Table of Contents

List of Tables

List of Figures

Summary

The main goal of the project was to design and simulate a gas to liquid plant which should produce at least 17 000 bbl/day of products. Further an optimization of the variables should be done, and also a pinch analysis to make the plant as thermal efficient as possible. An economic analysis was to be done to get a quick estimate on the costs and profitability of such a process. At last a sensitivity analysis should be done too see how the profitability changes with different parameters such as natural gas price.

The simulation tool used for this project was Aspen HYSYS® version 7.2, with the Peng-Robinson equation of state. For the economics the method described in Sinnot and Towler (1) was used, in addition to Microsoft Excel. The optimization was done by comparing different cases in the simulation.

After the optimization was finished, the optimal purge ratio was found to be 3.5 %, and the optimal recycle ratio back to the Fischer-Tropsch reactor was found to be 96 %. The optimal reactor volume was found to be 2700 m^3. The plant was very profitable at today's (2011) product prices with a natural gas price of 0.5 $/MMbtu. With an interest rate of 9 %, and a lifetime of 20 years, the net present value of the project was 4.6 billion dollars, the internal rate of return 106 %, and the payback time was slightly over one year. The plant was found to be profitable even at natural gas prices of 12 $/MMbtu, as long as the product prices are over 100 $/bbl.

1. Introduction

The Fischer-Tropsch (FT) process is named after the two Germans F. Fischer and H.Tropsch. They were the first to industrialize this process in the years 1925-1935. Before that two Frenchmen, Sabatier and Senderens, had described the main reaction for the process (2).The Fischer-Tropsch synthesis is a gas-to-liquid (GTL) process, where natural gas is converted into liquid fuels such as diesel. The Fischer Tropsch synthesis is good way to utilize stranded gas[1] (3).

The fuels from the Fischer-Tropsch synthesis are extremely clean. The diesel made from FT-synthesis has lower sulfur content than the new Ultra Low Sulfur Diesel (ULSD) made from petroleum (4). The sulfur content is less than 1 ppm, there are less than 1% aromatics, the diesel has high cetane number (70-80) and the energy density is 7% higher than regular fuel (5)

There are several major GTL plants using the Fischer-Tropsch process. Among them are the ORYX plant in Qatar, which is owned by Qatar Petroleum and Sasol, which has a capacity of producing 34 000 bbl/day. Shell also has large GTL plants, the Shell Bintulu plant in Malaysia and the new PEARL project in Qatar (6). The PEARL project will operate at full production in 2012, and have a capacity of 140 000 bbl/day (7).

A typical GTL plant consists of three sections, first the syngas preparation, then the FT synthesis and last is the upgrading of the products (3). The syngas production is the most expensive part. There are several processes for preparing the syngas. Steam reforming (SR), autothermal reforming (ATR), partial oxidation (POX) or a combination of these (8). In the FT synthesis there are mainly two types of reactors being used; the slurry bubble column and the multitubular fixed bed reactor. The slurry bubble column reactor has been suggested because it is more isothermal, and is more appropriate for heat removal in the exothermic FT reaction (9).

There are two different FT synthesis, high and low temperature. The high temperature synthesis uses iron catalyst, operates at 330-350 °C and produces mainly light hydrocarbons (gasoline). The low temperature FT synthesis can use both cobalt and iron catalyst, operates at temperature about 200 °C and produces mainly heavier hydrocarbons (diesel and wax) (5).

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[1] The gas reserve is too remote from a natural gas market, and making pipelines is to expensive

The two main catalysts used in the Fischer-Tropsch are cobalt and iron. There are several differences in the use of these two catalysts. The iron catalyst is used for the high-temperature synthesis, but both can be used for the low temperature synthesis. The iron catalyst usually produces more olefins, and the cobalt catalyst is more resistant to attrition and is more suitable for slurry reactors. The cobalt catalyst operates at a narrow temperature interval, and a small temperature change can significantly change the methane selectivity (10). A comparison between them is given in Table 1.

Table 1 - Comparison between cobalt and iron catalyst (10)

Parameter	Cobalt	Iron
Cost	More expensive	Less expensive
Lifetime	Resistant to deactivation	Less resistant to deactivation
Activity at low conversion	Same as iron	Same as cobalt
Productivity at high conversion	Higher, less significant effect of water on the rate of CO conversion	Lower, strong negative effect of water on the rate of CO conversion
Maximum chain growth probability	0.94	0.95
WGS reaction	At high conversion	
Maximum sulfur content	< 0.1 ppm	< 0.2 ppm
Flexibility (temperature and pressure)	Less flexible	Flexible
H_2/CO ratio	~2	0.5-2.5
Attrition resistance	Good	Not very resistant

2. Process description

The main goal of the project was to simulate a plant that produces at least 17 000 bbl/day of products. A typical process flow diagram for a GTL plant is illustrated in Figure 1. In this figure the Air Separation Unit (ASU) and the upgrading unit is not shown.
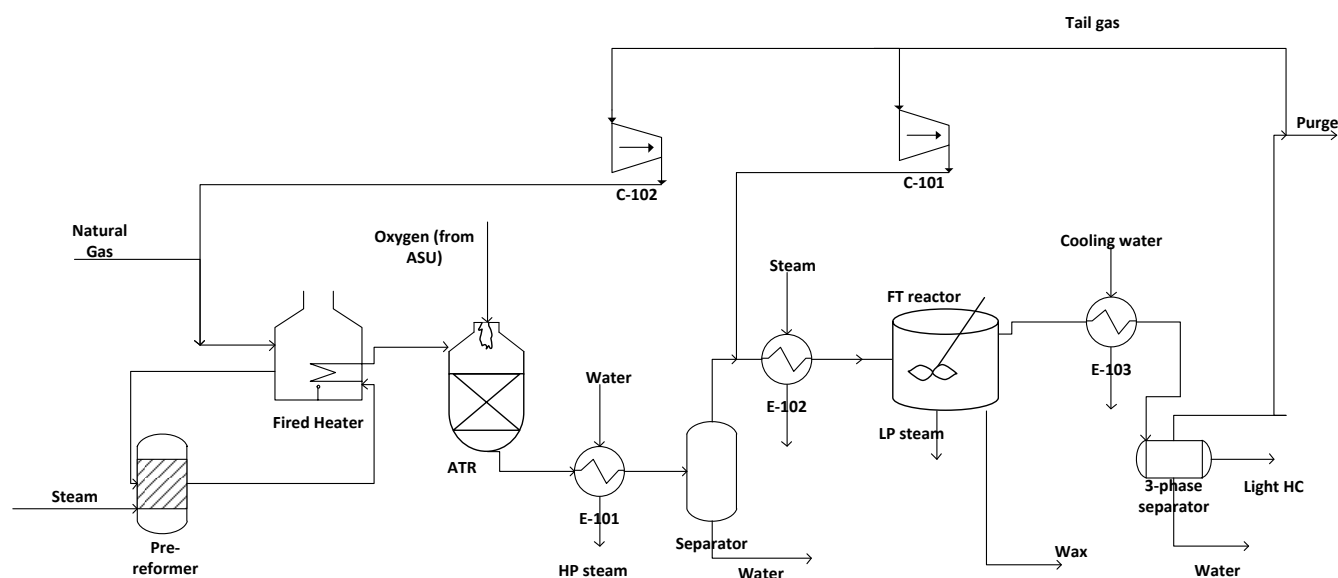
Figure 1 - Flowsheet

