LTO, llc

Natural Gas Production in Peru

Group 1:

Laura Causey
Michael Eatmon
Sara Habib
Kien Nguyen
Sheau Te

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Description of Project

This project's main goal is to determine the best investment plan for the natural gas reservoir located in the western part of Peru. The Camisea reservoir contains 11 trillion cubic feet of proven natural gas reserves. This reserve is large enough that no limits exist on the amount of natural gas that the plant will be able to process from the Camisea plant within the lifetime of the project. The proposed project takes into account the different investment options for a thirty-year period. Forecasting of product prices and demands are taken into account in order to reach a reliable investment option. A planning model solved using mathematical programming was used.

Results

The investment planning model for the Camisea natural gas reservoir in Peru, suggests that an initial investment of \$7 billion or \$9 billion maximum initial investment should be invested in the plant depending on the amount the company is willing to initially invest. The amount spent should be \$4.5 billion on the LNG and NG pipelines that connect the Camisea field with the plant location in Pisco, Peru. This will result in a net present value of \$52.5 billion for the initial investment of \$7 billion and a net present worth of \$55.0 billion for an initial investment of \$9 billion over the life span of the plant which is assumed to be 30 years. The model suggests that initially, liquefied natural gas should be sold to California, and ethylene glycol, ammonium nitrate, methanol, diesel, and polypropylene process should be built. In the second year, the vinyl chloride and polyvinyl chloride processes are constructed.

Future Recommended Studies

In order to find the best investment for the Camisea natural gas reserve, further investment options should be added to the overall model. The price and demand forecasting for each project can be updated. Export taxes and government regulations for each country should also be accounted for. Also the scope of the export areas could be examined. More detailed economic analysis would ensure the accuracy of these initial results.

1.0 Introduction

1.1 Camisea Natural Gas Project Background

In 1980, Shell Oil discovered a natural gas reservoir in the Amazon jungle. The Camisea natural gas reservoir located in eastern Peru contains 11 trillion cubic feet of proven natural gas reserves, which is equivalent to 600 million barrels. This reservoir is considered to be the largest in South America and is located in the Urubamba River Basin. A map of Peru is depicted in Figure 1.1.1. Table 1.1.1 shows the chemical composition of the natural gas found in the Camisea reserves.

Figure 1.1.1 Peru Map

Table 1.1.1 Natural Gas Chemical Composition

Camisea Reservoir Natural Gas Chemical Composition (mass%)	
Methane	0.883509
Ethane	0.104537
Propane	0.002009
iButane	0.000033
nButane	0.000041
iPentane	0.000003
nPentane	0.000002
C6	0.00002
N2	0.007275
CO2	0.002572
Total	1.00000

Currently, the natural gas is extracted from the Camisea reservoir at a rate of 400 million cubic feet per day. The natural gas from the reservoir is transported via two natural gas pipelines. One pipeline transports liquid natural gas extracted from the reservoir. This liquid extracted from the natural gas contains ethane, propane and higher hydrocarbons. The second pipeline transports the natural gas extracted from the reservoir, which has been processed though an initial natural gas distillation column. The liquid natural gas pipeline travels from Camisea to Pisco, the western coast of Peru, to the city of Pisco. At Pisco, the natural gas will be processed by the different processes, or will be converted to liquefied natural gas and exported. The natural gas pipeline will go from the Camisea reservoir to Pisco for processing, and then further continue on to Lima for consumption. Figure 1.1.2 shows the path of the current natural gas pipelines.

Figure 1.1.2 Path of Natural Gas Pipelines

1.2 Project Objectives

This project examines the investment and business planning for natural gas produced from the Camisea reservoir. The objective of this project is to evaluate the different possible products that can be produced by natural gas. The fixed capital investment and operating cost for each of the potential processes were determined as a function of capacity. Market analysis, demand, and price forecasting was conducted for each of the products. An investment planning model was developed and programmed in GAMS/CPLEX and a business plan is presented comparing different plans based on initial capital investment.

1.3 Project Goals

In order to complete the project in the allotted time, initial goals were set. The Peruvian petrochemical market was researched. Import and export markets for products derived by natural gas were investigated. The main focus of the first phase of the project was to generate a working GAMS business model.

