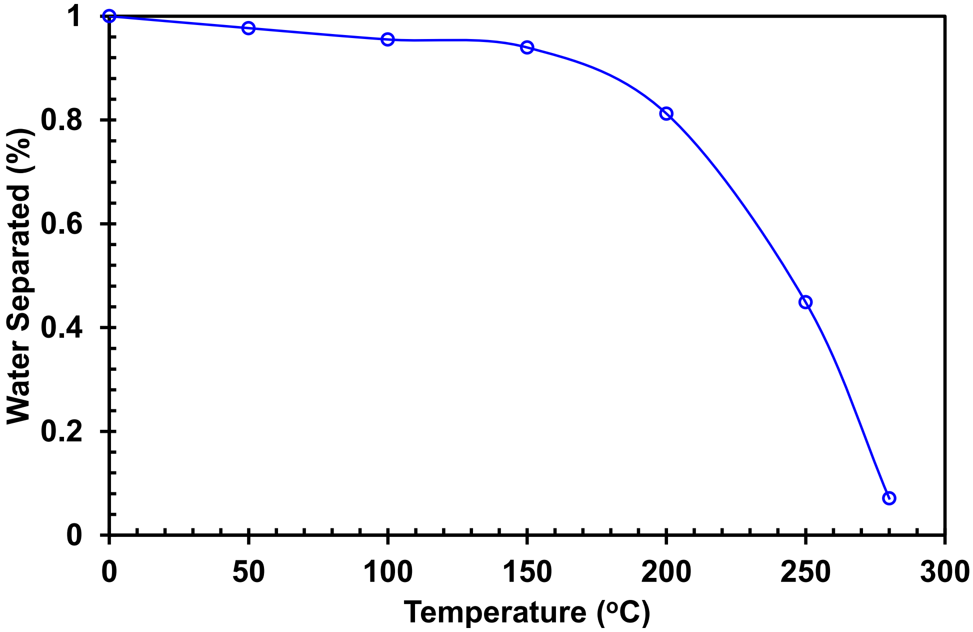


Figure 4-3: Effect of temperature on water removal efficiency

4.3.2.3 Semi-Permeable membrane

The H_2/CO ratio in the dry syngas exiting the flash drum was 3/1, which is greater than the optimal value of 2/1. A H_2/CO of 2/1 is the ideal ratio required to enhance the yield of middle distillates when using a cobalt-based F-T process [51]. Hence, to adjust the ratio from 3/1 to 2/1, a semi-permeable membrane was used to separate some H_2, resulting in a hydrogen stream. In this unit, hydrogen preferentially permeates through the membrane, producing a purified hydrogen "permeate" stream and a hydrogen-depleted "ratio-adjusted syngas" stream at 85 to 95 volume % purity [54]. Such membrane separators can be made of different materials, including polymers, ceramics, and metal alloys. However, significant research over the past decade was focused on using metal membranes, such as palladium, supported on porous ceramics, for H_2 separation [55, 56].

The separated hydrogen could be mixed with the fuel gas and sent to turbines for electricity generation or sold to refineries.

After adjusting the H_2/CO ratio, a compression step was necessary to increase the pressure to 25.5 bar and the temperature to 220 °C, which is the desired temperature for LTFT.

4.3.2.4 F-T MCR

The microchannel reactor used in this study consists of 21 rows; each row contains 100 microchannels with dimensions of 3 mm × 4.5 mm. The dimensions of the reactor are length = 0.305 m, width = 0.1 m and height = 0.15 m as shown in Figure 4-4. The superficial gas velocity through each channel of the reactor is 1.27 m/s. The MCR has water flowing in the cross flow channels adjacent to the packed channels in order to keep the reactor operating under virtually isothermal conditions. Keeping the temperature constant is vital in order to control the F-T products distribution. The products distribution was represented using the Anderson-Schulz-Flory (ASF) distribution, as discussed elsewhere [20]. It should be mentioned that cobalt catalyst was reported to achieve between 72% and 80% CO conversion with an $\alpha = 0.917$ for F-T MCR [57].

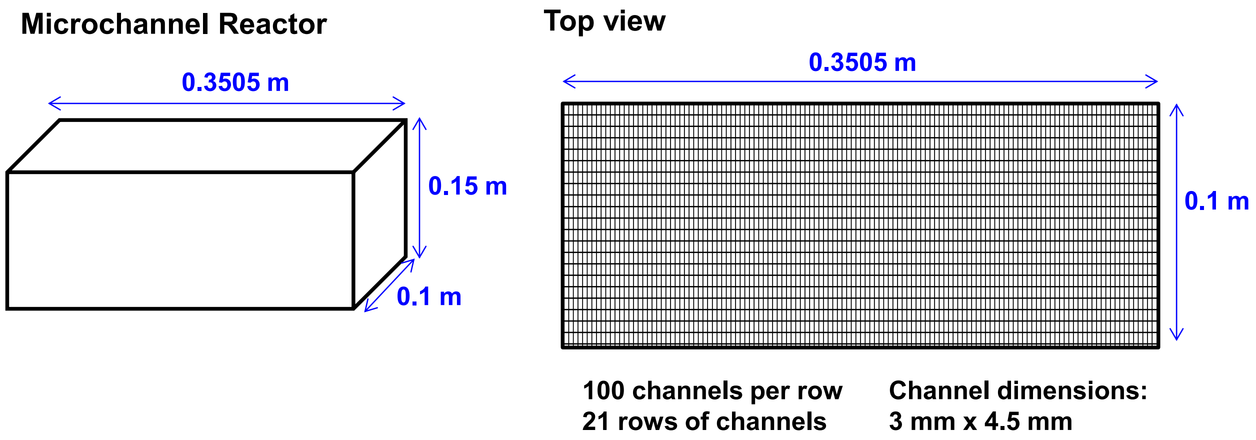


Figure 4-4: Dimensions of the microchannel reactor used in this study

In this study, methane was considered the representative element of natural gas, while propane (C_3H_8), Isooctane (C_8H_{18}), n-hexadecane ($C_{16}H_{34}$) and n-hexacosane ($C_{26}H_{54}$) were selected to represent LPG (Liquefied petroleum gases), naphtha, diesel and wax, respectively. In order to determine the selectivity of each pseudo-component at $\alpha = 0.917$, Equation (4-4), which is the linearized form of the ASF distribution, was used. The mole fraction (y_n) values were determined for every carbon number (n) in the range of 1 to 30 as can be seen in Figure 4-5.

$$y_n = (1 - \alpha) \cdot \alpha^{n-1} \tag{4-4}$$

Equation (4-5) was then used to determine the mole fraction for each pseudo-component by adding the mole fractions for every component within the range of C_1-C_2 for light gases, C_3-C_4 for LPG, C_5-C_{11} for Naphtha, C_{12}-C_{18} for Diesel and C_{19}-C_{30} for wax.

$$y_{k-l} = \frac{\sum_{n=k}^{l} y_n}{\sum_{m=1}^{30} y_m = 1.0} \tag{4-5}$$

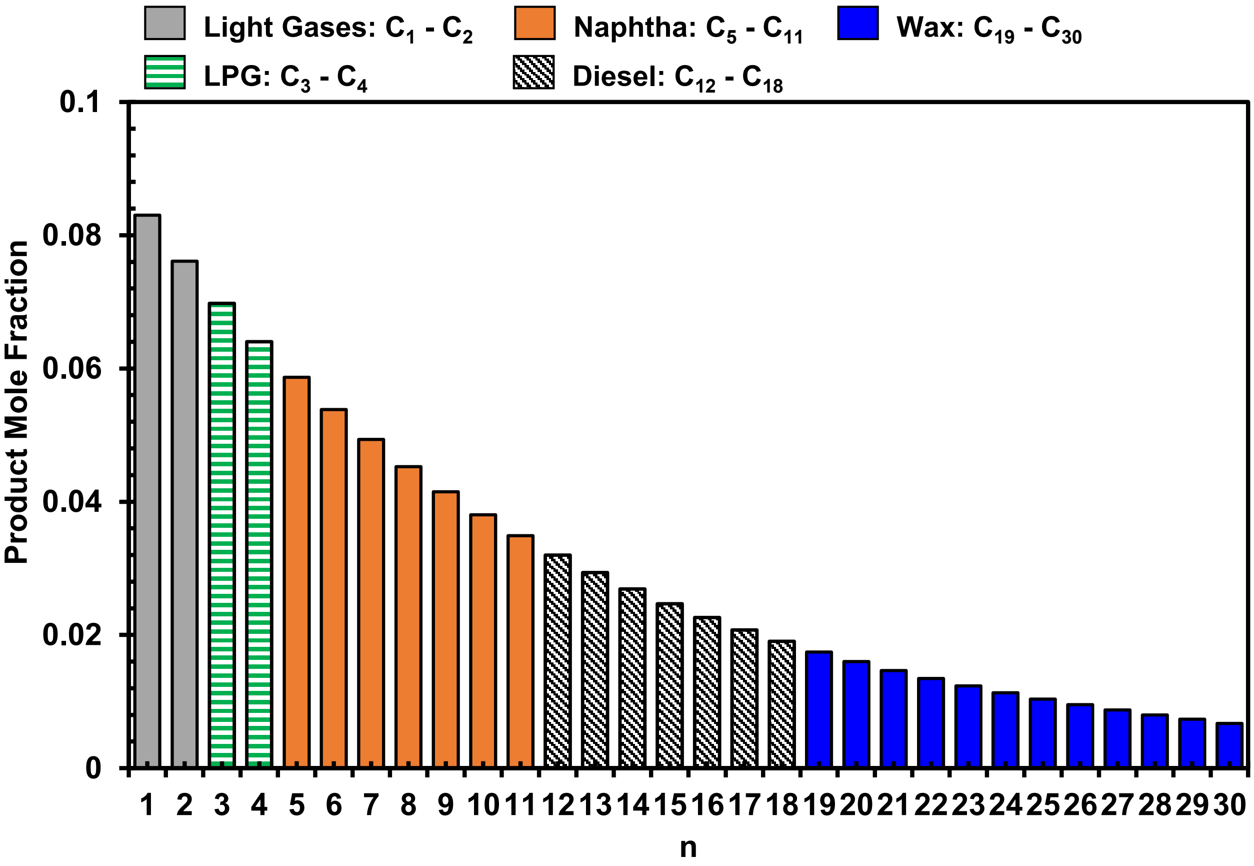


Figure 4-5: Product distribution for α = 0.917

The values obtained represent the mole fraction of each cut considering only the produced hydrocarbon mole fraction, which can be interpreted as the selectivity to a given fraction of the syncrude, defined as the number of moles of a pseudo-component divided by the total number of moles of CO converted into hydrocarbons. These values were calculated as given in Table 4-3.

Table 4-3: CO Selectivity towards each petroleum product

Product	Selectivity
Light Gases (C_1-C_2)	2.64%
LPG (C_3-C_4)	5.22%
Naphtha (C_5-C_{11})	27.59%
Diesel (C_{12}-C_{18})	28.80%
Wax (C_{19}-C_{30})	35.77%

Moreover, in order to solve the material balance equations, the stoichiometric coefficients were obtained using Equations (4-6) to (4-10).

Light Gases $CO + 3 H_2 \rightarrow CH_4 + H_2O$ (4-6)
Liquefied Petroleum Gases $3 CO + 7 H_2 \rightarrow C_3H_8 + 3 H_2O$ (4-7)
Naphtha $8 CO + 17 H_2 \rightarrow C_8H_{18} + 8 H_2O$ (4-8)
Diesel $16 CO + 33 H_2 \rightarrow C_{16}H_{34} + 16 H_2O$ (4-9)
Wax $26 CO + 53 H_2 \rightarrow C_{26}H_{54} + 26 H_2O$ (4-10)

Therefore, for n-alkanes, the following equation holds:

n-alkane $n CO + (2n + 1) H_2 \rightarrow C_nH_{2n+2} + n H_2O$ (4-11)

It is important to notice that the ratio between the stoichiometric coefficients of H_2 and CO in the selected syncrude products listed in Table 4-4 is close to 2/1 in the feed to the reactor, particularly for hydrocarbons with high carbon numbers.

Table 4-4: H_2/CO ratio of the selected F-T products

Product	Stoichiometric ratio (n_{H2}/n_{co})
Light Gases (C_1-C_2)	2.76
LPG (C_3-C_4)	2.29
Naphtha (C_5-C_{11})	2.13
Diesel (C_{12}-C_{18})	2.07
Wax (C_{19}-C_{30})	2.04

4.3.2.5 Three-phase separator

The overhead gas-phase from the F-T MCR is sent to a three-phase separator to remove both the water and liquid hydrocarbons. The mole fraction of the syncrude is low and so is its vapor pressure. Hence, another case study was carried out on the gas stream output from the three-phase separator to minimize the possibility of loss by vaporization of the light and valuable hydrocarbons, such as LPG into the gas-phase, subsequently the optimum temperature was found to be 10 ºC.

4.3.3 F-T Process Integration

4.3.3.1 Water Re-use

In order to optimize the F-T process in a MCR, it is necessary to integrate the energetic and material streams of the entire system. The goal is to reduce the duty of the heat exchangers; the consumption of natural gas for the catalytic combustion in the SMR; and the coolant water in some equipment, specially reactors and separators. The demand of utilities, which are purchased only for heat exchange tasks, is directly proportional to the feed flow rate, but it is also dependent on the arrangement of heat exchangers. There are two streams, which can enhance the process profitability in the case of integration: (1) the excess water removed in the flash drum after the reformer and (2) the water removed from the bottom of the three-phase separator as shown in Figure 4-6. It should be noted that the heat utility labeled Q-FT-Water is the same as that labeled Q-FT, therefore the cold water is heated using the heat generated from the F-T MCR. This means that the two recycled water streams could be utilized to remove the heat generated by the highly exothermic F-T reactions, which take place in the MCR. The water will flow in the adjacent channels and, after exchanging heat with the process fluid, can be mixed with the natural gas fed into the reformer. This integration dramatically reduces the fresh process water requirements from about 37,057 kg/h to only 4,321 kg/h.