The inlet conditions used in the simulation is given in Table 2.

Table 2 - Inlet conditions

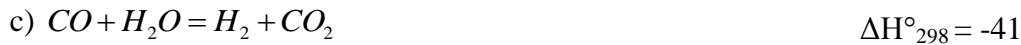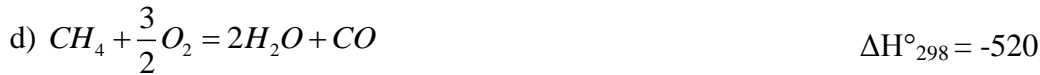|	Value
Temperature	50 °C
Pressure	50 bars
Flow rate	8218 kmole/h
Composition:	Mole fractions
CH_4	0.950
C_2H_6	0.020
C_3H_8	0.015
i-C_4 (iso-butane)	0.010
i-C_5 (iso-pentane)	0.005

2.1 Syngas Production

The natural gas enters a fired heater to reach the desired temperature before entering the adiabatic pre-reformer. This temperature is normally between 350-550 °C (11). Water steam is added to the pre-reformer along with the heated natural gas. In the pre-reformer following reactions takes place (3) (11). The enthalpies are given in [kJ/mole]

a) $C_nH_m + nH_2O = nCO + \left(n + \dfrac{m}{2}\right)H_2$ $\qquad$ $\Delta H°_{298} = 1175$ (for nC_7H_{16})

b) $3H_2 + CO = CH_4 + H_2O$ $\qquad$ $\Delta H°_{298} = 206$

c) $CO + H_2O = H_2 + CO_2$ $\qquad$ $\Delta H°_{298} = -41$

In the pre-reformer all higher hydrocarbons, except methane, are converted into CO and H_2 according to reaction a). The hydrogen and carbon monoxide are brought into equilibrium with methane and water according to reaction b). Reaction c) which is the water-gas-shift reaction also takes place in the pre-reformer.

After the pre-reformer the gas is heated before entering an oxygen blown ATR. The temperature of the inlet gas should be around 650 °C (11). The oxygen blown into the ATR should be around 200 °C (3). Inside the ATR there are three reactions happening (11).

d) $CH_4 + \dfrac{3}{2}O_2 = 2H_2O + CO$ $\qquad$ $\Delta H°_{298} = -520$

e) $CH_4 + H_2O = CO + 3H_2$ $\qquad$ $\Delta H°_{298} = 206$

c) $CO + H_2O = H_2 + CO_2$ $\qquad$ $\Delta H°_{298} = -41$

After the ATR, the gas is cooled down, and water is removed in a separator. Then the syngas is heated to 180-250 °C (12) before entering the FT reactor.

2.2 Fischer-Tropsch reactor

In the FT-reactor the syngas reacts to form higher hydrocarbons after reaction f)

f) $CO + 2H_2 = (-CH_2-) + H_2O$
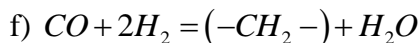
The catalyst used in this reactor is a cobalt catalyst, and the kinetic model used, is found in literature (13), and can be seen in equation 1.

$$r_{CO} = \frac{k_1 P_{H_2}^{0,6} P_{CO}^{0,65}}{1 + K_1 P_{CO}} \quad , \quad \left[\frac{kmole_{CO}}{m_{reactor}^3 s} \right] \tag{1}$$

Where the parameters k_1 and K_1 are given below

$$k_1 = 1.6 \cdot 10^{-5} \exp\left(-\frac{37326}{RT}\right) \quad , \quad \left[\frac{kmole_{CO}}{Pa^{1.25} m_{reactor}^3 s} \right] \tag{2}$$

$$K_1 = 1.096 \cdot 10^{-12} \exp\left(-\frac{-68401.5}{RT}\right) \quad , \quad \left[Pa^{-1} \right] \tag{3}$$

The kinetic model found in the literature was given in mole per second and per gram surface metal. To convert the expression to the wanted units, some assumptions have been made. The catalyst density is assumed to be 2000 kg/m^3, the weight fraction of cobalt in the catalyst is 20%, and 10% of the cobalt is exposed as surface atoms. The catalyst volume fraction is 10%. (14)

To calculate the product distribution, the products are assumed to follow the ideal ASF distribution[2]. The chain growth probability, α_{ASF}, was set constant to 0.9, even though it is actually an empirical correlation of partial pressures and temperature (15). The distribution can be seen in Figure 2 below.
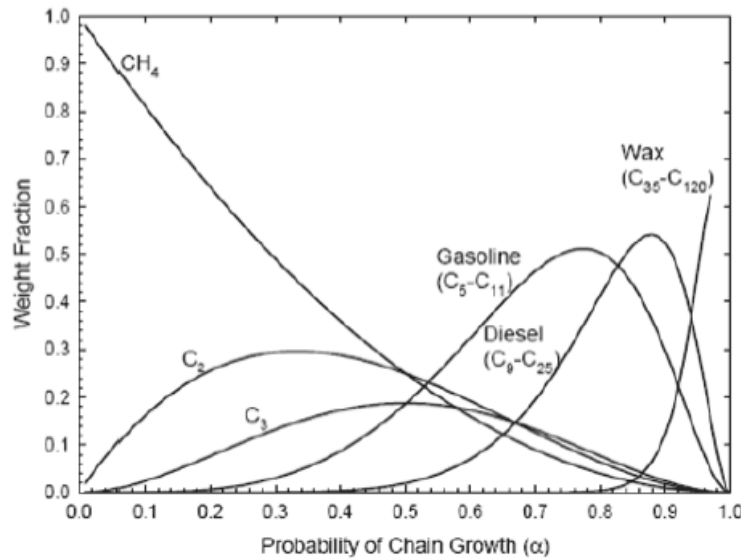
Figure 2 - ASF distribution

After the FT reactor the crude oil goes further to the upgrading part, where hydrocracking and isomerization takes place (16). The gas out of the three-phase separator recycles back to the pre-reformer and FT reactor. There is also a purge gas to avoid accumulation of CO_2. The purge gas is used as fuel in the fired heater.