Fifteen different processes were considered as potential investments. The selection of each process was determined by ranking the products by value and demand. The fifteen processes chosen were: natural gas distillation, ethylene synthesis, vinyl chloride process, ethylene glycol process, polyethylene (low, high, and linear-low density) production, polypropylene production, thermal coupling-oligomerization, ammonia synthesis, nitric acid formation-fertilizer, methanol synthesis, Fischer-Tropsch synthesis, and liquid natural gas process.

Other processes were examined and the products produced were not in high demand thus they were not considered as model inputs. With an unlimited surplus of time all options would have been explored and a perfect model would be compiled. Natural gas is used in producing non-ferrous metals, since these are not in high-demand in the region this process was eliminated as a potential process for the GAMS model. Natural gas is used to produce non-metallic minerals such as bricks, tiles, glass and cement. Currently in the region, other energy is used to produce these products and breaking into this market seems unrealistic, thus it was not included as a potential process for the GAMS model. Other processes that were examined but did not qualify for this project will be further discussed in the final phase of this project.

2.0 Project

2.1 Natural Gas Process Diagram

Many different processes were considered as potential routes for the natural gas from the Camisea reservoir. Figure 2.1.1 shows the process flow diagram including all the processes considered for this project.

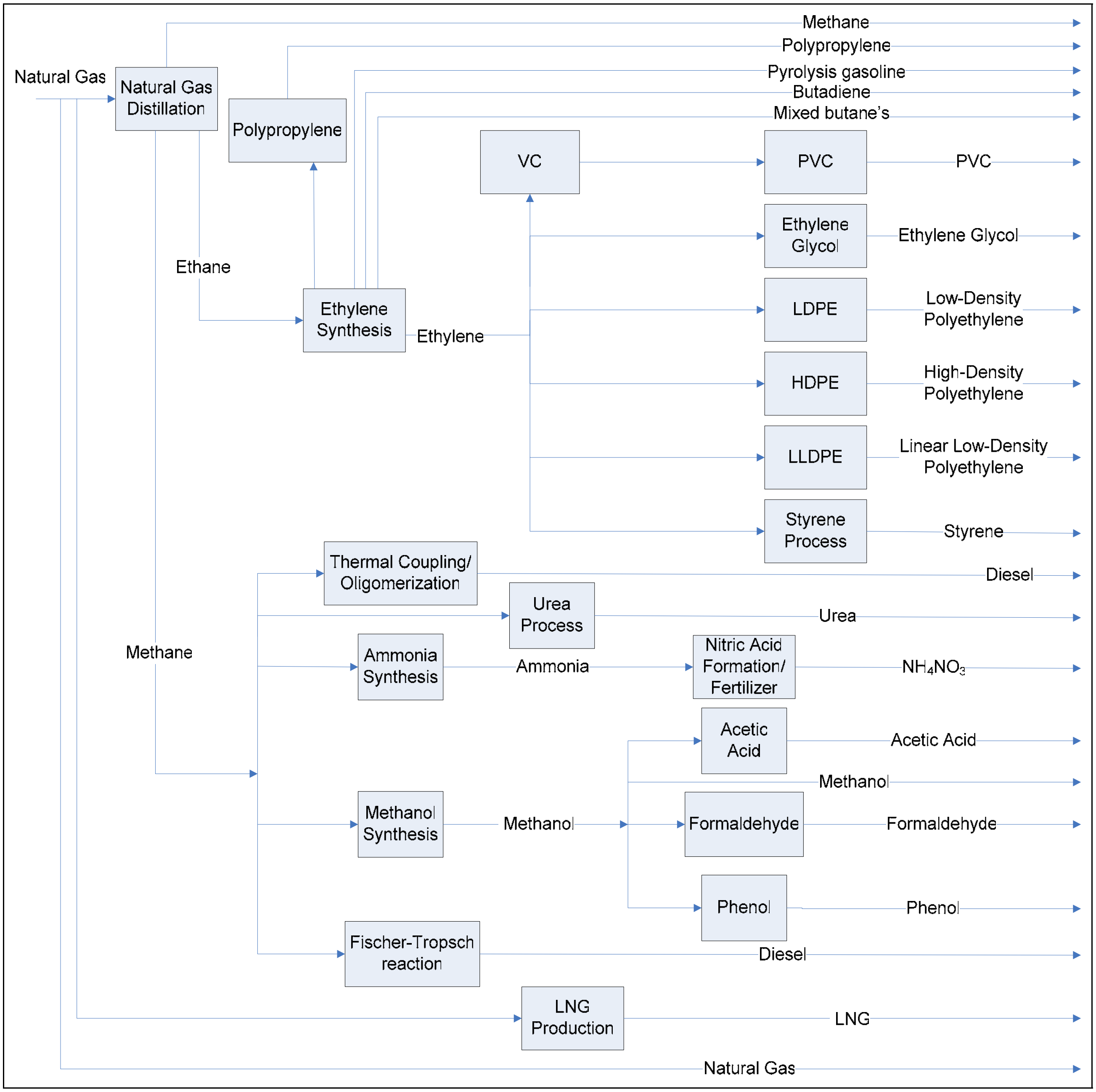


Figure 2.1.1 Process Flow Diagram of All Potential Processes

2.2 Process Selection

In order to refocus the scope of the project, and initial check was made for each process. Processes that were deemed to initially be not profitable were cut from being considered for detailed economic calculations. Five processes were initially eliminated from the overall project. The first process eliminated was the plant producing acetic acid. The supply and demand for this product was not high enough in the Peru and surrounding markets to make this process profitable. Along with this process, the process forming formaldehyde was eliminated because the current market for that product is dominated by Brazil. Since the current demand for formaldehyde is satisfied, this process was not considered.