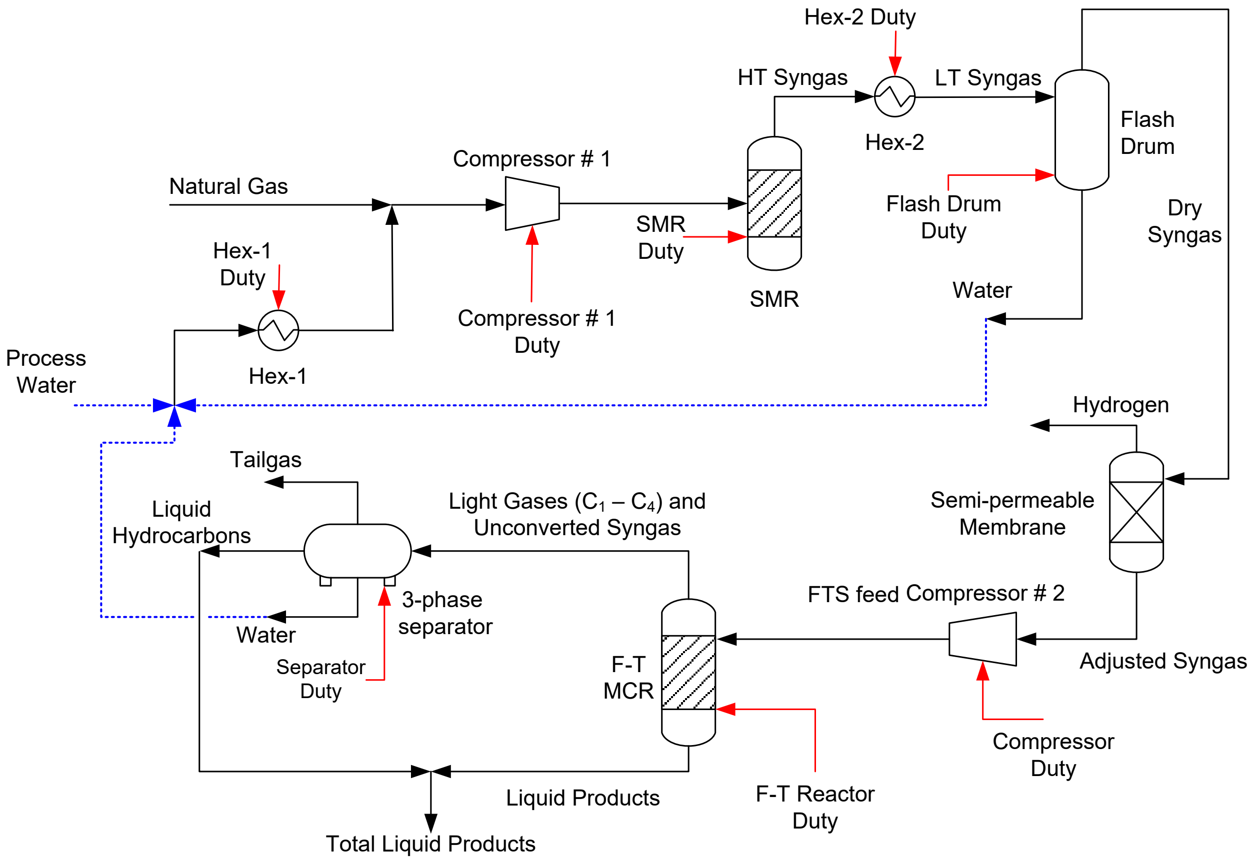


Figure 4-6: **Water re-use scheme used in this study**

It is important to mention that in the commercial Shell's Pearl GTL, located in Qatar, the water cogeneration plays a vital role, since the plant is located in an arid country, which lacks natural water resources. The Pearl's two fixed-bed trains (140,000 bpd capacity) is water self-sufficient, using most of the excess water produced for steam generation, cooling utility, and contaminant removal [57, 58].

The fact is a complete optimization of the proposed F-T process in a MCR could be a complex and challenging task. This is because, if only 2 utilities are considered for heat exchange (hot and cold) for a process including 5 streams, there will be 126 possible options of heat exchanger and only one is the optimum solution [58]. The design of heat exchangers, however, is out of the scope of the present study, since the GTL units are commercialized as manufactured modules. The presence of an endothermic process followed by an exothermic one elucidates the opportunity for energy integration, whereas the presence of water and tail gas as by-products of the system is behind the motivation for the search for material integration opportunity.

4.3.3.2 Tailgas recycle

The tail gas represents the stream coming from the three-phase separator. The effects of tailgas recycle on the operational and economics process performances were investigated. The tailgas recycle configuration is shown in Figure 4-7. The recycle ratio was defined as follows:

$$Recycle\ Ratio = \frac{Molar\ flow\ of\ Tailgas\ recycled}{Total\ Molar\ flow\ of\ Tailgas\ from\ the\ 3-phase\ separator} \times 100\% \qquad (4\text{-}12)$$

The recycle ratio was varied from 0% to 90% and the corresponding yields and economic parameters were determined.

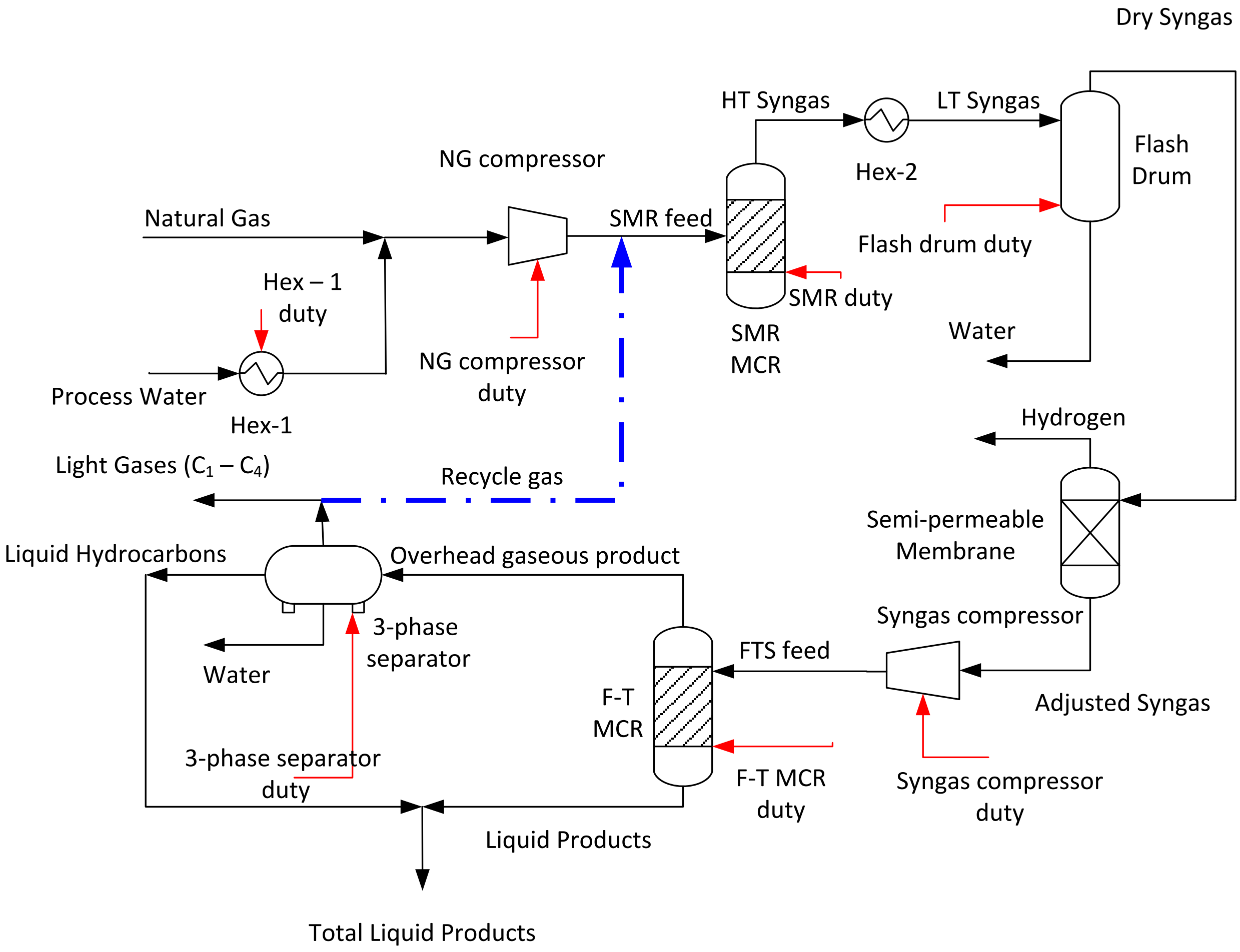


Figure 4-7: **Tailgas recycle configuration used in this study**

4.4 DMTM PROCESS SIMULATION

Figure 4-8 shows a schematic of the DMTM process flow diagram, which was implemented in HYSYS and a more detailed flow diagram, can be found in Appendix B. The UNIQUAC fluid package was chosen to represent the chemical species, because it accounts for all the binary interactions of the oxygenates [59, 60]. Since methane conversion is considerably low (about 3%

per pass) [41, 61], the recycle ratio of the unreacted methane was varied in order to determine its effect on enhancing oxygenates yield and increasing the overall methane conversion.

In this process, the natural gas was initially sent through a semi-permeable membrane to separate the methane from the rest of the feed constituents. The methane is subsequently mixed with both the recycle stream and air before entering the DMTM reactor. The recycle ratio was varied by changing the split between recycle and purge rate on the separator as follows:

$$Recycle\ Ratio = \frac{Molar\ flow\ rate\ of\ recycled\ CH_4\ from\ DMTM\ reactor}{Total\ Molar\ flow\ rate\ of\ unconverted\ CH_4\ from\ DMTM\ reactor} \tag{4-13}$$

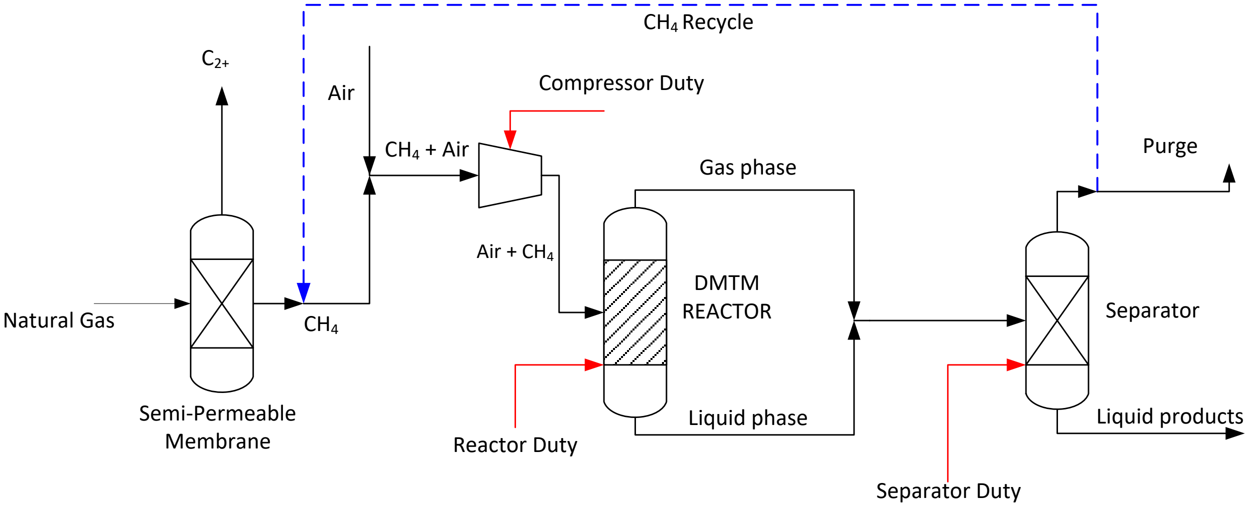


Figure 4-8: **DMTM flow diagram used in the simulation**

The amount of air used was controlled so that the oxygen mole fraction is kept constant at 4% of the mixed stream [23, 61]. This was due to the fact that high oxygen concentrations were reported to increase methane conversion and reduce methanol selectivity [61]. The upper oxygen concentration limit is 10%, above which the (methane + O_2) mixture is within the explosion envelope [61, 62]. The (gas + air) stream is then compressed up to 80 bar, which is the required reaction pressure, then enters the DMTM reactor at this pressure and 500 °C [61, 63].

4.4.1 Units Description and Assumptions

4.4.1.1 Semi-permeable Membrane for Methane Separation

The feed gas was initially fed to a semipermeable membrane to separate the methane from the rest of the gas constituents, this was carried out since methane is the main reactant in the DMTM process, and the effect of other feed components on the reaction process and kinetics within the DMTM jet reactor has not been investigated. Research efforts investigating membranes for gas separation technology have primarily aimed at polymeric membranes, such as cellulose acetate and polyaramide [64, 65].

4.4.1.2 DMTM Reactor

The DMTM reactor requires the input of conversion into the simulator for each of the resulting product. Due to their versatile industrial usage and high profitability, only methanol, formaldehyde, and ethanol were taken into account. The selectivity was calculated be 50.2% for methanol, 45.3% for formaldehyde, and 4.5% for ethanol, The theoretical selectivities of 6.2, 5.1 and 0.4 gallons of methanol, formaldehyde and ethanol, respectively, for every 1000 SCF of methane converted in the DMTM reactor, were used as reported in multiple references [61, 63, 66]. In this study, the diameter of the stainless steel DMTM reactor used was taken to be 30 mm [62, 63]. The DMTM reactor used in this study has the following characteristics [44]:

Residence time: 30 ms

Superficial Gas Velocity = 159.6 m/s

Reactor Diameter = 30 mm

Reactor Length: 5 m

4.4.1.3 Product Separator

Upon leaving the DMTM reactor, the stream containing liquids and unconverted hydrocarbons enters the separator at 500 ^{o}C and 80 bar, which separates the desired cold products at 25 ^{o}C and 78 bar and recycles the unconverted methane. This step corresponds to a fractional distillation unit. The split of methane could be adjusted by sending it back to the process or purging in order to keep the overall material balance in the process constant. Unquestionably, purging methane to the environment is not recommended.

Adjusting the split ratio will affect the yield of the cold products obtained and subsequently the economic cost analysis. After a split ratio is chosen, the desired portion of recycled methane gas is sent to be mixed with the natural gas feed stream.

5.0 RESULTS AND DISCUSSION

5.1 FISCHER-TROPSCH PROCESS SIMULATION RESULTS

5.1.1 F-T Process Operational Performance

The composition of the tailgas stream leaving the top of the 3-phase separator is shown in Figure 5-1. As can be seen, the tailgas primarily consists of C_1-C_2 hydrocarbons, H_2 and CO. All process stream names and flow rates for the F-T process simulations are provided in Tables Table A-1 to A-10 in Appendix A.