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[2] Anderson-Schultz-Flory distribution

3. Assumptions, modeling and simulation

The simulation program used in this simulation is Aspen HYSYS® version 7.2.

The fluid package selected is the Peng-Robinson EOS[3], this is because this EOS should be applicable to calculate all fluid properties of natural gas processes (17).

3.1 Modeling in HYSYS

The fired heater seen on Figure 1 is modeled as two heaters, and the heat exchangers are modeled as heaters or coolers, this is done to simplify the simulations. After the simulation was finished a pinch analysis and heat integration was done for the entire process.

The pre-reformer was modeled as an equilibrium reactor. Reaction a) is actually a conversion reaction, but since the temperature inside the reactor is high, all higher hydrocarbons are completely converted, so it's ok to use an equilibrium reactor. The temperature in the pre-reformer was set to 455 °C, because that temperature is large enough to give an ethane conversion of 99,999 %. If the temperature was higher, the ethane conversion would have been slightly higher, but the fired heater and pre-reformer would also be more expensive.

The ATR is modeled as an equilibrium reactor. This assumption is made because the outlet temperature is so high (1030 °C) that equilibrium is a good assumption. The temperature in to the ATR is set to 650 °C. This temperature could have been lower, but then the ASU would have been more expensive. The temperature should not exceed 675 °C due to limitations on construction material (14), (18).

The pressure drops in heat exchangers are set to 35 kPa for liquids and 1 bar for gases (19).

The Fischer-Tropsch reactor is modeled as a CSTR reactor, and the reaction is heterogeneous catalytic. The overall FT reaction was modeled as one reaction, and written as

$$CO + UH_2 = p_1 C_1 + p_2 C_2 + .. + .. + p_{N-1} C_{N-1} + p_{[N,\infty]} C_{[N,\infty]} + H_2 O \qquad (4)$$
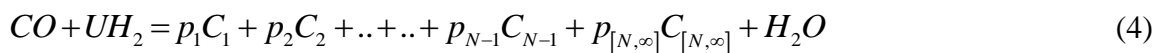
Since the database in HYSYS only contains hydrocarbons up to C_{30}, a remainder distribution was defined based on the average carbon number for hydrocarbons above C_{20} (20).

$$x_{[N,\infty]} = N + \frac{\alpha}{1-\alpha} = 30 \qquad (5)$$

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[3] Equation Of State

This means that $C_{[N,\infty]} = C_{30}$ represents $\sum_{N=21}^{\infty} C_N$ (6)

The U and p_i values are calculated in Appendix B.

The pressure drop in the reactor was neglected, because it is a slurry bubble column, and therefore the pressure drop will be low. The pressure drops in the heat exchangers are quite large, so it compensates for the neglected pressure drops in the reacors.

There is assumed only formation of parafins in the FT synthesis, olefin and alcohol formation is neglected.

3.2 Mass and energy balances

When using a simulation tool like HYSYS it is important to check the overall mass and energy balances to make sure they are conserved.

3.2.1 Total mass balance

The general mass equation at steady state is

$$\sum m^*_{In} = \sum m^*_{Out} \tag{7}$$

In this project that gives

$$\sum m^*_{Natural\,gas} + \sum m^*_{Steam\,in} + \sum m^*_{Oxygen} = \\ \sum m^*_{Stream\,7} + \sum m^*_{Wax} + \sum m^*_{Water\,out} + \sum m^*_{Light\,HC} + \sum m^*_{Purge} \tag{8}$$

The values for each stream are summarized in Table 3.

Table 3 - Total mass balance

Inlet streams	m* [kg/h]	Outlet Streams	m* [kg/h]
Natural gas	143368.0	Stream 7	111067.3
Steam in	112053.9	Wax	53966.6
Oxygen	159699.6	Water out	150989.1
		Light HC	60906.8
		Purge	38179.3
Sum In	415121.5	Sum Out	415109.1
Difference In-Out	12.4		
% wrong of inlet stream	0.003		

3.2.2 Total energy balance

The general energy equation for a steady state process is given below

$$\sum Q_{in} = \sum Q_{InletStreams} + \sum Q_{Added} = \sum Q_{OuletStreams} + \sum Q_{removed} = \sum Q_{out} \qquad (9)$$

The values gotten from HYSYS are given in Table 4 below.

Table 4 - Energy balance

Inlet streams	Q [kW]	Outlet Streams	Q [kW]
Natural gas	-174569.5	Stream 7	-489088.4
Steam in	-392010.8	Wax	-23033.4
Oxygen	7151.0	Water out	-660862.1
Duty 1	56153.5	Light HC	-46340.2
Duty 2	51978.5	Purge	-83633.1
Duty 4	93312.0	Duty 3	390099.5
Duty 6	1760.0	Duty FT	361833.7
Duty 7	154.3	Duty 5	194948.6
Sum In	-356071.0	Sum Out	-356075.4
Difference In-Out	-4.4		
% wrong of inlet stream	0.001		

The mass and energy balances are not completely conserved. This is due to the recycle loops in HYSYS which contains a small error.

4. Optimization

After the base case simulation was finished, an optimization had to be done in order to make the process as effective and economically cheap as possible.

4.1 Optimization of process variables

The adjusted variables were:

- Steam to carbon ratio
- Oxygen to carbon ratio
- Recycle ratio
- Purge ratio
- Volume of FT reactor

The steam to carbon ratio was adjusted to get the desired H_2/CO ratio of 2, according to reaction f (see chapter 2.2).