The process producing urea was also not considered for economic analysis. This is because the international demand for urea is decreasing and many other more economical products can be made from the natural gas. The processes forming phenol and styrene were also removed from the overall process flow diagram because both processes produce products that are a part of a saturated market. For phenol, the current demand is satisfied and with increasing prices and low margins, this process is not profitable. Styrene was not considered because local companies have satisfied the market for styrene, and the increased cost of transportation for exportation would make this process even less profitable.

Individual gases such as ethane and pentane were not sold since they could be used in other processes to yield higher profit products. The ethane is used to produce ethylene from the ethylene synthesis plant. The ethylene is then used to produce higher profit products such as polyvinyl chloride, ethylene glycol, low-density polyethylene, high-density polyethylene, and linear low-density polyethylene. The products consistently have a higher profit yield than the sale of pure ethane.

The final process eliminated from the overall process diagram was thermal coupling and oligomerization. Thermal coupling is used to convert a monomer into an oligomer. These higher carbon chains are used to produce jet fuel and premium gasoline. This process was initially considered to be a good venture, and initial economic analysis was completed. Due to limited literature over the process, this process was eliminated from the project. Currently only laboratory processes data is available and no current large scale information of the process is obtainable. Since a selective group of companies are looking to patent this process on an industrial scale, information considering the patent is confidential. It is recommended that upon obtaining large scale plant information that this process be added to the overall mathematical model design.

A complete process flow diagram for every potential plant considered was formed. The material balances for each process were performed and entered into the overall process-to-process flow diagram. Figure 2.2.1 shows the process-to-process flow diagram containing all possible interconnections. The planning model should choose which process to build, when to build the process, at what initial capacity to operate the process, and when to expand the plant.