The simulation results indicated that increasing the syngas recycle ratios in the F-T process increase the overall C_5^+ yield by up to 86% for the 72% CO conversion case and by up to 69.5% for the 80% CO conversion case, as shown in Figure 5-2. The effect of tailgas recycle on increasing the C_{5+} yield was significantly higher when compared with simulation results obtained in slurry bubble column reactors [52]. Moreover, increasing the tail gas recycle ratios result in an enhancement of the yield of each liquid hydrocarbon, as can be seen in Figures 5-3 and 5-4.

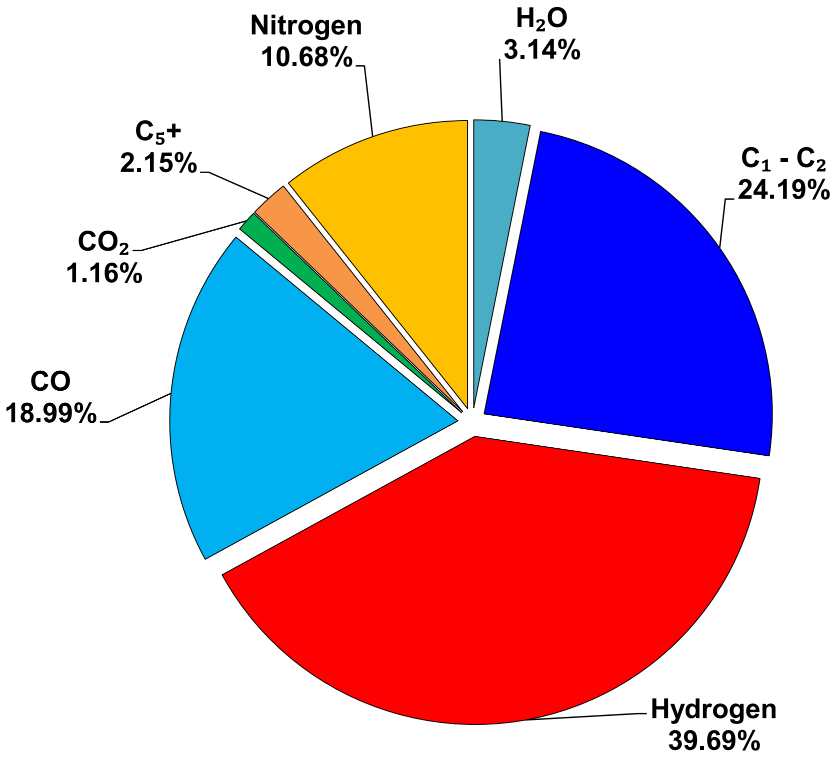


Figure 5-1: Tailgas composition (mole %)

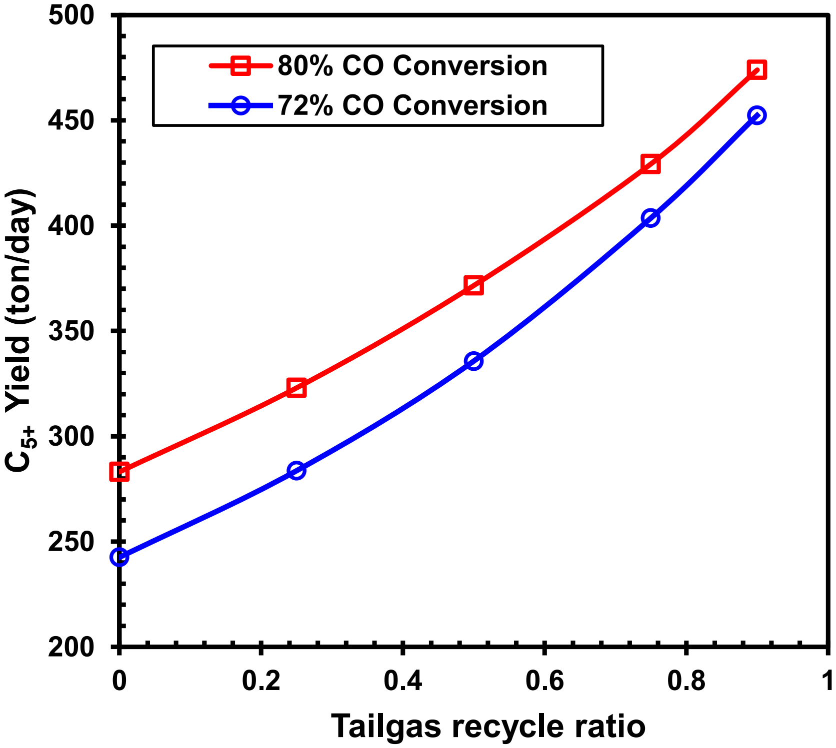


Figure 5-2: Effect of tailgas recycle ratio at 72% and 80% CO conversion on the yield of C_5^+

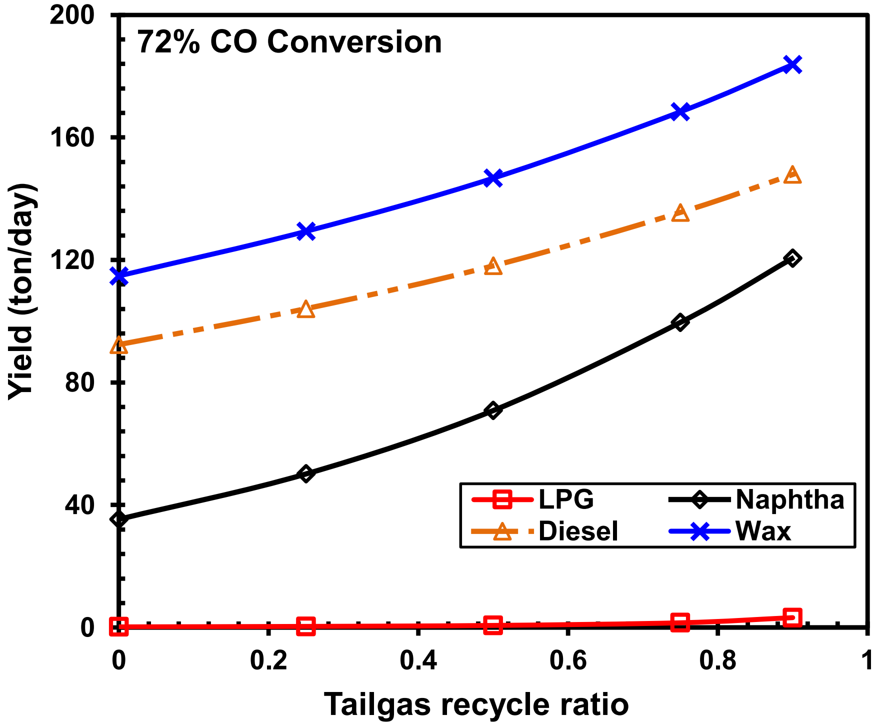


Figure 5-3: Effect of tailgas recycle ratio on the yield of synthetic hydrocarbon products at 72% of CO conversion

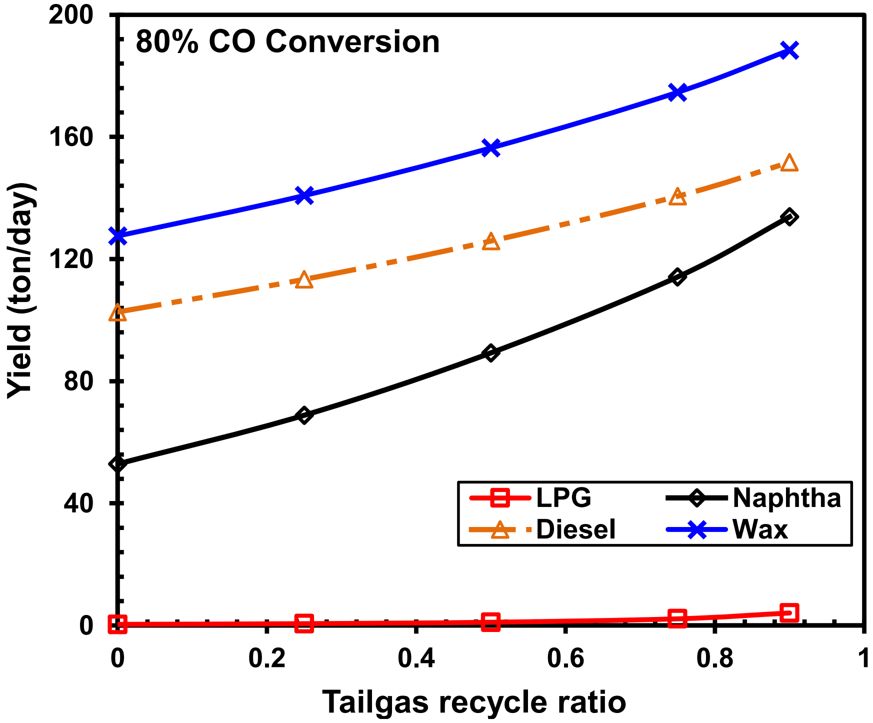


Figure 5-4: Effect of tailgas recycle ratio on the yield of synthetic hydrocarbon products at 80% of CO conversion

The effect of the tailgas recycle ratio on the energy requirements of different process units is shown in Figure 5-6; and as can be seen both the three-phase separator and the F-T MCR duties are negative, which means that there is excess energy to be removed from these units. On the other hand, the reformer duty is significantly endothermic, thus requiring a source of energy. It should be noted that both the natural gas and syngas compressor duties (electricity) are slightly positive, when compared with the major reaction or separation units. These results show that there is a significant potential for heat integration throughout the process.

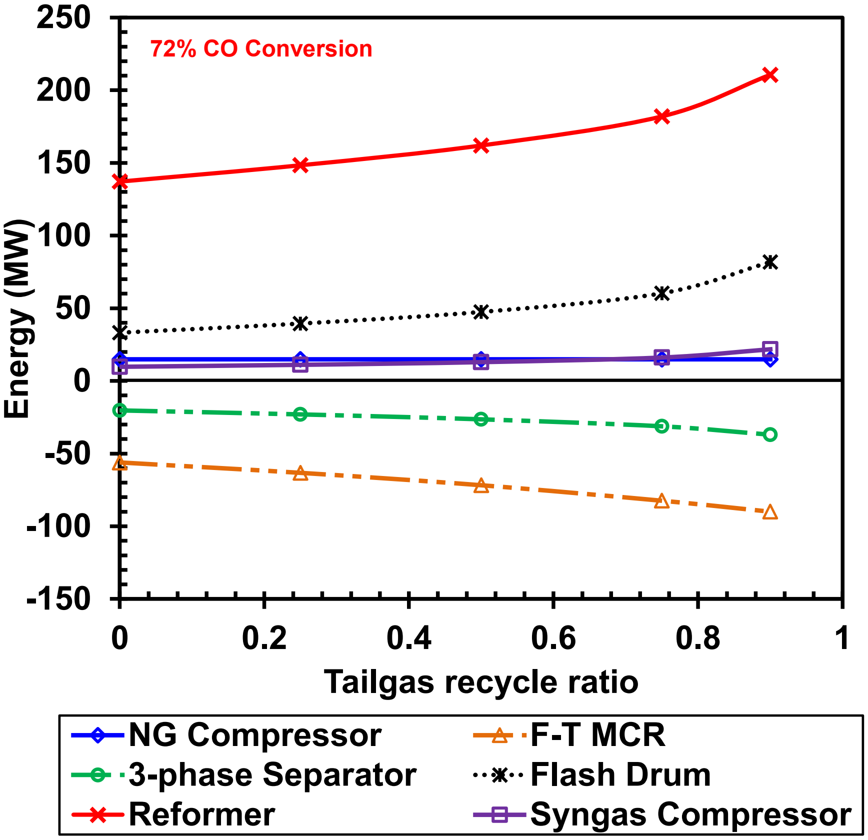


Figure 5-5: Effect of tailgas recycle ratio on the on Energy streams at 72% of CO conversion

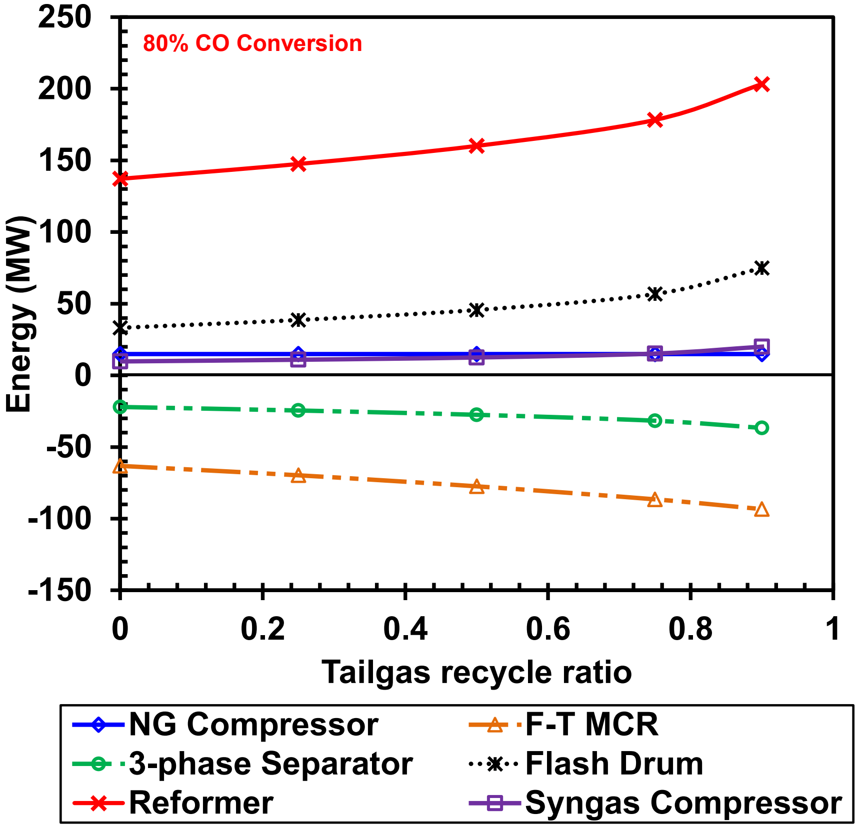


Figure 5-6: Effect of tailgas recycle ratio on the on Energy streams at 80% of CO conversion.