The oxygen to carbon ratio was adjusted to keep the temperature out of the ATR constant at 1030 °C

The optimal recycle ratio back to the FT reactor was found by comparing the recycle ratio vs. both the steam to carbon ratio and the production. The steam to carbon ratio should be as low as possible, because steam is expensive, but it is not favorable with a steam to carbon ratio below 0.4, because then soot formation can happen in the ATR (8). All other variables were held constant under these tests.
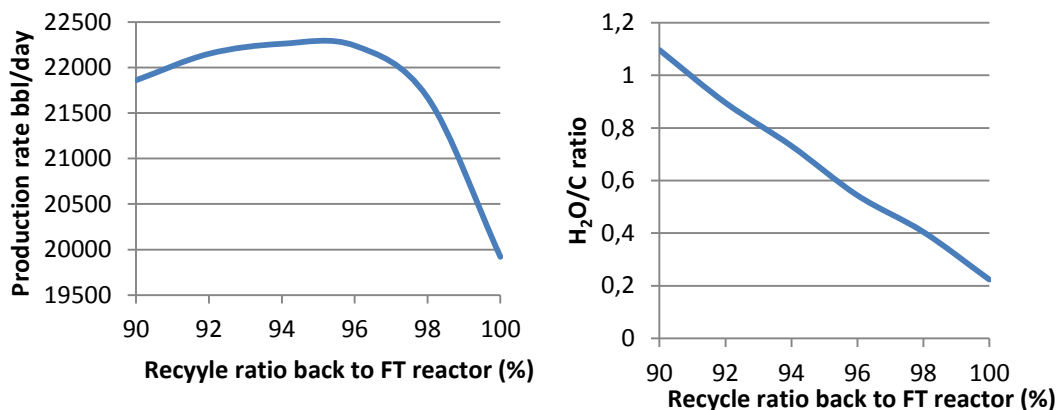
Figure 3 - a) Recycle ratio vs. production rate, b) Recycle ratio vs. steam/carbon ratio

From Figure 3 it can be seen that the optimal recycle ratio back to the Fischer-Tropsch reactor is 96%. This is a tradeoff between the production rate and the steam/carbon ratio.

The optimal purge ratio was found by comparing the purge ratio vs. the carbon efficiency and the steam to carbon ratio, whilst holding all other variables constant.

Figure 4 - a) Purge vs. carbon efficiency   b) steam/carbon ratio

The optimal purge ratio is a tradeoff between the carbon efficiency and the steam consumption seen in Figure 4, and is found to be 3.5%

The optimal reactor volume was found by comparing the reactor volume against the production rate, whilst holding all other variables constant.
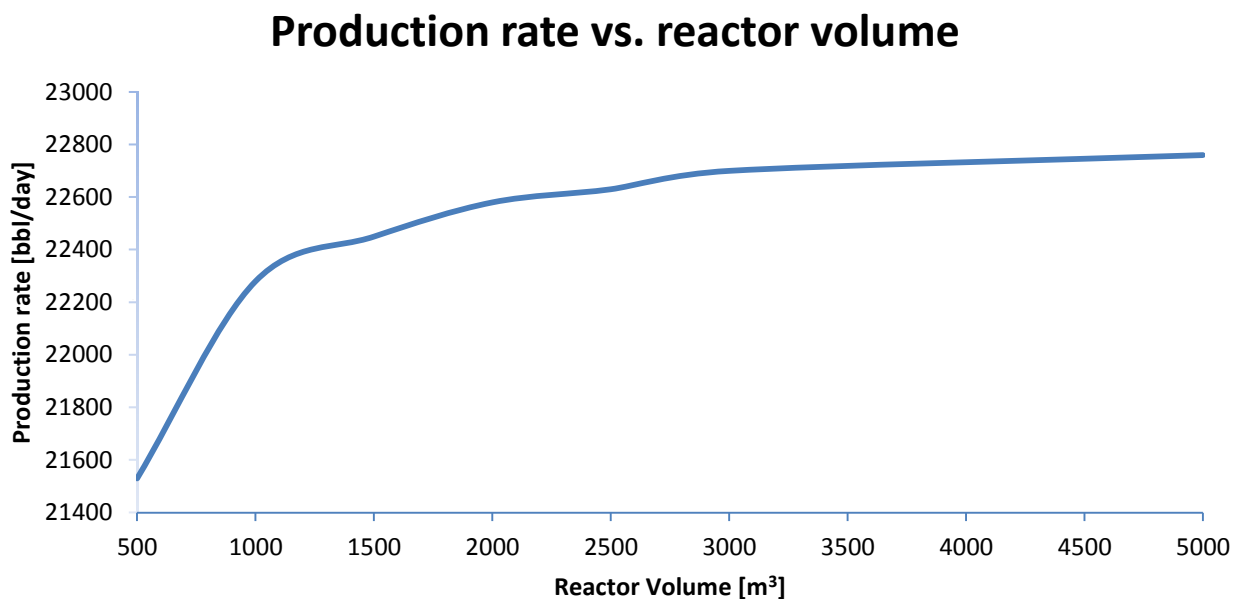
Figure 5 - Reactor volume vs. production rate

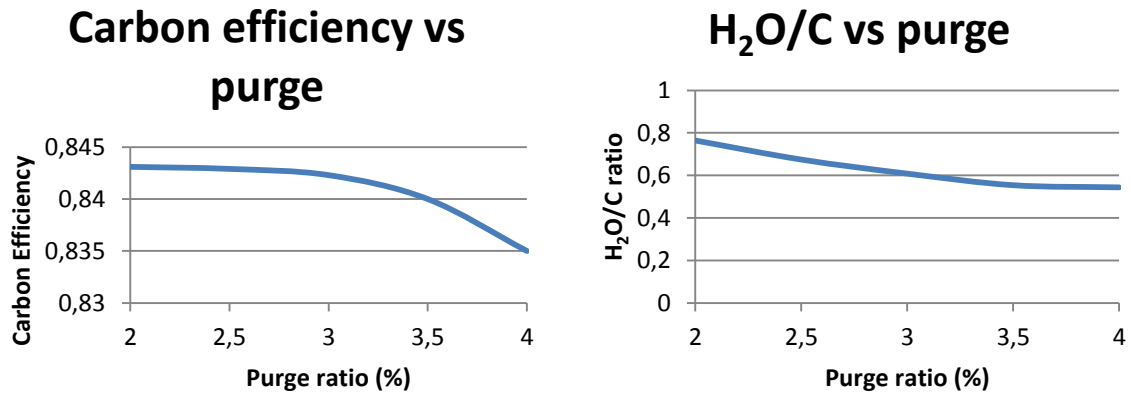The optimal reactor volume will be a tradeoff between the production and the cost of the reactor. The slope seems to flatten at about 2700 m^3 so that is the volume used in this project.

More information about the optimization can be found in Appendix C.