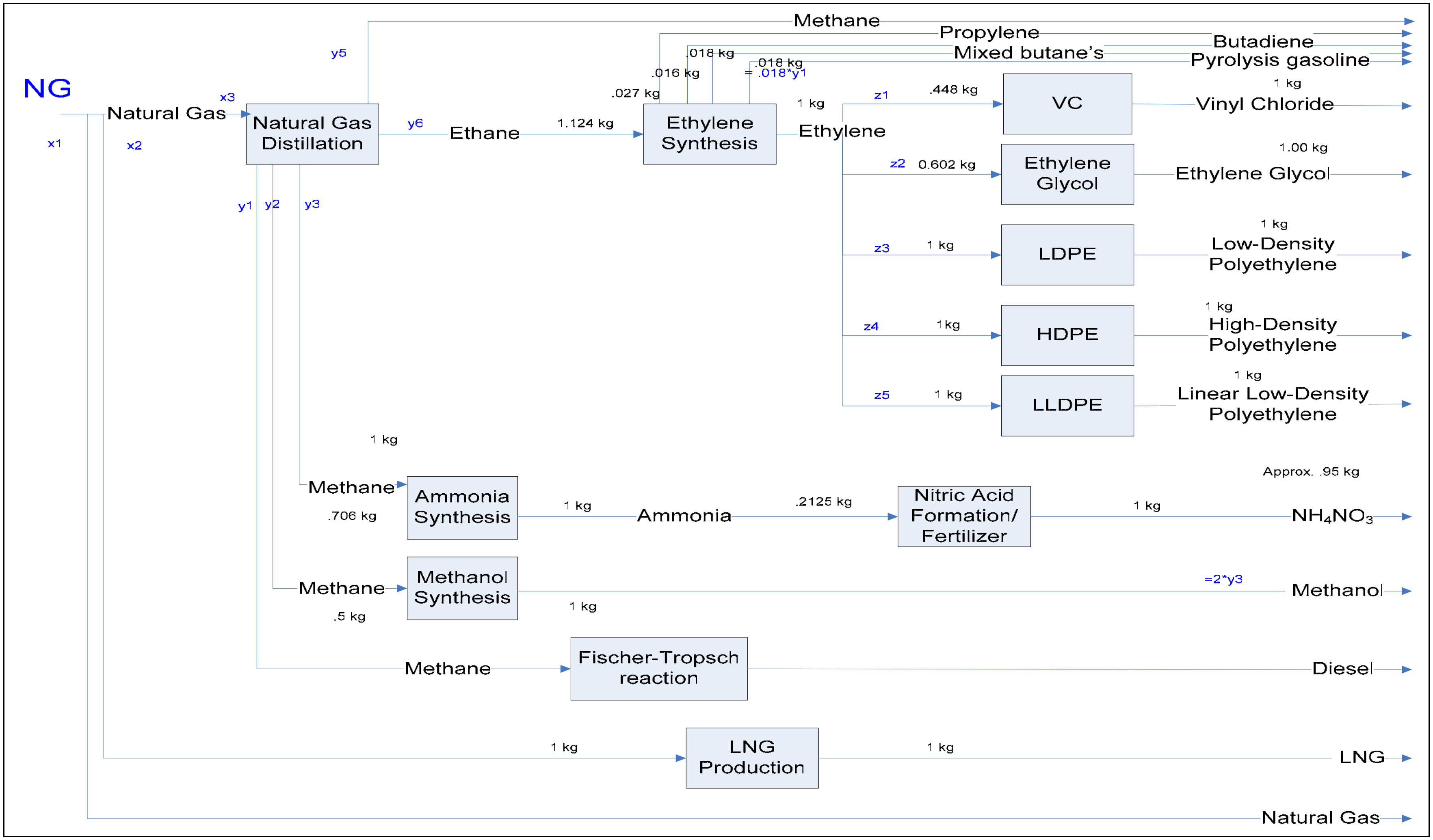


Figure 2.2.1 Process Flow Diagram

3.0 Model Processes

3.1 Natural Gas Distillation

An example of a natural gas distillation column in shown below.