5.1.2 F-T Process Economic Assessment

Figure 5-7 shows the effect of the tailgas recycle ratio on the overall process cash flow, which represents the difference between the total annual plant investment and the total operating cost. As can be seen in this figure, increasing the tailgas recycle ratio increases the cash flow to the process, up to a recycle ratio of 75%; beyond which further increase of the tailgas recycle results in a slight decrease in the cash flow. The payback period (PBP) was determined for every recycle ratio, as shown in Figure 5-8; and as can be observed the payback period ranges from 7.0 to 4.5 years for all the tailgas recycle ratios used. The behavior is primarily due to the relatively high fixed capital investment, when compared with the cash flow, which reduces the impact of the tailgas recycle ratio on the time required to generate the return on investment.

The effects of the tailgas recycle ratios on the internal rate of return (IRR) are illustrated in Figure 5-9. As can be seen in this figure, the internal rate of return follows a similar trend to that of the cash flow (CF) shown in Figure 5-7. It is important to note that the rate of return has to be greater than 10% in order for the investment to be deemed feasible within the oil and gas industry [67], therefore, a syngas recycle ratio of 8% and 30% would be required in order to meet this criterion at CO conversions of 80 and 72%, respectively.

The effect of the tailgas recycle ratio on the plant net present value (NPV) is shown in Figure 5-10; and as can be concluded the NPV follows a similar trend to that of the cash flow and the payback period. Actually, the overall economic performance of the F-T process was found to be optimal around 75% tailgas recycle ratio, below which the yield of synthetic hydrocarbon products is too low, and above which the cost of utilities play an important role over the net cash flow of the plant.

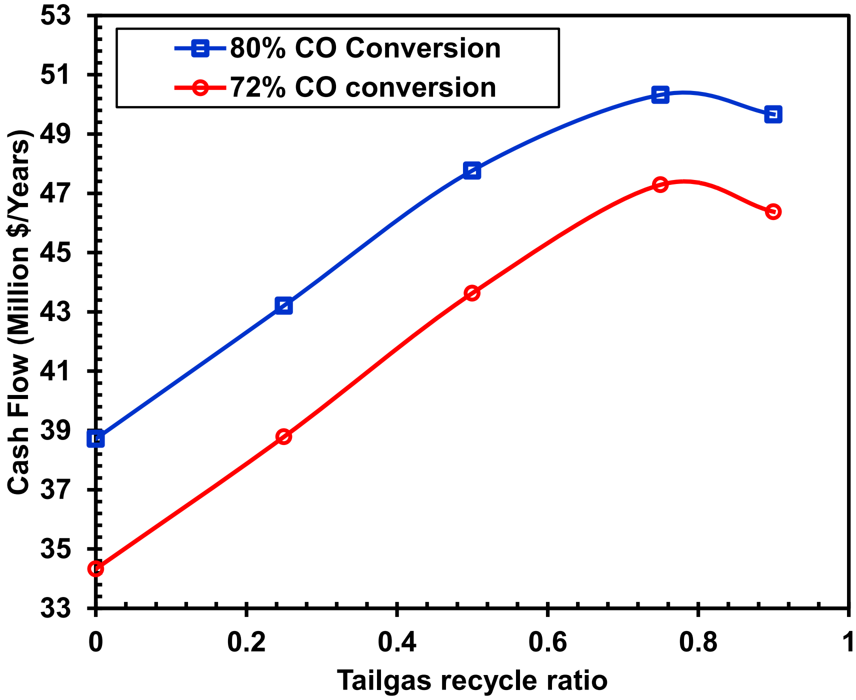


Figure 5-7: Effect of tailgas recycle ratio on the cash flow at 72% and 80% of CO conversion.

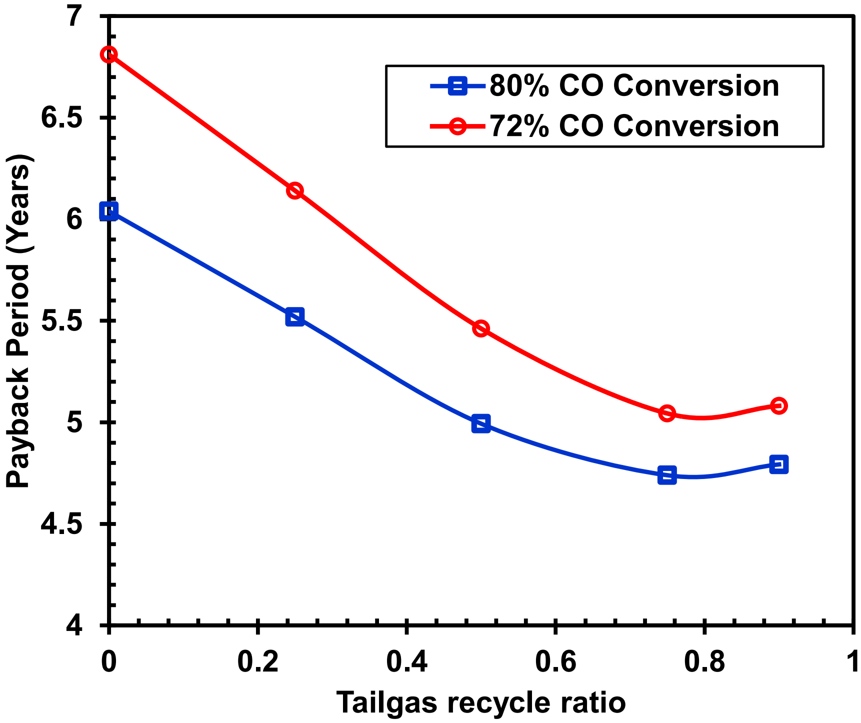


Figure 5-8: Effect of tailgas recycle ratio on the payback period

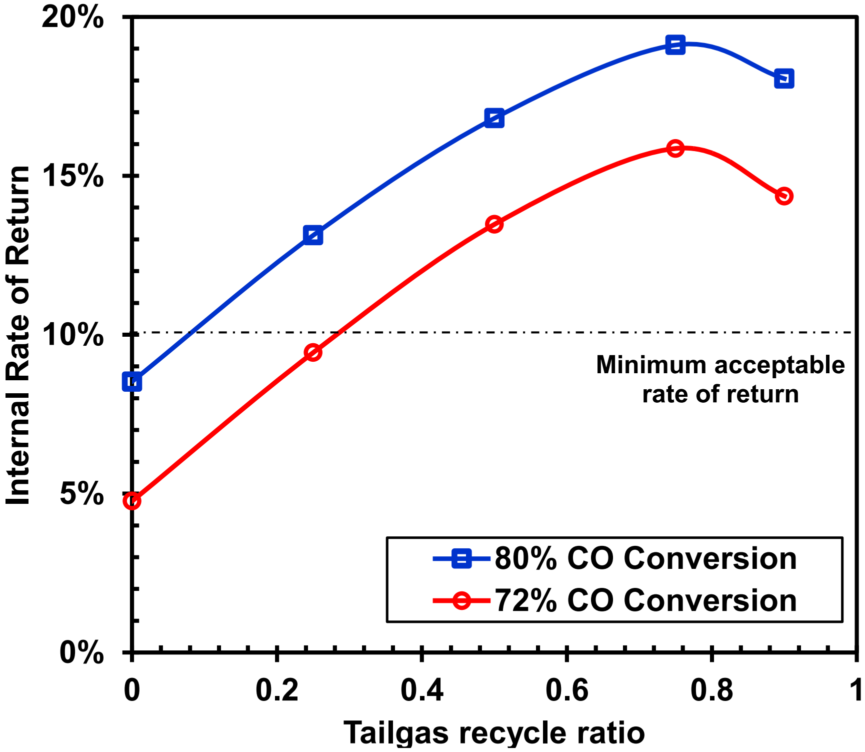


Figure 5-9: Effect of tailgas recycle ratio on the internal rate of return

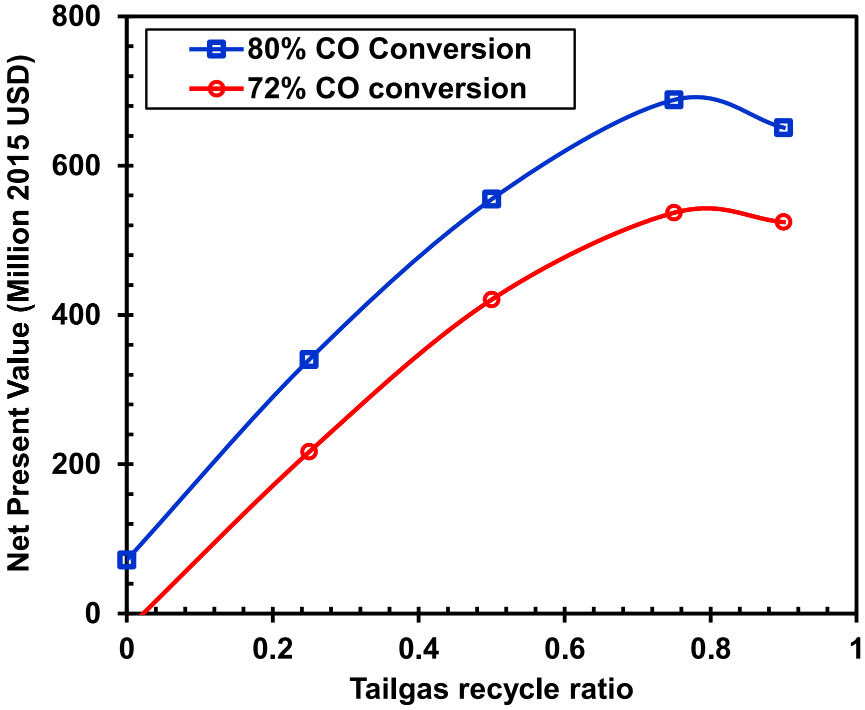


Figure 5-10: Effect of tailgas recycle ratio on the net present value

Figure 5-11 shows the effects of fluctuating the FCI and the cash flow rate on the internal rate of return without recycle at 80% CO Conversion inside the F-T MCR reactor. An increase in cash flow could be driven by a surge in oil price, whereas the opposite would happen when the remote gas has to be purchased. It was determined that an upsurge on cash inflow increases the internal rate of return, making the process more profitable throughout its lifetime, with no noticeable non-linearity at high cash flow fluctuations. Also, increasing the fixed capital investment, results in an increased reduction in the internal rate of return, making the process unprofitable at higher FCI requirements.

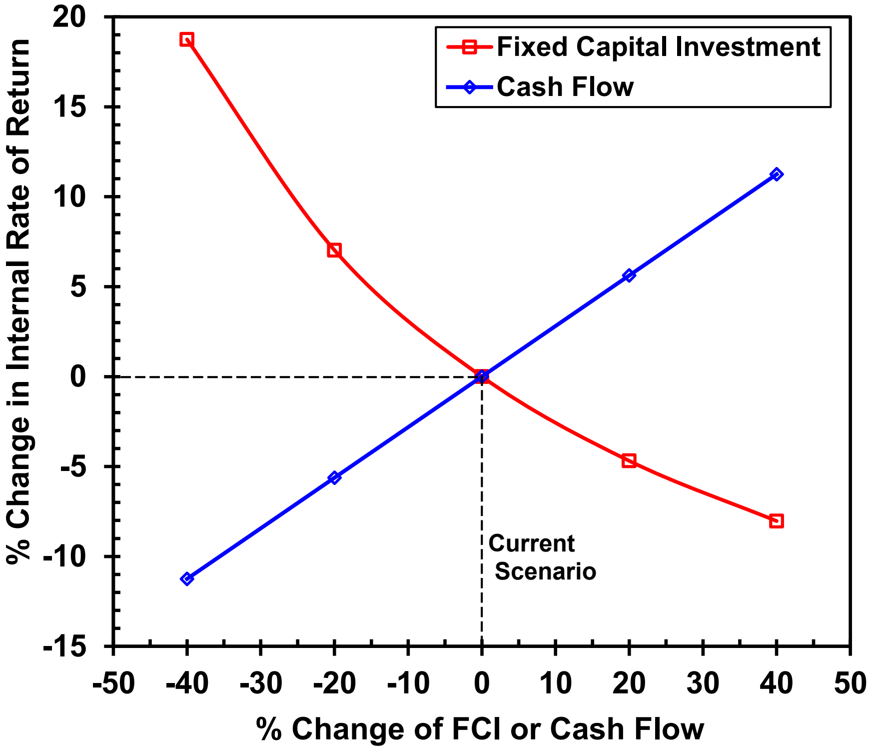


Figure 5-11: Economic sensitivity analysis for the F-T process (No recycle)

5.2 DMTM PROCESS SIMULATION RESULTS

5.2.1 DMTM Process Operational Performance

All the DMTM process stream names and flow rates for the DMTM process simulations are provided in Tables B-1 to B -10 in Appendix B. The simulation results indicated that increasing the methane recycle ratios in the DMTM process increase the overall product yield as shown in Figure 5-12. In addition, increasing the methane recycle ratio results in a more significant increase in the yield of methanol and formaldehyde when compared with that of ethanol. It should be mentioned that although high methane recycle ratios result in high yield of liquid products, increasing the recycle ratio demands more energy to compress, react, and separate the gas as shown in Figure 5-14.

The energy requirement accountable for separating the products (Splitter Duty) is considerably lower than the electricity consumed for compression (Compressor Duty) or the equivalent utility consumed by the DMTM reactor (DMTM Duty).

As the volumetric flow rate of the hot Recycle Gas increases, the temperature of the Gas + Recycle stream, and subsequently that of the (Air + Gas) will also increase, requiring different utility requirements. An estimate of the required utility can be determined based on the temperature difference between the inlet (Air + Gas) stream and that of the reactor temperature (500 °C).