4.2 Heat integration

The heating and cooling demands are huge in this plant, and therefore it's important to look for several possibilities of heat integration to make sure as little heat as possible gets wasted.

4.2.1 First heat Exchanger (E-101)

The hot syngas coming out of the ATR is 1030 °C, and it was decided to make benefit of that to produce high pressure steam at 510 °C and 110 bars. A pinch analysis was done as showed in Figure 6

Figure 6 - Pinch analysis

The minimum temperature approach was set to 10 °C, and the amount of high pressure (HP) steam that can be produced is 361.3 ton/hour. There is no hot work needed in this heat exchanger, but there is cold work needed, which is covered by using cooling water. Four heat exchangers and a steam drum are needed f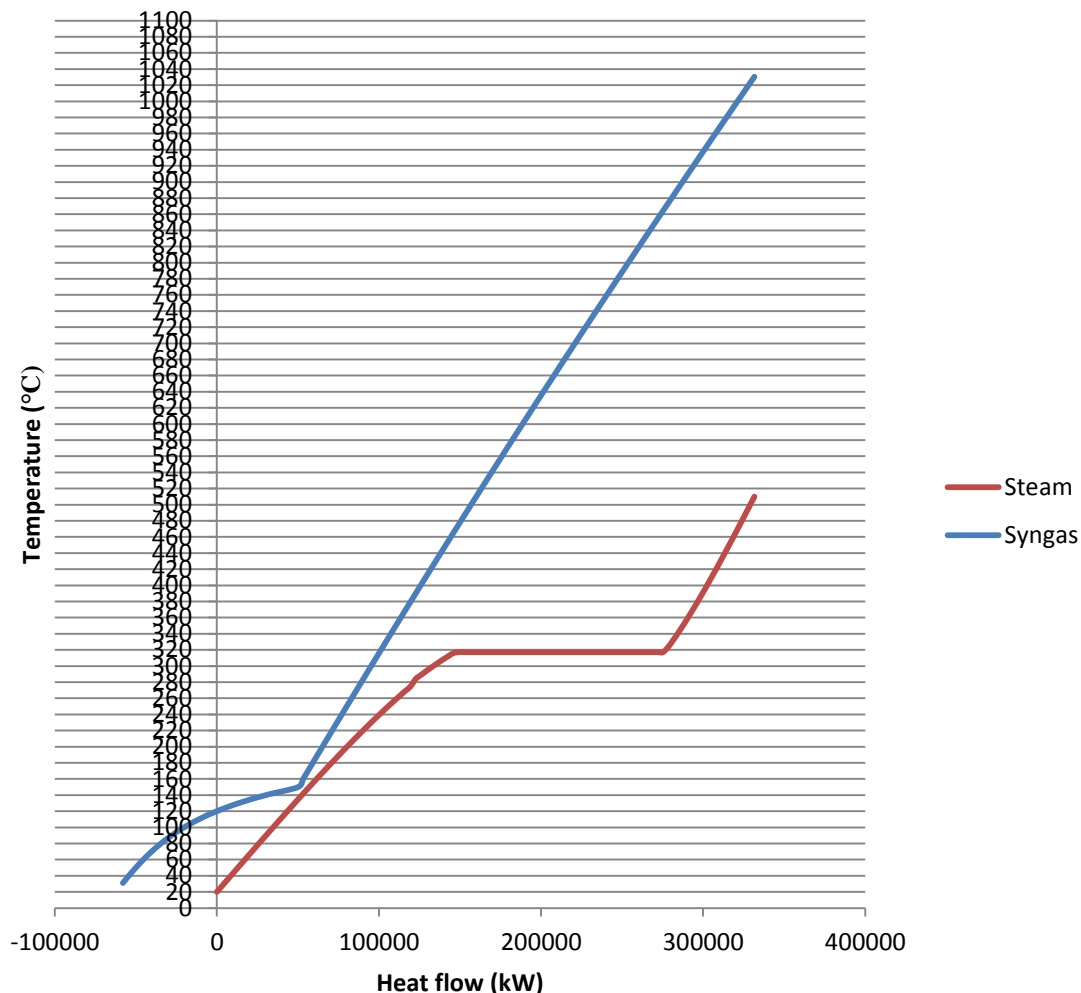or this combination (see Appendix D), one for heating the water, one for evaporating the water, one for superheating the water steam and one to cool the hydrocarbon fluid to the wanted temperature.

The syngas leaving the ATR at 1030 °C is so hot, that it was decided not to pair this stream against another stream because of possible material problems.

4.2.2 Reaction heat

The exothermic reaction in the FT reactor produces a lot of energy, and it was decided to produce low pressure (LP) steam of that energy. The LP steam is at 10 bars and 190 °C. The amount of steam that can be produced is 470.3 ton/hour. Inside the reactor there are cooling tubes connected to a steam drum for transferring heat and producing steam.

4.2.3 Last two heat exchangers (E-102 and E-103)

Several opportunities were tried for these two exchangers (see Appendix D).
The best solution was found when using steam and cooling water instead of matching them against each other. Two hydrocarbon fluids in a heat exchanger have bad heat transfer so the exchanger would have been extremely huge in that case.

For E-102 the best solution was found to be a series of two exchangers, where LP steam is used the first, and the extra hot work needed is covered by HP steam in a second exchanger. The LP steam used is 118 ton/hour, and the amount of HP steam used is 8 ton/hour.

For E-103 cooling water was used at an amount of 16 000 m^3/hour, entering at 8.5 °C, and leaving at 18.5 °C (21)

4.3 Process Summary

A summary of the process parameters, after the optimization and the heat integration, are given in Table 5 below

Table 5 - Process parameters

Variable	Value	Unit
Oxygen consumption	$1.597*10^5$	kg/h
Natural gas consumption	$1.434*10^5$	kg/h
Net HP-steam production	241 341	kg/h
Net LP-steam production	352 600	kg/h
LPG production	805.4	bbl/day
Gasoline production	6294	bbl/day
Diesel production	6687	bbl/day
Wax production	7844	bbl/day
Process water used	831 600	kg/h
Purge ratio	3.5	%
Recycle ratio back to FT reactor	96	%
Carbon Efficiency	0.8467	
CO conversion in FT reactor	0.8941	
H_2/CO ratio into FT reactor	1.997	
Steam/C ratio into ATR	0.5193	
Oxygen/C ratio into ATR	0.4837	
CO_2 fraction in tail gas	0.6528	
Reactor volume	2700	m^3

5. Economics

To calculate the prices of the different equipment some assumptions have been made. The density of stainless steel is assumed to be 8000 kg/m^3, all other factors and economic assumptions are gotten from literature (1). The prices of utilities are given in Table 6.