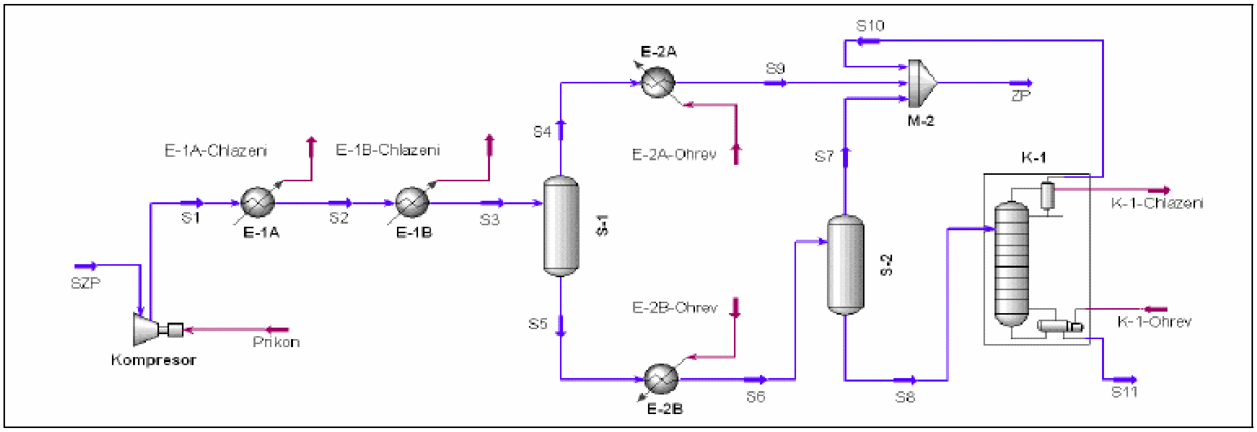


Figure 3.1.1 Natural Gas Distillation FCI versus Capacity

The fixed capital investment of a distillation process was found in PT&W. In an atmospheric distillation column processing 1,600,000 m^3 per day, the fixed capital investment is $23 million. The power factor was used to find other values for the capital investments. They ranged from $8.8 million to $35.7 million for flows from 2.9 kg/s to 21.5 kg/s. The graph may be seen below in Figure 3.1.2.

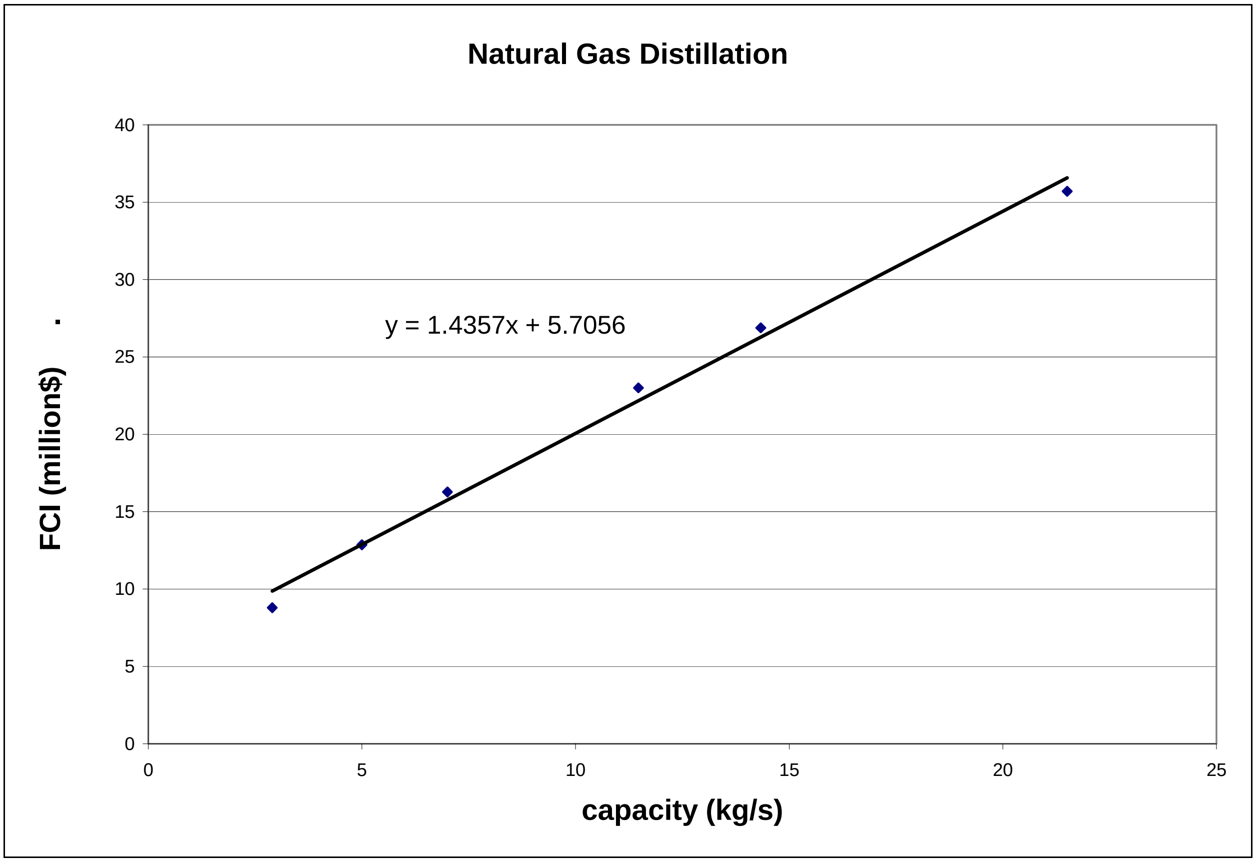


Figure 3.1.2 Natural Gas Distillation FCI versus Capacity

The operating cost trend was found using values given by the website Energy and Energy Savings in Distillation. It was assumed to be linear with respect to flow rate. The operating costs found ranged from 8.5 million dollars/yr and 20.5 million dollars/yr at flow rates of 2.6 kg/s to 7 kg/s, respectively. The trend line is shown in Figure 3.1.3.