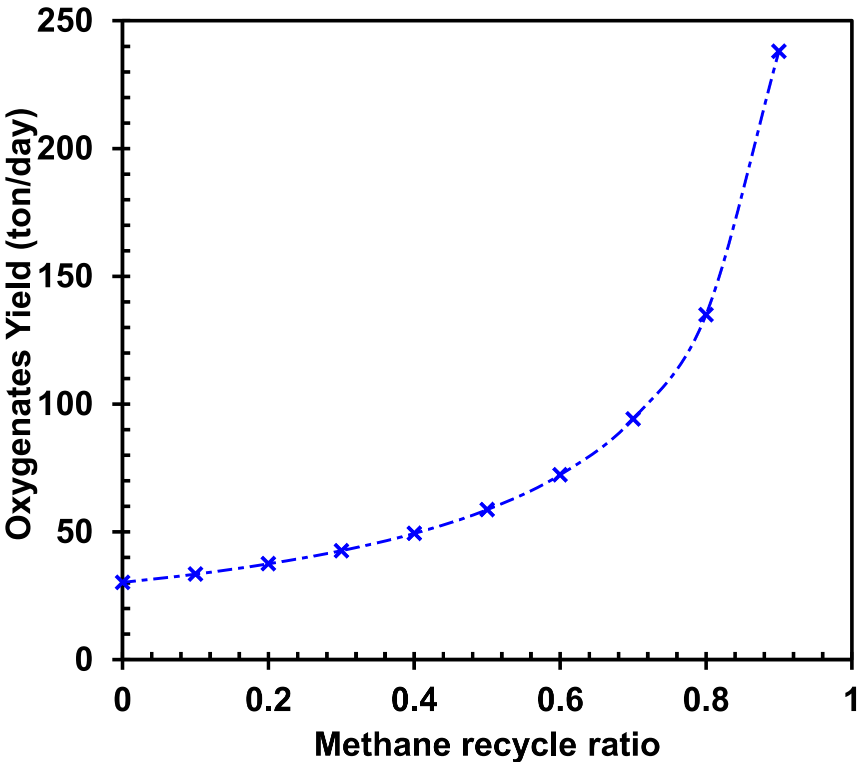


Figure 5-12: Effect of methane recycle ratio on the yield of oxygenates

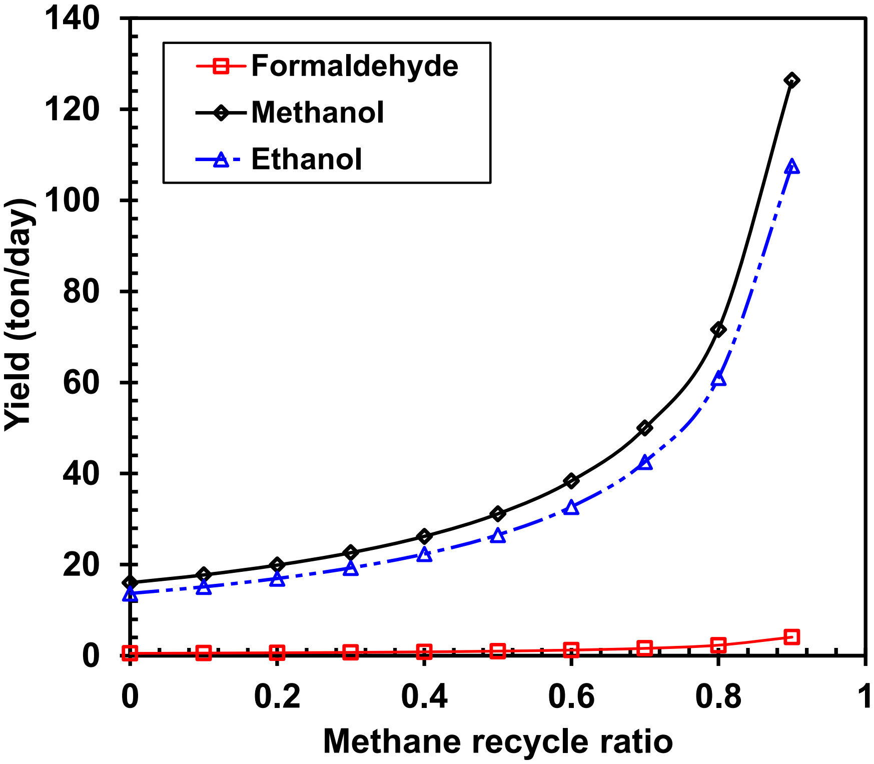


Figure 5-13: Effect of methane recycle ratio on the products yield

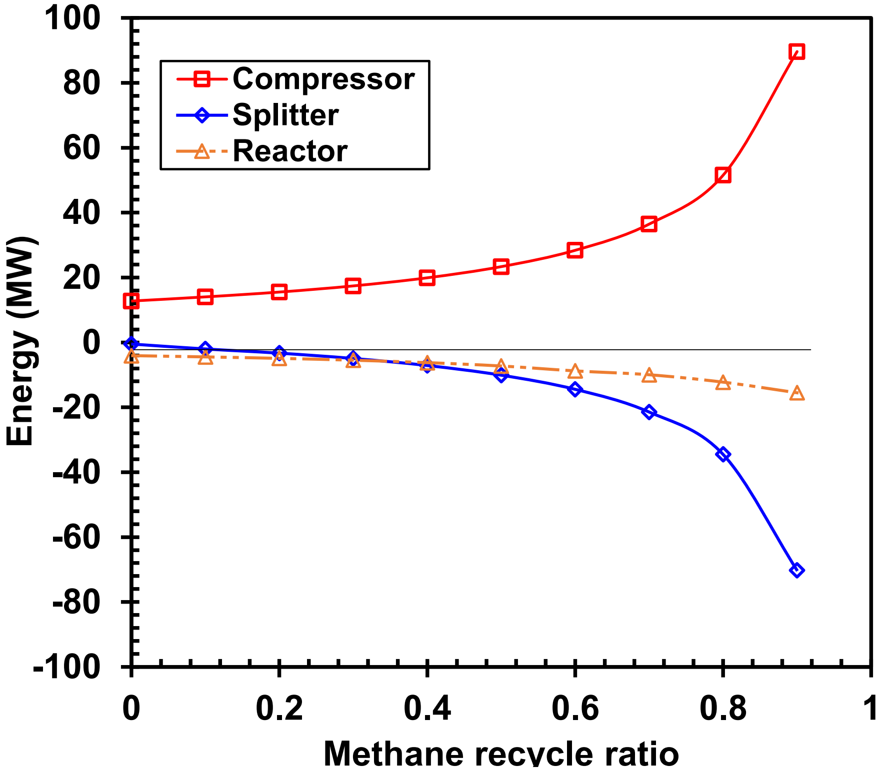


Figure 5-14: Effect of methane recycle ratio on the energy duty

5.2.2 DMTM Process Economic Assessment

Figure 5-15 shows the effect of methane recycle ratio on the overall process cash flow, which represents the difference between total annual plant investment and the total operating cost. As can be seen in this figure, the cash flow increases with increasing the recycle ratio, which is expected since a high product yield is achieved. It should be noted that these cash flow values were lower than those reported by GasTechno, who reported a cash flow of around 33-Million USD for their process [43].

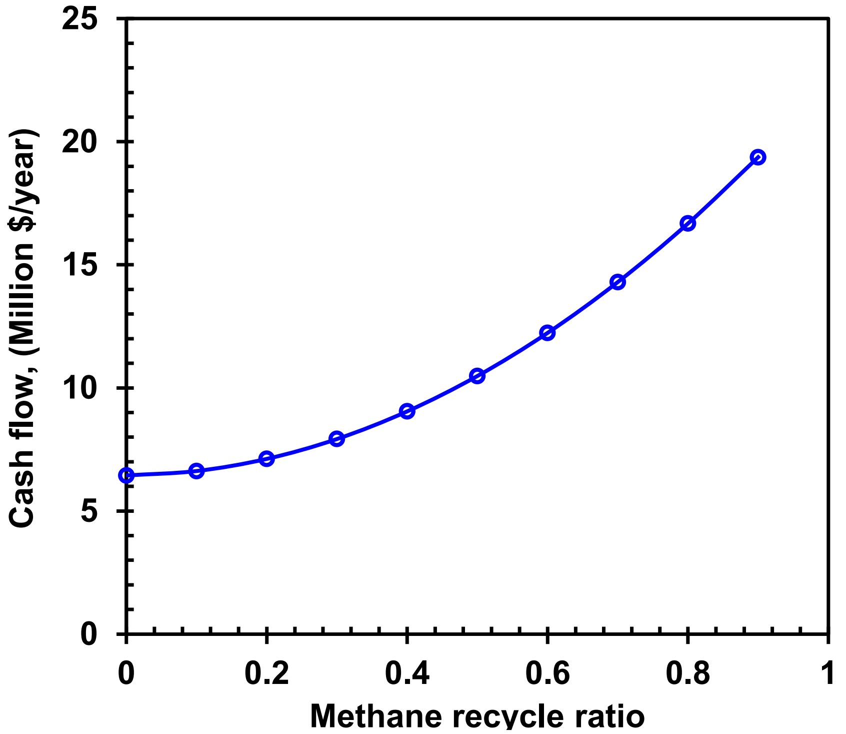


Figure 5-15: Effect of methane recycle ratio on the annual cash flow

The payback period was determined for every methane recycle ratio and the results are shown in Figure 5-16. As can be seen in this figure, increasing the methane recycle ratio decreases the payback period from 14 years up to a minimum of 1.5 year, which is in agreement with the results published by GasTechno for their DMTM process [43]. In addition, as shown in Figure 5-17, a methane recycle ratio of at least 55% is required in order to achieve a typical healthy internal rate of return of 10%.

Despite a positive annual cash flow, the overall feasibility is not proven before analyzing the cash flow throughout the plant's lifetime, represented using the NPV. Figure 5-18 shows the NPV of the plant over 20-year lifetime, and as can be seen, a minimum recycle ratio of 45% is required to maintain a positive NPV.

Generally, the DMTM process was determined to be profitable at a minimum methane recycle ratio of 45%, as it satisfies all the economic parameters. The results showed that higher methane recycle ratios led to more oxygenates production, which improved the profitability of the plant.