Table 6 - Utilities

Utility/Product	Price	Units
LPG	104.10	$/bbl
Gasoline	121.05	$/bbl
Diesel	126.27	$/bbl
Wax	100.41	$/bbl
HP steam	29.97	$/1000 kg
LP steam	28.31	$/1000 kg
Electricity	0.06	$/kWh
Distilled process water	0.067	$/1000 kg
Natural gas	0.5	$/MMbtu
Waste disposal	41	$/1000m^3

The product prices are found from OPEC (22), and the rest of the utilities are found in literature (23)

5.1 Fixed capital investment

To estimate the cost of the equipment, a correlation found in literature was used (1). The correlation is given in equation 10 under.

$$C_e = a + bS^n \tag{10}$$

Where a and b are cost constants, n is an exponent for that equipment and S is the size parameter.

The installation cost was found using the formula found in literature

$$C = \sum_{j=1}^{n} C_{e,j,CS} \left[(1+f_p)f_m + f_{er} + f_{el} + f_i + f_c + f_s + f_l \right] \tag{11}$$

The different factors are given in literature (1). The total equipment and installation cost are given in Table 7 under.

Table 7 - Equipment and installation cost (2011 price)

Equipment + Installation cost, 2011	$	%
Pump 1	108 175	0,03 %
Pump 2	127 109	0,04 %
Compressor 1	10 716 968	2,97 %
Compressor 2	4 526 185	1,25 %
FT reactor	26 726 266	7,40 %
Pre-reformer	2 446 200	0,68 %
ATR	10 246 411	2,84 %
Water Separator	1 288 270	0,36 %
3-phase separator	1 874 278	0,52 %
Fired heater	22 013 481	6,09 %
Heat Exchanger 1	221 400	0,06 %
Heat Exchanger 2	477 039	0,13 %
Heat Exchanger 3	4 786 404	1,32 %
Heat Exchanger 4	1 484 715	0,41 %
Heat Exchanger 5	2 782 782	0,77 %
Heat Exchanger 6	162 443	0,04 %
Cooling tubes FT reactor	1 578 507	0,44 %
Heat Exchanger 7	3 964 723	1,10 %
Steam Drum 1	3 220 949	0,89 %
Steam Drum 2	435 790	0,12 %
ASU	178 785 610	49,48 %
Catalyst	36 383 940	10,07 %
Upgrading Unit (13 %)	46 972 981	13,00 %
Total Equipment + installation cost	361 330 625	100,00 %
Engineering (10% of eq.+inst. Cost)	36 133 062	
Contingency (10% of eq.+inst. Cost)	36 133 062	
Fixed capital investment	433 596 750	

The cost of the upgrading unit was set to 13% of the total equipment and installation cost (24). More info can be found in Appendix E.

5.2 Working capital

The working capital is additional money needed in order to run the plant until the plant starts to earn income. The working capital is returned at the end of the project time. For petrochemical industries the working capital is typically 15 % of the equipment and installation cost (1).

5.3 Production costs

The production costs are the costs which are dependent on production, such as electricity and operators.

The number of operators needed is estimated from literature (23).

$$N_{OL} = \left(6.29 + 31.7P^2 + 0.23N_{np}\right)^{0.5}$$

(12)

Where P is the number of processing steps involving handling of particulate solids, and N_{np} is the sum of equipment (including compressors, towers, reactors, heaters and heat exchangers). N_{OL} is the number of operators needed per shift, and there are 4.5 shift positions needed (23). In total there are 15 operators needed.

A summary of the production cost can be found in Table 8. All the percentages are found in literature (1)

Table 8 - Operating Cost

		$/year
Electricity	Power consumption from HYSYS used	63 008 208
Natural gas	Flow from HYSYS used, 0.5 $/MMbtu	26 945 608
Operators	15 operators, 60 000 $/year	900 000
Supervision	25% of operating labour	225 000
Direct salary overhead	50% of operating labour + supervision	562 500
Maintenance	4% of equipment + installation cost	14 453 225
Property taxes and insurance	1.5 % of equipment + installation cost	5 419 959
Rent of land	1.5% of equipment + installation cost	5 419 959
General plant overhead	65% of operating labour + supervision + maintenance + direct overhead	10 491 471
Allocated environmental charges	1% of equipment + installation cost	3 613 306
Running license fees and royalty payments	0.5% of equipment + installation cost	1 806 653
waste disposal	41 $/1000 m3	86 297
Process water	0.067 $/1000 kg	445 738
Total Operating cost		133 377 924

5.4 Income

The income is from the products and steam sold. A summary is given in Table 9. The data used in calculations are given in Table 5 and Table 6.

Table 9 - Annually income

Income		
LPG	27 947 380	$/year
Gasoline	253 962 900	$/year
Diesel	281 455 830	$/year
Wax	262 538 680	$/year
HP steam	57 863 918	$/year
LP steam	79 856 848	$/year
Total income	963 625 556	$/year

5.5 Profitability

The profitability is seen from the net present value (NPV) and the internal rate of return. The net present value is found from equation 13 where CF is cash flow in year j, and i is the interest rate. The internal rate of return (IRR) is the rate that gives NPV=0.