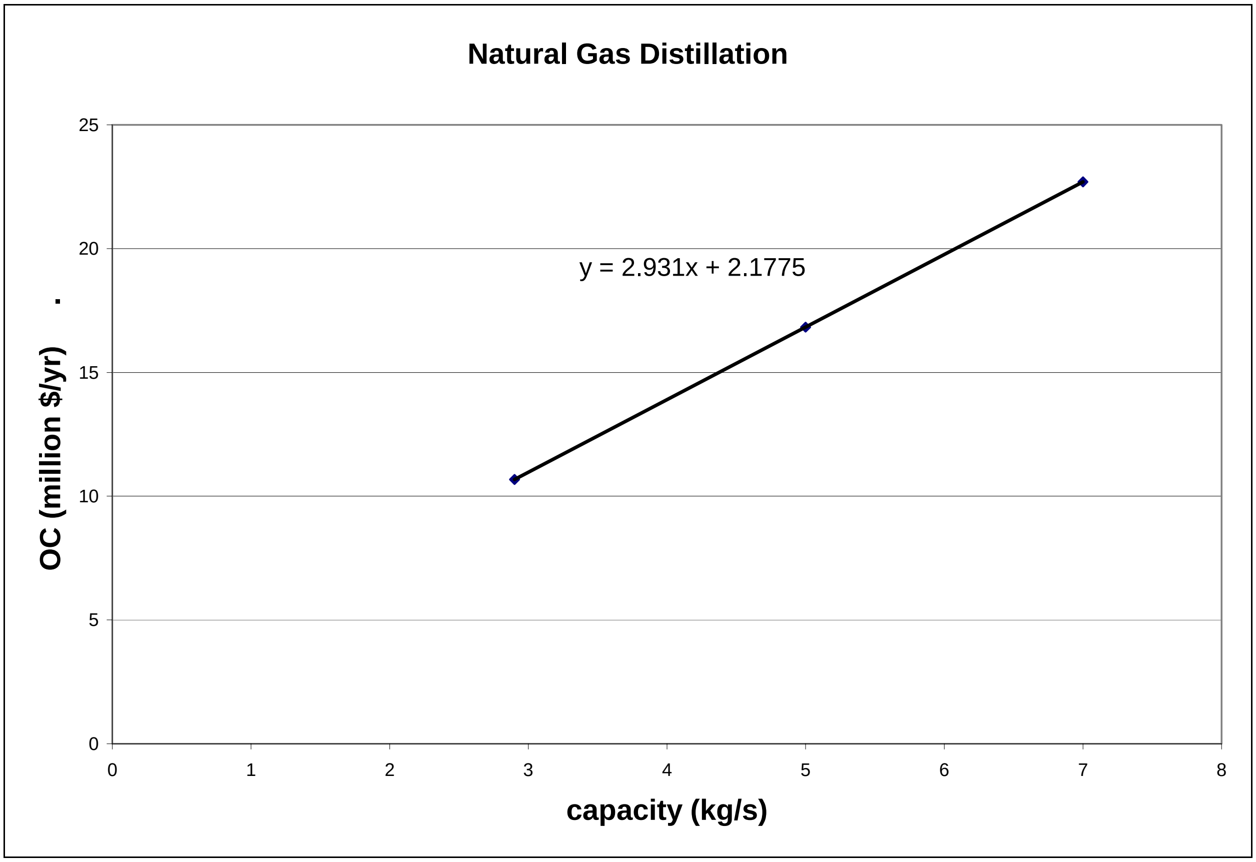


Figure 3.1.3 Natural Gas Distillation Operating Cost versus Capacity

A material balance for the natural gas distillation is shown in Figure 2.2.1. This information and the equations found were entered into the GAMS program, and were used to determine the cost of the process for different capacities.

3.2 Ethylene Synthesis

Two types of technology have been considered for the ethylene synthesis process, ABB Lummus Global SRT cracking technology and Stone & Webster ethylene technology. The description of these methods was given in Petrochemicals Production Processes.