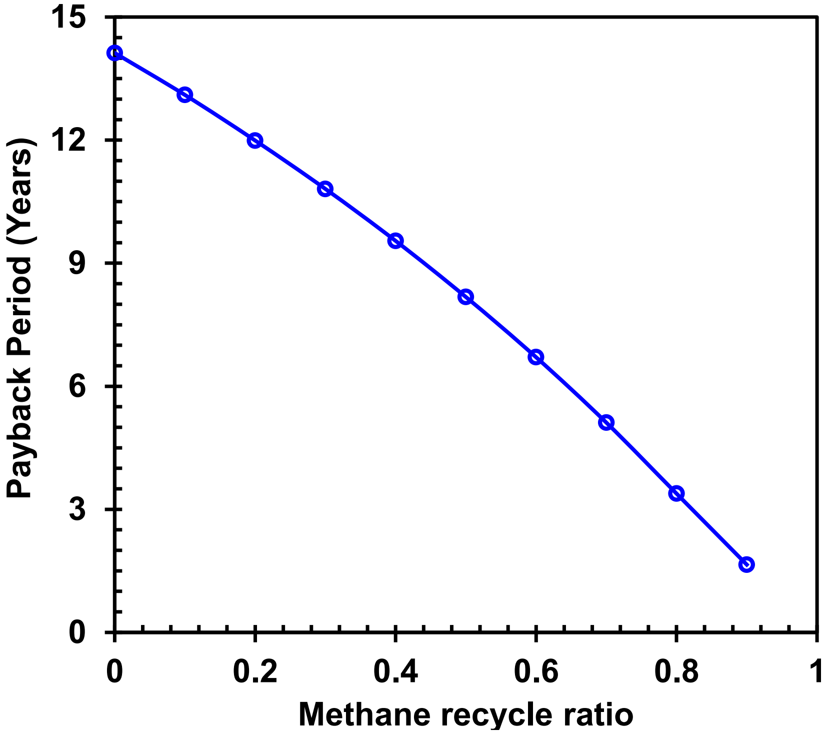


Figure 5-16: Effect of methane recycle ratio on the payback period

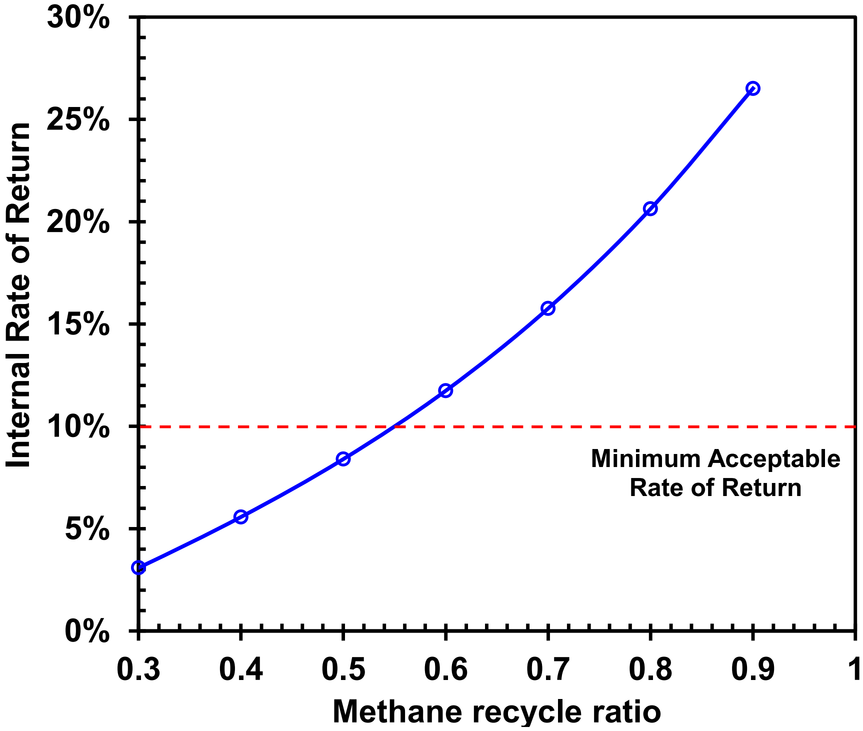


Figure 5-17: Effect of methane recycle ratio on the internal rate of return

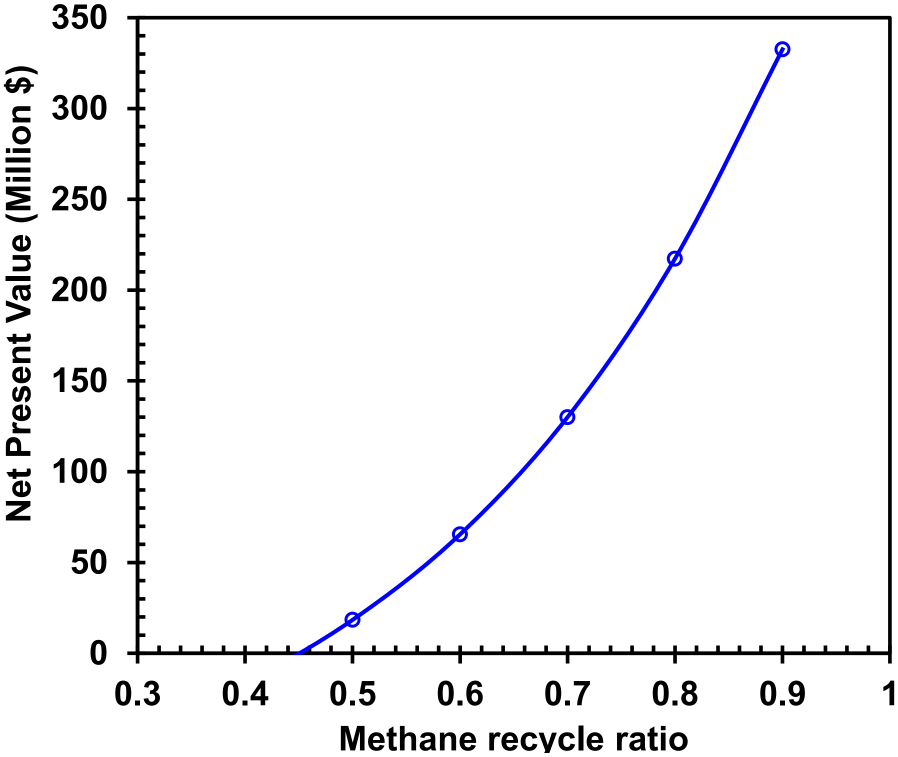


Figure 5-18: Effect of methane recycle ratio on the net present value

5.3 COMPARISON BETWEEN THE TWO PROCESSES

Figure 5-19 shows the net energy per capital invested for both processes, which was calculated by dividing the net process energy requirement by the required FCI. As can be seen in this figure, the DMTM process has a higher net energy per capita in comparison with that of the F-T MCR process at recycle ratios up to 55%, beyond which the net energy per capita for the DMTM process becomes lower than that of the F-T MCR. Moreover, when determining the net energy per product, which was determined by dividing the net process energy requirement by the overall products yield, the DMTM process has a significantly higher value at all recycle ratios, when compared to that of the F-T MCR. The higher net energy per product of the DMTM process compared with that of the F-T in MCR is in agreement with the analyses given by Klerk [41]. The significantly higher net energy per product for the DMTM processes is due the relatively high-energy utility requirements, such as oxidant compression, methane gas recycling and high temperature operation compared to the relatively lower yield to that of the F-T MCR. Although this makes the F-T in MCR more favorable, it should be noted that both technologies still have a relatively high net energy when compared with that of the conventional GTL technologies [15, 36], which is a major challenges facing the development of feasible modular technologies [41].

From an economic perspective, the profitability index, which is the ratio of the NPV to the FCI, is calculated for both processes according to Equation (5-1):

$$Profitability\ Index = \frac{Net\ Present\ Value}{Fixed\ Capital\ Inevstment} \tag{5-1}$$

The profitability index is a measure of the amount of value generated when compared with the original investment, and has to be greater than 1 for a project to be economically favorable [68].

Figure 5-21 shows the profitability index for both processes and as can be seen in this figure, the profitability index of the F-T MCR process is greater than 1 at recycle ratios of 15 and 25% for CO conversions of 80 and 72%, respectively. Whereas that for the DMTM is greater than 1 at recycle ratios of 55% and above.

Furthermore, the unit cost, which is the cost required to produce one ton of products, was determined for both processes and the data are presented in Figure 5-22. As can be seen in this figure, the unit cost for the F-T in MCR process is not very sensitive to the tail-gas recycle ratios, however, the DMTM is significantly more sensitive to the methane recycle ratio, which is primarily due to the lower conversion rates (3%). Nonetheless, at methane recycle ratios greater than 75%, the unit cost of the DMTM process becomes lower than that of the F-T MCR.

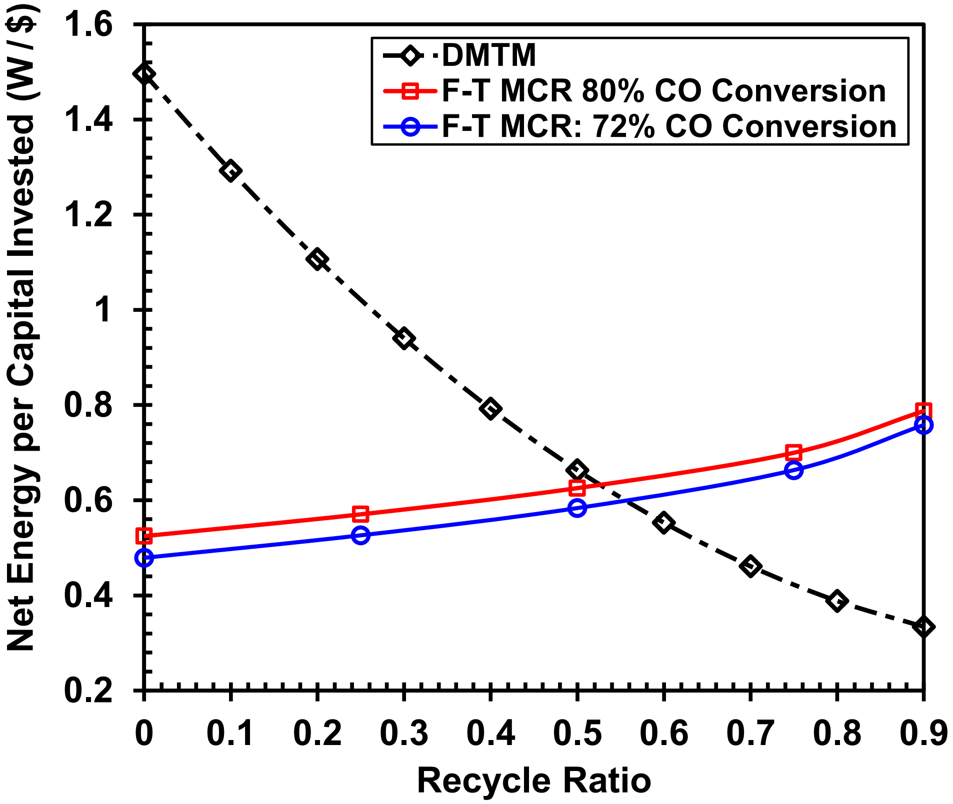


Figure 5-19: Comparison between the net energy per capital invested for both processes

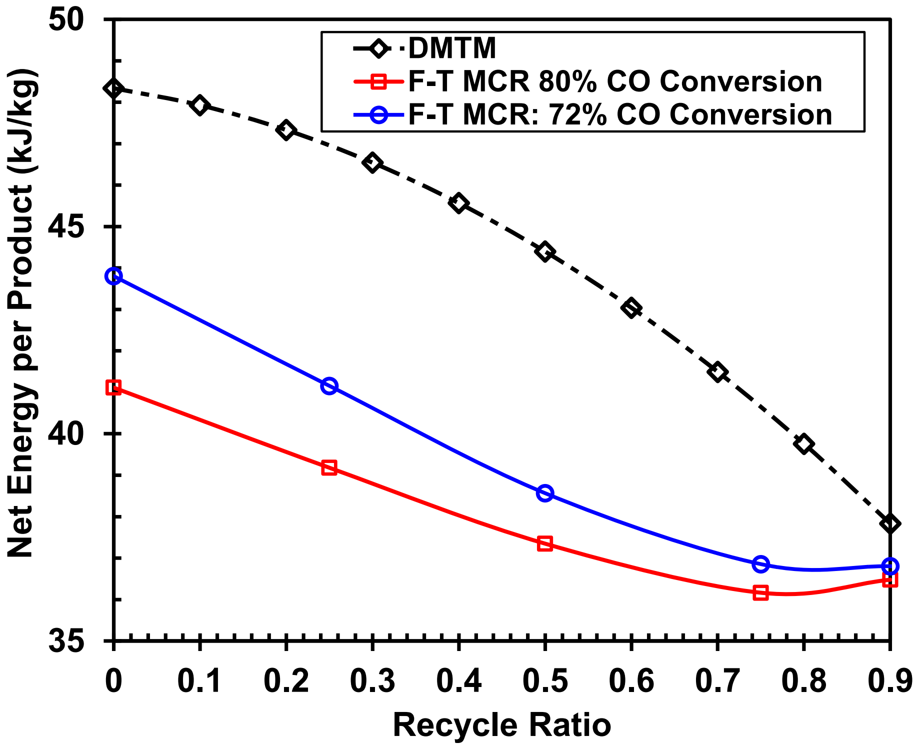


Figure 5-20: Comparison between the net energy per product for both processes

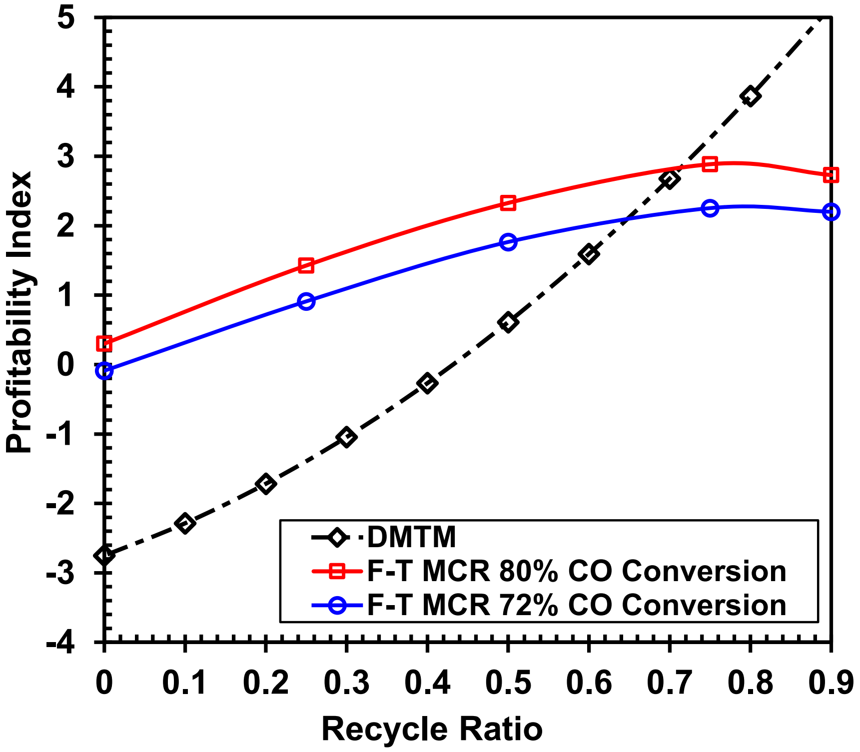


Figure 5-21: Comparison between the profitability index for both processes

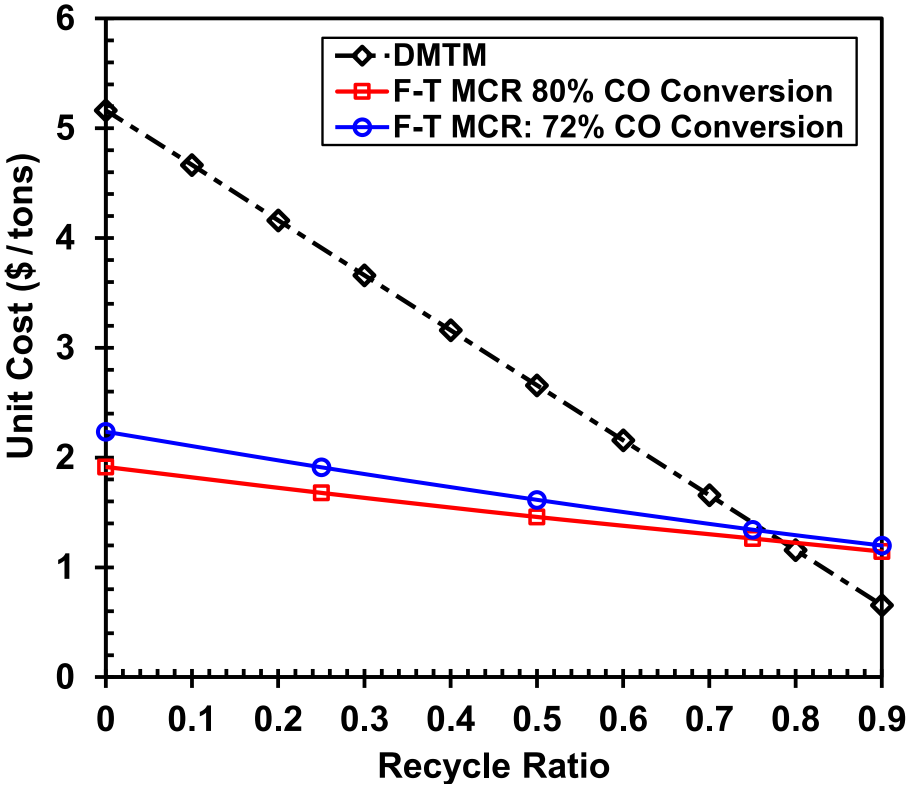


Figure 5-22: Comparison between the unit cost for both processes

6.0 CONCLUDING REMARKS

In this study, two processes, namely F-T in a MCR and DMTM were simulated and economically evaluated, using HYSYS v7.2, in order to determine their potential use in converting natural gas produced from small-scale remote gas or oil reservoirs into liquid products. The process operational performance and economic performance in terms NPV, IRR and PBP were calculated. The effect of the tail-gas recycle ratio in the F-T process and the methane recycle ratio in the DMTM process on the operational and economic performances of both processes were investigated. From the simulation results, the following concluding remarks can be made:

1. For the F-T in MCR process, in order to maintain an internal rate of return greater than the minimal acceptable 10%, the tail-gas recycle ratio had to be maintained higher than 8% and 30% at CO conversions of 80 and 72%, respectively. For the DMTM process, however, a minimum methane recycle ratio of 60% was required to achieve any profitability.
2. The products yields for the F-T in MCR process were found to be less sensitive to the tail-gas recycle ratios, whereas those of the DMTM process were more sensitive to the methane recycle ratio. This behavior was primarily due to the very low methane per-pass conversions of the DMTM process (3%) when compared with that of the F-T in MCR (72 and 80%).
3. The F-T in MCR process was found to have a profitability index greater than 1 at any tail-gas recycle ratio higher than 15 and 25% at CO conversions of 80 and 72%, respectively, whereas

the DMTM process required a minimum methane recycle ratio of 55% to achieve a favorable profitability index.

4. The unit cost (capital per product yield) of the DMTM process was found to be very sensitive to the methane recycle ratio as it could be reduced by up to 80% at higher values, whereas the unit cost of the F-T in MCR was found to be less insensitive to the recycle ratio.
5. The DMTM process was found to have a significantly higher net energy per capital invested and per product yield when compared with that of the F-T in MCR process; however, both processes were determined to have higher net energy requirements per capital and per yield when compared with that of the conventional GTL technologies.