$$NPV = \sum_{j=0}^{lifetime} \frac{(CF)_j}{(1+i)^j} \tag{13}$$

The interest rate was set to 9%, and the lifetime was set to 20 years. Taxes were set to 30%, and a reducing balance depreciation of 20 % was used. Figure 7 shows the NPV for the cases with 0.5 and 5 $/MMbtu for natural gas.
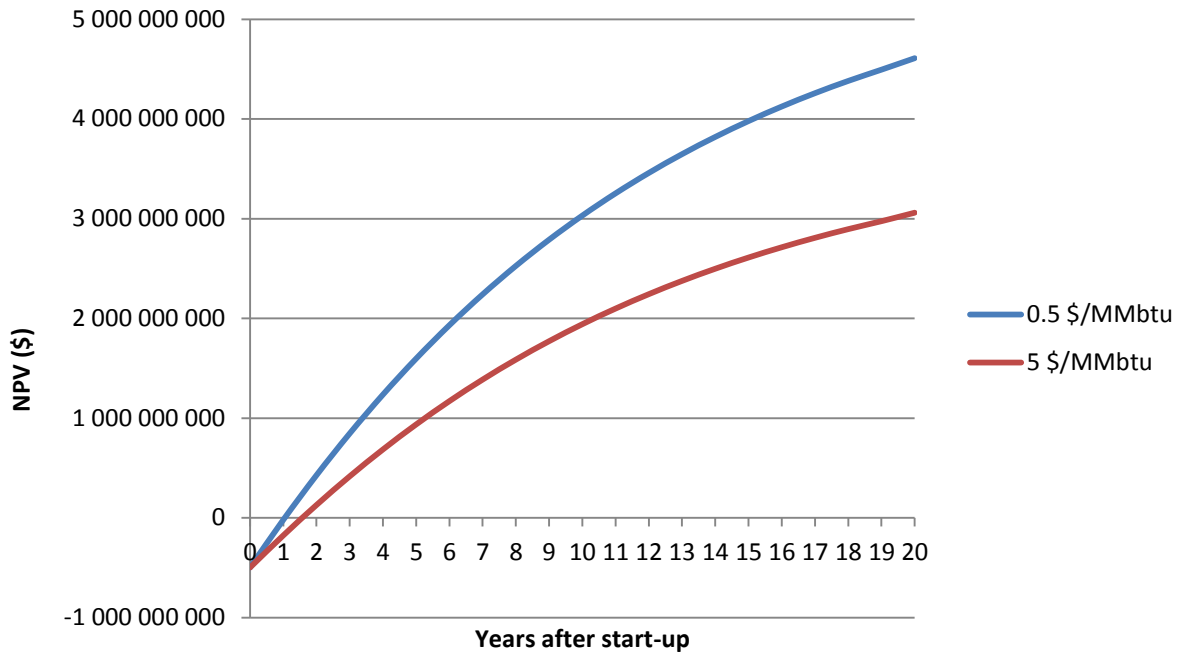
Figure 7 - NPV for different natural gas prices

As it can be seen in Figure 7 the plant is highly profitable at these natural gas prices, when the prices for steam and products are as given in Table 6. For a natural gas price at 0.5 $/MMbtu the internal rate of return is 106 %, and the net present value is 4.6 billion dollars. For the case where the natural gas price is 5 $/MMbtu the internal rate of return is 73 %, and the net present value is 3.1 billion dollars. Both of the cases have a payback time between one and two years.
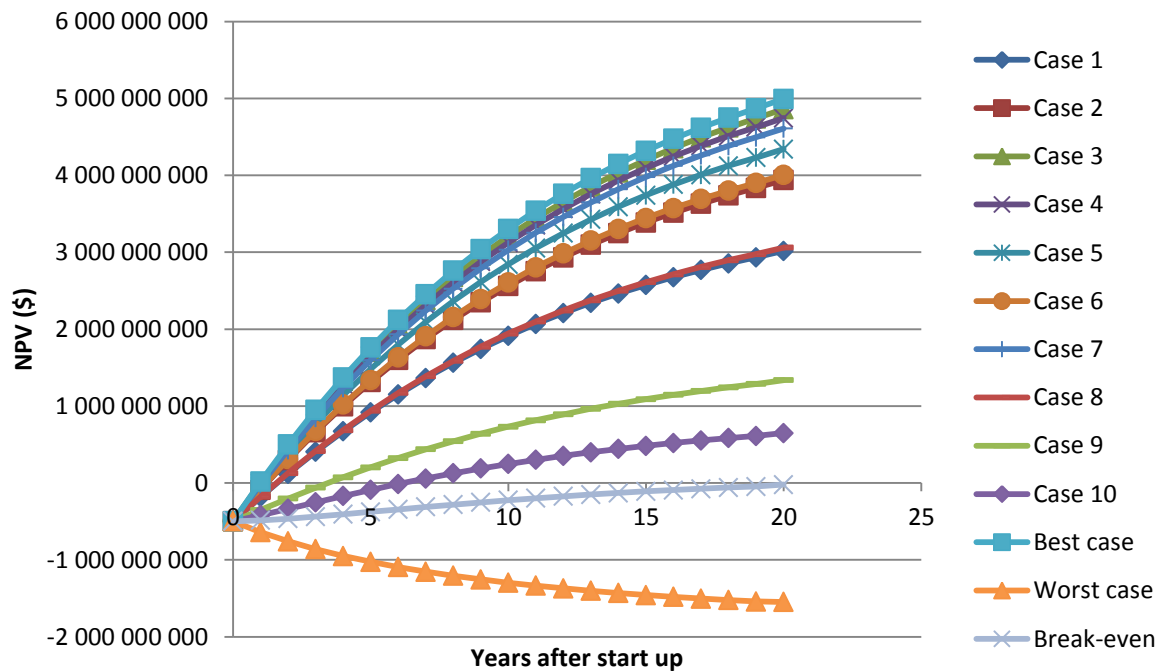
6. Sensitivity analysis

In the sensitivity analysis it is interesting to see how the profitability changes with product price, natural gas price and electricity price. In Table 10 and Figure 8 below a summary of different cases are compared. If nothing special is noted, then the values in Table 6 are used.

Table 10 - Adjusted variables in sensitivity analysis

Case	Adjusted variable	NPV	IRR
Case 1	Average product price 80 $/bbl	3 017 637 279	72 %
Case 2	Average product price 100 $/bbl	3 939 089 719	92 %
Case 3	Average product price 120 $/bbl	4 860 542 159	112 %
Case 4	0.04 $/kWh el. price	4 743 551 340	109 %
Case 5	0.10 $/kWh el. price	4 340 930 027	101 %
Case 6	0.15 $/kWh el. price	4 005 412 266	93 %
Case 7	0.5 $/Mmbtu natural gas price	4 609 344 236	106 %
Case 8	5 $/Mmbtu natural gas price	3 059 706 716	73 %
Case 9	10 $/Mmbtu natural gas price	1 337 887 251	37 %
Case 10	12 $/Mmbtu natural gas price	649 159 464	23 %
Best case	0.04 $/kwh, 0.5 $/Mmbtu, 120 $/bbl	4 994 749 263	115 %
Worst case	0.15 $/kwh, 12 $/Mmbtu, 80$/bbl	-1 546 479 462	
Break even	12 $/Mmbtu, 100 $/bbl	~0	9%

Figure 8 - Sensitivity analysis

As it can be seen from Figure 8, the profitability is very dependent on the natural gas price, and also the product price is of high importance. The electricity price doesn't have the same impact on the results. If the electricity and natural gas prices are high, meanwhile the product prices are low, then the project is not profitable as seen from the worst case.

The break-even point is when the NPV is zero, and that happens if the average product price is 100 $/bbl, and the natural gas price is 12 $/MMbtu. At product prices lower and natural gas prices higher than this, the plant will no longer be profitable.

7. Discussion

7.1 Economic discussion

The economic discussion is divided into three parts to give a better view.

7.1.1 Fixed capital investment

The fixed capital investment found in this analysis is a bit low when comparing to actual plants. Here it was found to be 434 million dollars for a 21 600 bbl/day plant. This is a bit low when comparing to the ORYX plant, which has a capacity of 34 000 bbl/day, and investment cost of one billion dollar (14).

There is a large uncertainty in the equipment and installation costs, using the methods found in literature (1). All reactors are modeled as simple pressure vessels, but in reality they are much more complex. The ATR for instance also contains a burner and a refractory lining to protect the pressure vessel from hot gases (25), which is not included in this analysis. The thickness of the vessels is also an uncertain factor, because the temperatures are so high that the stress-temperature relationship used is not valid, and thus the vessels may have been estimated to thin, which again makes the fixed capital investment too low.

The ASU is the most expensive part of the plant, and the largest uncertainties are associated with it. The price used is from a smaller plant, and it is scaled up to the current oxygen production

7.1.2 Operating costs

The operating costs are highly dependent on the natural gas prices. At natural gas prices of 0.5 $/MMbtu the project is very profitable. To get natural gas prices that low, the plant must be in an area were the gas is stranded. According to Platts (26) the price for natural gas is about 4.3 $/MMbtu at the Gulf coast, and ranging from 9-12 $/MMbtu in central Europe (27). A natural gas price at 0.5 $/MMbtu is very low, and not likely unless the plant is in a very remote area.

If the plant is in a very remote area other problems such as higher electricity prices must be considered. The fixed capital investments will probably also be higher due to higher transport costs, and it's likely that the salaries to the staff will be higher because the plant is so remote that high salaries is the only way to get people to work there. On the positive side the land needed to rent (or buy) will probably be cheaper due to the remote location.

The number of operators needed to run the plant in this analysis is probably estimated too low, because the ASU and the upgrading unit is considered one unit, even though both of them consists of several units.

Another factor that should be considered is the financing of the plant. If the company building the plant needs to loan money from a bank, the rents should be taken into account in the economic analysis, which they are not here.

7.1.3 Income

The steam prices used in the economic analysis are the prices when steam is bought and not the selling price, this will lead to a higher income due to steam sale.

In addition the prices of products are very high at the moment compared to the last few years (22). This is a factor that makes the plant very profitable compared to earlier reports (14). If the selling price dropped to about 80 $/bbl of product, the plant will still be profitable at low natural gas price as seen in Figure 8. The break-even is when the natural gas costs 12 $/MMbtu, and the products are sold at 100 $/bbl.

7.2 Equipment selection

The steam/carbon ratio into the ATR is fairly low (0.52), thus it is important to study further on the ATR, to avoid side reaction which can make soot formation appear after equation 14. Especially the burner design should be studied further (3).

$$C_n H_m = nC + \frac{m}{2} H_2 \qquad (14)$$

Soot formation can also be a problem in the pre-reformer, where carbon whisker tends to grow with a nickel crystal at the top (3). Soot formation leads to catalyst deactivation and is favorised at high temperatures and low steam/carbon ratio.

The temperature in the ATR and out of the ATR is so high that other material than stainless steel should be considered. This is to avoid metal dusting, which is a severe form of corrosion that occurs when a material is in an environment with high carbon activity. The corrosion can break up bulk metal to metal powder which interferes the process. A mechanism to avoid metal dusting can be to protect the metal with a chromium layer (28).

7.3 Reactor configuration

The reactor is modeled as a CSTR in HYSYS, even though it is a slurry bubble column. This is not an optimal model, and better results could have been gained by using an axial dispersion model (ADM).

The reactor is also modeled as one large reactor (2700 m^3). A reactor that huge can be difficult to operate and install, thus it could have been better to have several smaller reactors in parallel. The design of the reactor should also be studied more because handling of slurry can promote plugging in the reactor (29). The volume of the reactor might not be optimal, because it's compared only against the production rate. A more optimal volume could have been found if a profitability analysis had been done for different reactor volumes.

8. Conclusion

This investigation showed that the optimal operation of the FT plant is at a purge ratio of 3.5 %, and a recycle ratio back to the FT reactor of 96 %. The optimal reactor volume is found to be 2700 m^3.

After a pinch analysis was done, the plant uses excess heat to produce low and high pressure steam. The net amount of low pressure steam that is produced and sold is 353 tons/hour, and the net amount of high pressure steam that is produced and sold is 241 tons/hour.

The plant is highly economical at today's (2011) product prices, if the natural gas price is 0.5 $/MMbtu. With a life time of 20 years and an interest rate of 9 %, the net present value becomes 4.6 billion dollars, and the internal rate of return is 106 %.

The break-even point is when the natural gas costs 12 $/MMbtu and the average product prices are 100 $/bbl.

Trondheim 09.12.2011   Ole Kristian Nordvåg

List of Symbols

Symbol	Units	Description
a		Cost constant
A	m^2	Area
b		Cost constant
Ce	$	Cost
CF	$	Cash flow
C_p	J/(kg*°C)	Specific heat capacity
D	m	Diameter
f_i		Lang's factors
H	kJ/mole	Enthalpy
i		Interest rate
IRR		Internal rate of return
j		Years after start up
k_1	$kmole_{CO}/(Pa^{1.25}*m^3_{reactor}*s)$	Rate constant
K_1	Pa^{-1}	Rate constant
L	m	Length of tubes
m*	kg/h	Mass flow
n		Cost exponent
N_{np}		Sum of equipments
N_{OL}	operators	Number of operators
NPV	$	Net present value
p_i		Parameter in ASF distribution
P_i	Pa	Design pressure
P		Number of processing steps involving handling of solids
Q	kW	Heat flow
r	$kmole_{CO}/(s*m^3_{reactor})$	Reaction rate
S		Size parameter
t	m	Thickness
T	°C	Temperature
U	W/m^2	Total heat transfer coefficient
U'		Parameter in ASF distribution

V	m^3	Volume
α		Chain growth probability
ρ	kg/m^3	Density
σ	ksi	stress

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