APPENDIX A

F-T STREAM COMPOSITIONS

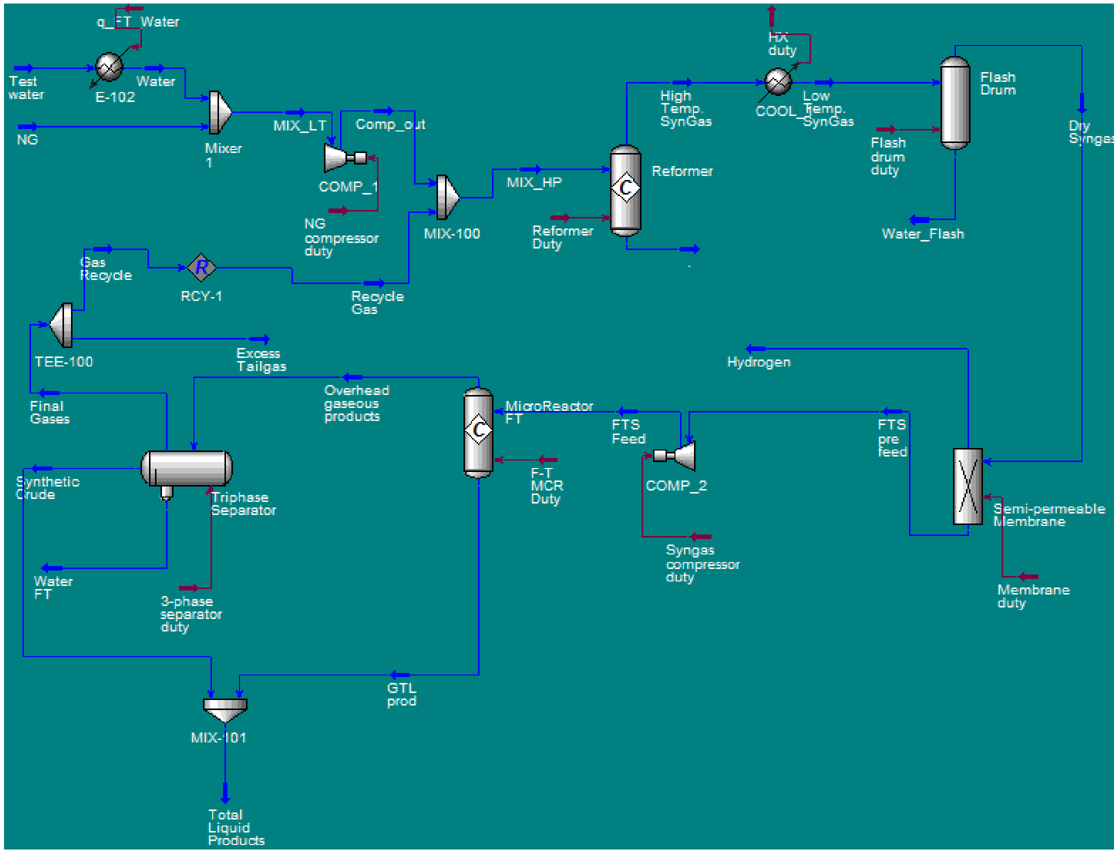


Figure A-1: F-T process flow diagram built in HYSYS

Table A-1: F-T Stream properties at 0% of tailgas recycle ratio and 80% of CO conversion

Name	**Vapor Fraction**	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Std Ideal Liq Vol Flow**
		^0C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	9345.51	103978.53	273.24
Low Temp. SynGas	1.00	100.00	3.50	9345.51	103978.53	273.24
Dry Syngas	1.00	25.00	3.50	7653.37	73494.35	242.69
Water_Flash	0.00	25.00	3.50	1692.14	30484.18	30.55
Hydrogen	1.00	25.00	0.73	1634.74	4410.11	46.42
FTS pre feed	1.00	25.00	20.00	6018.63	69084.24	196.27
FTS Feed	1.00	220.00	25.50	6018.63	69084.24	196.27
Overhead gaseous products	1.00	220.00	25.00	3200.26	59102.20	103.05
GTL prod	0.00	220.00	25.00	36.01	9979.91	12.57
MIX_HP	1.00	452.99	4.00	5867.55	103976.97	186.32
Final Gases	1.00	50.00	4.00	1830.47	29282.78	71.14
Synthetic Crude	0.00	50.00	4.00	38.06	5827.68	7.88
Water FT	0.00	50.00	4.00	1331.73	23991.74	24.04
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	0.00	0.00	0.00
Recycle Gas	1.00	50.00	4.00	0.00	0.00	0.00
Excess Tailgas	1.00	50.00	4.00	1830.47	29282.78	71.14
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.05	162.42	4.00	74.07	15807.59	20.45
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-2: F-T Stream properties at 25% of tailgas recycle ratio and 80% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	10042.73	112600.38	298.17
Low Temp. SynGas	1.00	100.00	3.50	10042.73	112600.38	298.17
Dry Syngas	1.00	25.00	3.50	8446.61	83845.75	269.36
Water_Flash	0.00	25.00	3.50	1596.12	28754.63	28.81
Hydrogen	1.00	25.00	0.73	1787.39	4833.83	50.75
FTS pre feed	1.00	25.00	22.65	6659.22	79011.92	218.61
FTS Feed	1.00	220.00	25.50	6659.22	79011.92	218.61
Overhead gaseous products	1.00	220.00	25.00	3547.84	67978.70	115.69
GTL prod	0.00	220.00	25.00	39.95	11030.86	13.90
MIX_HP	1.00	425.21	4.00	6375.29	112598.73	206.23
Final Gases	1.00	50.00	4.00	2030.97	34486.34	79.64
Synthetic Crude	0.00	50.00	4.00	47.17	7014.51	9.52
Water FT	0.00	50.00	4.00	1469.71	26477.84	26.53
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	507.74	8621.59	19.91
Recycle Gas	1.00	50.00	4.00	507.74	8621.77	19.91
Excess Tailgas	1.00	50.00	4.00	1523.23	25864.76	59.73
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.05	159.00	4.00	87.12	18045.38	23.42
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-3: F-T Stream properties at 50% of tailgas recycle ratio and 80% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	10909.17	125695.99	330.22
Low Temp. SynGas	1.00	100.00	3.50	10909.17	125695.99	330.22
Dry Syngas	1.00	25.00	3.50	9420.01	98868.09	303.34
Water_Flash	0.00	25.00	3.50	1489.16	26827.91	26.88
Hydrogen	1.00	25.00	0.71	1955.21	5314.86	55.50
FTS pre feed	1.00	25.00	20.00	7464.80	93553.23	247.84
FTS Feed	1.00	220.00	25.50	7464.80	93553.23	247.84
Overhead gaseous products	1.00	220.00	25.00	4010.91	81328.35	133.62
GTL prod	0.00	220.00	25.00	44.39	12222.25	15.40
MIX_HP	1.00	393.65	4.00	7028.79	125694.25	232.61
Final Gases	1.00	50.00	4.00	2322.69	43433.73	92.57
Synthetic Crude	0.00	50.00	4.00	58.94	8541.12	11.64
Water FT	0.00	50.00	4.00	1629.28	29353.50	29.41
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	1161.35	21716.86	46.29
Recycle Gas	1.00	50.00	4.00	1161.24	21717.29	46.28
Excess Tailgas	1.00	50.00	4.00	1161.35	21716.86	46.29
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	155.14	4.00	103.33	20763.38	27.04
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-4: F-T Stream properties at 75% of tailgas recycle ratio and 80% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	12231.76	154483.23	382.94
Low Temp. SynGas	1.00	100.00	3.50	12231.76	154483.23	382.94
Dry Syngas	1.00	25.00	3.50	10857.32	129721.24	358.13
Water_Flash	0.00	25.00	3.50	1374.44	24761.99	24.81
Hydrogen	1.00	25.00	0.67	2130.12	5879.68	60.40
FTS pre feed	1.00	25.00	20.00	8727.20	123841.56	297.73
FTS Feed	1.00	220.00	25.50	8727.20	123841.56	297.73
Overhead gaseous products	1.00	220.00	25.00	4869.51	110411.26	170.42
GTL prod	0.00	220.00	25.00	48.64	13427.37	16.92
MIX_HP	1.00	349.36	4.00	8109.80	154481.39	278.86
Final Gases	1.00	50.00	4.00	2986.71	67187.55	123.23
Synthetic Crude	0.00	50.00	4.00	74.93	10650.28	14.55
Water FT	0.00	50.00	4.00	1807.87	32573.44	32.64
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	2240.04	50390.66	92.42
Recycle Gas	1.00	50.00	4.00	2242.25	50504.42	92.53
Excess Tailgas	1.00	50.00	4.00	746.68	16796.89	30.81
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	149.90	4.00	123.58	24077.65	31.47
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-5: F-T Stream properties at 90% of tailgas recycle ratio and 80% of CO conversion

Name	**Vapor Fraction**	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Std Ideal Liq Vol Flow**
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	14287.95	217301.69	472.41
Low Temp. SynGas	1.00	100.00	3.50	14287.95	217301.69	472.41
Dry Syngas	1.00	25.00	3.50	12970.26	193557.75	448.62
Water_Flash	0.00	25.00	3.50	1317.70	23743.94	23.79
Hydrogen	1.00	25.00	0.59	2232.58	6401.28	63.14
FTS pre feed	1.00	25.00	20.00	10737.67	187156.47	385.47
FTS Feed	1.00	220.00	25.50	10737.67	187156.47	385.47
Overhead gaseous products	1.00	220.00	25.00	6578.42	173385.33	249.02
GTL prod	0.00	220.00	25.00	49.04	13767.99	17.33
MIX_HP	1.00	291.80	4.00	9995.28	217299.77	363.71
Final Gases	1.00	50.00	4.00	4579.40	126084.04	196.92
Synthetic Crude	0.00	50.00	4.00	91.00	12919.70	17.65
Water FT	0.00	50.00	4.00	1908.03	34381.59	34.45
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	4121.46	113475.63	177.23
Recycle Gas	1.00	50.00	4.00	4127.73	113322.80	177.39
Excess Tailgas	1.00	50.00	4.00	457.94	12608.40	19.69
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	142.96	4.00	140.03	26687.68	34.97
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-6: F-T Stream properties at 0% of tailgas recycle ratio and 72% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	9345.51	103978.53	273.24
Low Temp. SynGas	1.00	100.00	3.50	9345.51	103978.53	273.24
Dry Syngas	1.00	25.00	3.50	7653.37	73494.35	242.69
Water_Flash	0.00	25.00	3.50	1692.14	30484.18	30.55
Hydrogen	1.00	25.00	0.73	1634.74	4410.11	46.42
FTS pre feed	1.00	25.00	20.00	6018.63	69084.24	196.27
FTS Feed	1.00	220.00	25.50	6018.63	69084.24	196.27
Overhead gaseous products	1.00	220.00	25.00	3512.18	60461.62	113.66
GTL prod	0.00	220.00	25.00	30.32	8620.72	10.85
MIX_HP	1.00	452.99	4.00	5867.55	103976.97	186.32
Final Gases	1.00	50.00	4.00	2318.42	34583.25	86.08
Synthetic Crude	0.00	50.00	4.00	29.82	4909.55	6.57
Water FT	0.00	50.00	4.00	1163.94	20968.81	21.01
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	0.00	0.00	0.00
Recycle Gas	1.00	50.00	4.00	0.00	0.00	0.00
Excess Tailgas	1.00	50.00	4.00	2318.42	34583.25	86.08
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	163.88	4.00	60.14	13530.27	17.42
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-7: F-T Stream properties at 25% of tailgas recycle ratio and 72% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	10183.51	114263.84	302.72
Low Temp. SynGas	1.00	100.00	3.50	10183.51	114263.84	302.72
Dry Syngas	1.00	25.00	3.50	8588.40	85527.43	273.93
Water_Flash	0.00	25.00	3.50	1595.11	28736.41	28.79
Hydrogen	1.00	25.00	0.73	1816.91	4914.04	51.59
FTS pre feed	1.00	25.00	20.00	6771.49	80613.39	222.34
FTS Feed	1.00	220.00	25.50	6771.49	80613.39	222.34
Overhead gaseous products	1.00	220.00	25.00	3946.74	70869.56	129.21
GTL prod	0.00	220.00	25.00	34.42	9741.68	12.26
MIX_HP	1.00	419.53	4.00	6517.15	114262.20	210.67
Final Gases	1.00	50.00	4.00	2596.10	41142.16	97.33
Synthetic Crude	0.00	50.00	4.00	38.57	6089.60	8.19
Water FT	0.00	50.00	4.00	1312.07	23637.79	23.69
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	649.03	10285.54	24.33
Recycle Gas	1.00	50.00	4.00	649.60	10285.23	24.35
Excess Tailgas	1.00	50.00	4.00	1947.08	30856.62	72.99
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	160.13	4.00	72.99	15831.28	20.46
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-8: F-T Stream properties at 50% of tailgas recycle ratio and 72% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	11215.87	129711.66	340.36
Low Temp. SynGas	1.00	100.00	3.50	11215.87	129711.66	340.36
Dry Syngas	1.00	25.00	3.50	9728.11	102908.82	313.50
Water_Flash	0.00	25.00	3.50	1487.76	26802.84	26.86
Hydrogen	1.00	25.00	0.71	2016.77	5484.01	57.24
FTS pre feed	1.00	25.00	20.00	7711.35	97424.81	256.26
FTS Feed	1.00	220.00	25.50	7711.35	97424.81	256.26
Overhead gaseous products	1.00	220.00	25.00	4507.41	86369.06	150.62
GTL prod	0.00	220.00	25.00	39.23	11053.31	13.92
MIX_HP	1.00	383.08	4.00	7340.73	129709.92	242.55
Final Gases	1.00	50.00	4.00	2969.48	51876.07	113.35
Synthetic Crude	0.00	50.00	4.00	50.73	7699.29	10.42
Water FT	0.00	50.00	4.00	1487.20	26793.69	26.85
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	1484.74	25938.03	56.68
Recycle Gas	1.00	50.00	4.00	1473.18	25732.96	56.23
Excess Tailgas	1.00	50.00	4.00	1484.74	25938.03	56.68
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	155.68	4.00	89.96	18752.60	24.33
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-9: F-T Stream properties at 75% of tailgas recycle ratio and 72% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	12801.03	163403.01	402.61
Low Temp. SynGas	1.00	100.00	3.50	12801.03	163403.01	402.61
Dry Syngas	1.00	25.00	3.50	11432.21	138741.21	377.90
Water_Flash	0.00	25.00	3.50	1368.82	24661.80	24.71
Hydrogen	1.00	25.00	0.67	2234.21	6173.84	63.35
FTS pre feed	1.00	25.00	20.00	9198.00	132567.37	314.55
FTS Feed	1.00	220.00	25.50	9198.00	132567.37	314.55
Overhead gaseous products	1.00	220.00	25.00	5521.06	120048.45	193.52
GTL prod	0.00	220.00	25.00	44.43	12516.11	15.75
MIX_HP	1.00	334.36	4.00	8682.25	163401.18	298.02
Final Gases	1.00	50.00	4.00	3756.54	79409.46	149.20
Synthetic Crude	0.00	50.00	4.00	68.62	10082.81	13.70
Water FT	0.00	50.00	4.00	1695.90	30556.19	30.62
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	2817.40	59557.10	111.90
Recycle Gas	1.00	50.00	4.00	2814.70	59424.21	111.70
Excess Tailgas	1.00	50.00	4.00	939.13	19852.37	37.30
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	149.62	4.00	113.05	22598.92	29.46
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

Table A-10: F-T Stream properties at 90% of tailgas recycle ratio and 72% of CO conversion

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Std Ideal Liq Vol Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
MIX_LT	1.00	239.65	1.00	5867.55	103976.97	186.32
High Temp. SynGas	1.00	800.00	4.00	15258.97	240102.44	510.51
Low Temp. SynGas	1.00	100.00	3.50	15258.97	240102.44	510.51
Dry Syngas	1.00	25.00	3.50	13938.79	216309.85	486.66
Water_Flash	0.00	25.00	3.50	1320.17	23792.59	23.84
Hydrogen	1.00	25.00	0.57	2353.06	6787.43	66.52
FTS pre feed	1.00	25.00	20.00	11585.73	209522.42	420.14
FTS Feed	1.00	220.00	25.50	11585.73	209522.42	420.14
Overhead gaseous products	1.00	220.00	25.00	7575.92	196580.43	288.97
GTL prod	0.00	220.00	25.00	45.20	12938.93	16.26
MIX_HP	1.00	273.35	4.00	10978.13	240100.53	401.36
Final Gases	1.00	50.00	4.00	5689.99	151644.67	239.47
Synthetic Crude	0.00	50.00	4.00	85.44	12490.92	16.98
Water FT	0.00	50.00	4.00	1800.49	32444.83	32.51
Test water	0.00	25.00	1.00	3500.00	63052.85	63.18
Water	1.00	400.00	1.00	3500.00	63052.85	63.18
Gas Recycle	1.00	50.00	4.00	5120.99	136480.20	215.53
Recycle Gas	1.00	50.00	4.00	5110.58	136123.57	215.03
Excess Tailgas	1.00	50.00	4.00	569.00	15164.47	23.95
NG	1.00	25.00	1.00	2367.55	40924.12	123.14
Total Liquid Products	0.04	142.07	4.00	130.65	25429.85	33.25
Comp_out	1.00	452.99	4.00	5867.55	103976.97	186.32

APPENDIX B

DMTM STREAM COMPOSITIONS

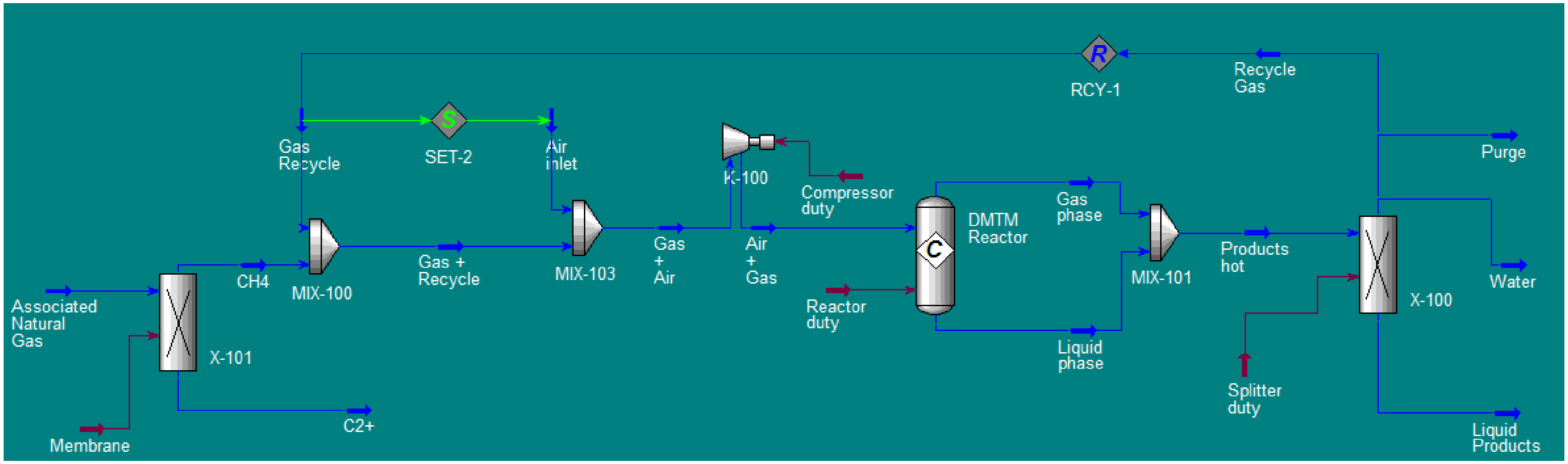


Figure B-1: DMTM process flow diagram built in HYSYS

Table B-1: DMTM Stream properties at 0% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	1361.24	21838.28	72.94
Gas phase	1.00	500.00	80.01	1894.34	37545.46	90.04
Recycle Gas	0.98	25.00	1.00	0.00	0.00	0.00
Gas + Air	1.00	25.00	1.00	1905.68	37545.52	91.10
Products hot	1.00	500.00	80.01	1894.34	37545.46	90.04
Liquid Products	0.00	25.00	1.00	40.15	1259.54	1.62
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	0.98	25.00	1.00	0.00	0.00	0.00
Purge	1.00	25.00	1.00	1835.35	35946.65	88.08
Air + Gas	1.00	563.28	80.50	1905.68	37545.52	91.10
Air inlet	1.00	25.00	1.01	544.44	15707.24	18.16
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	18.83	339.27	0.34
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -2: DMTM Stream properties at 10% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	1507.47	24184.14	80.78
Gas phase	1.00	500.00	80.01	2085.82	41232.29	99.32
Recycle Gas	1.00	25.00	1.00	146.22	2345.86	7.84
Gas + Air	1.00	25.00	1.00	2098.39	41232.35	100.48
Products hot	1.00	500.00	80.01	2085.82	41232.29	99.32
Liquid Products	1.00	25.00	1.00	1214.26	20161.73	64.48
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	146.22	2345.86	7.84
Purge	1.00	25.00	1.00	704.49	18348.97	26.62
Air + Gas	1.00	562.26	80.50	2098.39	41232.35	100.48
Air inlet	1.00	25.00	1.01	590.92	17048.21	19.71
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	20.86	375.72	0.38
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -3: DMTM Stream properties at 20% of methane recycle ratio

Name	**Vapor Fraction**	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Volumetric Flow**
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	1688.89	27094.64	90.50
Gas phase	1.00	500.00	80.01	2309.63	45409.41	110.36
Recycle Gas	1.00	25.00	1.00	327.64	5256.36	17.56
Gas + Air	1.00	25.00	1.00	2323.71	45409.48	111.67
Products hot	1.00	500.00	80.01	2309.63	45409.41	110.36
Liquid Products	0.00	25.00	1.00	49.81	1562.70	2.01
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	327.64	5256.36	17.56
Purge	1.00	25.00	1.00	1908.81	38169.41	90.37
Air + Gas	1.00	560.16	80.50	2323.71	45409.48	111.67
Air inlet	1.00	25.00	1.01	634.82	18314.84	21.17
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	23.37	420.94	0.42
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -4: DMTM Stream properties at 30% of methane recycle ratio

Name	**Vapor Fraction**	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Volumetric Flow**
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	1919.95	30801.52	102.88
Gas phase	1.00	500.00	80.01	2594.69	50729.49	124.43
Recycle Gas	1.00	25.00	1.00	558.70	8963.24	29.94
Gas + Air	1.00	25.00	1.00	2610.69	50729.57	125.92
Products hot	1.00	500.00	80.01	2594.69	50729.49	124.43
Liquid Products	0.00	25.00	1.00	56.63	1776.50	2.29
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	558.70	8963.24	29.94
Purge	1.00	25.00	1.00	1952.79	39511.22	91.72
Air + Gas	1.00	558.04	80.50	2610.69	50729.57	125.92
Air inlet	1.00	25.00	1.01	690.74	19928.05	23.04
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	26.56	478.53	0.48
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -5: DMTM Stream properties at 40% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	2224.25	35683.47	119.19
Gas phase	1.00	500.00	80.01	2970.10	57736.01	142.95
Recycle Gas	1.00	25.00	1.00	863.01	13845.18	46.24
Gas + Air	1.00	25.00	1.00	2988.64	57736.10	144.68
Products hot	1.00	500.00	80.01	2970.10	57736.01	142.95
Liquid Products	0.00	25.00	1.00	65.60	2058.07	2.65
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	863.01	13845.18	46.24
Purge	1.00	25.00	1.00	2010.71	41278.39	93.50
Air + Gas	1.00	555.89	80.50	2988.64	57736.10	144.68
Air inlet	1.00	25.00	1.01	764.38	22052.64	25.49
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	30.77	554.37	0.56
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -6: DMTM Stream properties at 50% of methane recycle ratio

Name	**Vapor Fraction**	**Temperature**	**Pressure**	**Molar Flow**	**Mass Flow**	**Volumetric Flow**
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	2643.19	42404.43	141.63
Gas phase	1.00	500.00	80.01	3486.92	67381.87	168.46
Recycle Gas	1.00	25.00	1.00	1281.95	20566.15	68.69
Gas + Air	1.00	25.00	1.00	3508.95	67381.99	170.51
Products hot	1.00	500.00	80.01	3486.92	67381.87	168.46
Liquid Products	0.00	25.00	1.00	77.96	2445.70	3.15
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	1281.95	20566.15	68.69
Purge	1.00	25.00	1.00	2090.45	43711.24	95.96
Air + Gas	1.00	553.72	80.50	3508.95	67381.99	170.51
Air inlet	1.00	25.00	1.01	865.77	24977.56	28.87
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	36.57	658.79	0.66
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -7: DMTM Stream properties at 60% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	3256.56	52244.69	174.50
Gas phase	1.00	500.00	80.01	4243.62	81504.53	205.80
Recycle Gas	1.00	25.00	1.00	1895.32	30406.41	101.56
Gas + Air	1.00	25.00	1.00	4270.76	81504.67	208.33
Products hot	1.00	500.00	80.01	4243.62	81504.53	205.80
Liquid Products	0.00	25.00	1.00	96.05	3013.25	3.88
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	1895.32	30406.41	101.56
Purge	1.00	25.00	1.00	2207.19	47273.21	99.55
Air + Gas	1.00	551.51	80.50	4270.76	81504.67	208.33
Air inlet	1.00	25.00	1.01	1014.20	29259.98	33.82
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	45.05	811.66	0.81
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -8: DMTM Stream properties at 70% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	4240.63	68032.03	227.23
Gas phase	1.00	500.00	80.01	5457.63	104162.37	265.71
Recycle Gas	1.00	25.00	1.00	2879.39	46193.75	154.29
Gas + Air	1.00	25.00	1.00	5492.98	104162.55	269.00
Products hot	1.00	500.00	80.01	5457.63	104162.37	265.71
Liquid Products	0.00	25.00	1.00	125.08	3923.79	5.05
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	2879.39	46193.75	154.29
Purge	1.00	25.00	1.00	2394.50	52987.90	105.31
Air + Gas	1.00	549.28	80.50	5492.98	104162.55	269.00
Air inlet	1.00	25.00	1.01	1252.35	36130.52	41.77
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	58.67	1056.93	1.06
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -9: DMTM Stream properties at 80% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	6076.98	97492.32	325.63
Gas phase	1.00	500.00	80.01	7723.07	146443.51	377.51
Recycle Gas	1.00	25.00	1.00	4715.73	75654.04	252.69
Gas + Air	1.00	25.00	1.00	7773.72	146443.77	382.22
Products hot	1.00	500.00	80.01	7723.07	146443.51	377.51
Liquid Products	0.00	25.00	1.00	179.24	5622.94	7.24
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	4715.73	75654.04	252.69
Purge	1.00	25.00	1.00	2744.02	63651.91	116.06
Air + Gas	1.00	547.01	80.50	7773.72	146443.77	382.22
Air inlet	1.00	25.00	1.01	1696.74	48951.45	56.59
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	84.07	1514.62	1.52
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

Table B -10: DMTM Stream properties at 90% of methane recycle ratio

Name	Vapor Fraction	Temperature	Pressure	Molar Flow	Mass Flow	Volumetric Flow
		^{0}C	bar	kmol/h	kg/h	m^3/h
Gas + Recycle	1.00	25.00	1.00	10718.45	171954.96	574.34
Gas phase	1.00	500.00	80.01	13449.09	253311.61	660.08
Recycle Gas	1.00	25.00	1.00	9357.20	150116.68	501.40
Gas + Air	1.00	25.00	1.00	13538.42	253312.07	668.39
Products hot	1.00	500.00	80.01	13449.09	253311.61	660.08
Liquid Products	0.00	25.00	1.00	316.14	9917.62	12.77
Associated Natural Gas	1.00	25.00	1.00	2360.40	59319.43	153.61
Gas Recycle	1.00	25.00	1.00	9357.20	150116.68	501.40
Purge	1.00	25.00	1.00	3627.45	90605.85	143.23
Air + Gas	1.00	544.72	80.50	13538.42	253312.07	668.39
Air inlet	1.00	25.00	1.01	2819.98	81357.11	94.05
Liquid phase	0.00	500.00	80.01	0.00	0.00	0.00
Water	0.00	25.00	1.00	148.29	2671.45	2.68
CH4	1.00	25.00	1.00	1361.24	21838.28	72.94
C2+	1.00	25.00	1.00	999.16	37481.15	80.67

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