The Lummus Global SRT cracking technology relies heavily on the cracking heater design. This piece of equipment represents 20-30% of the total equipment cost and consumes the majority of the energy in the process. It also determines the product slate, setting the profitability of the plant. A process flow diagram of the Lummus Global SRT Cracking Technology is shown in Figure 3.2.1.

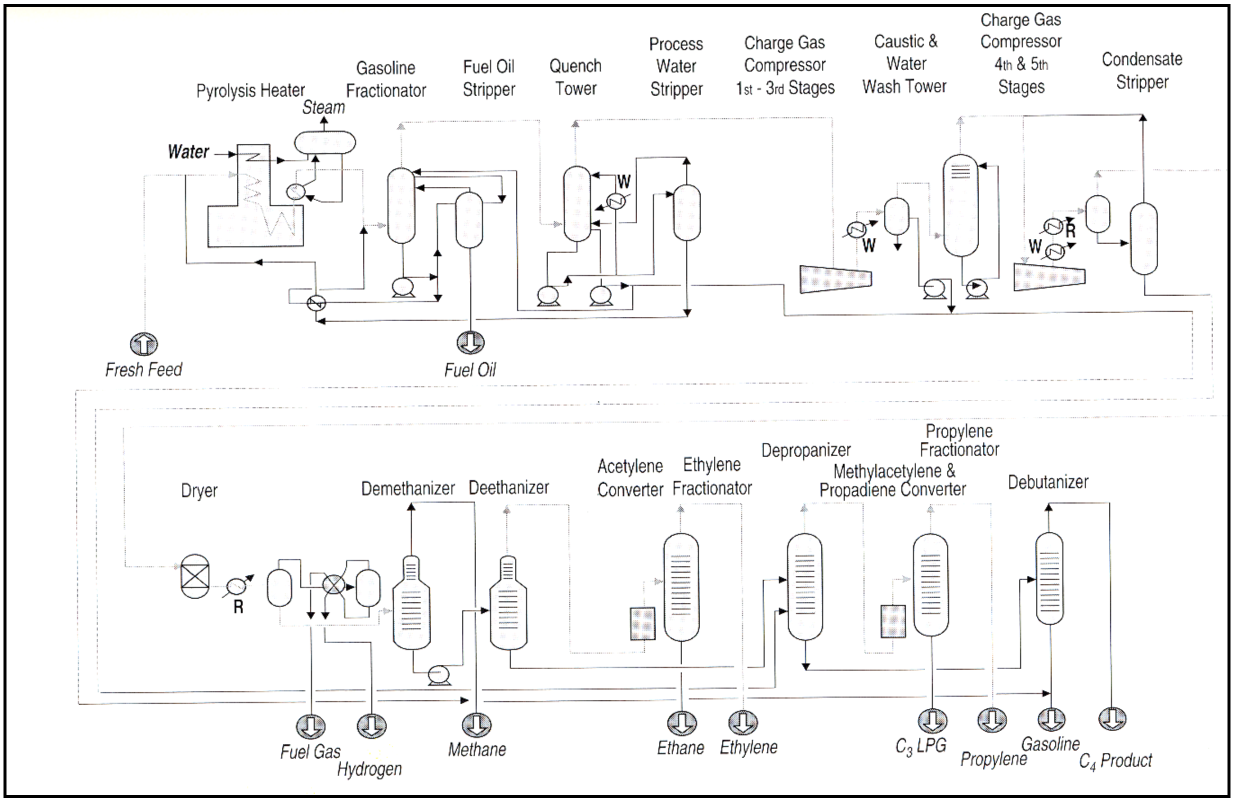


Figure 3.2.1 Lummus SRT Global Cracking Technology

The process consists of 6 areas: oil and water quenching, cracked gas compression and acid gas removal/drying, chilling train and demethanizer, deethanizer and hydrogenation/fractionation, purification of propylene and heavy products, and refrigeration systems.

According to Petrochemicals Production Processes more than one-third of the world's ethylene capacity is based on Lummus's SRT heater technology. In this process ethane feed plants have the lowest total capital investment, which makes this process the most appealing design considered. Therefore, the fixed capital investment was found based on the Lummus SRT technology. The trend line relating fixed capital investment to capacity is shown in Figure 3.2.2. The fixed capital investment ranges from $39 million to $95 million for capacities of 6 kg/s to 32 kg/s, respectively.

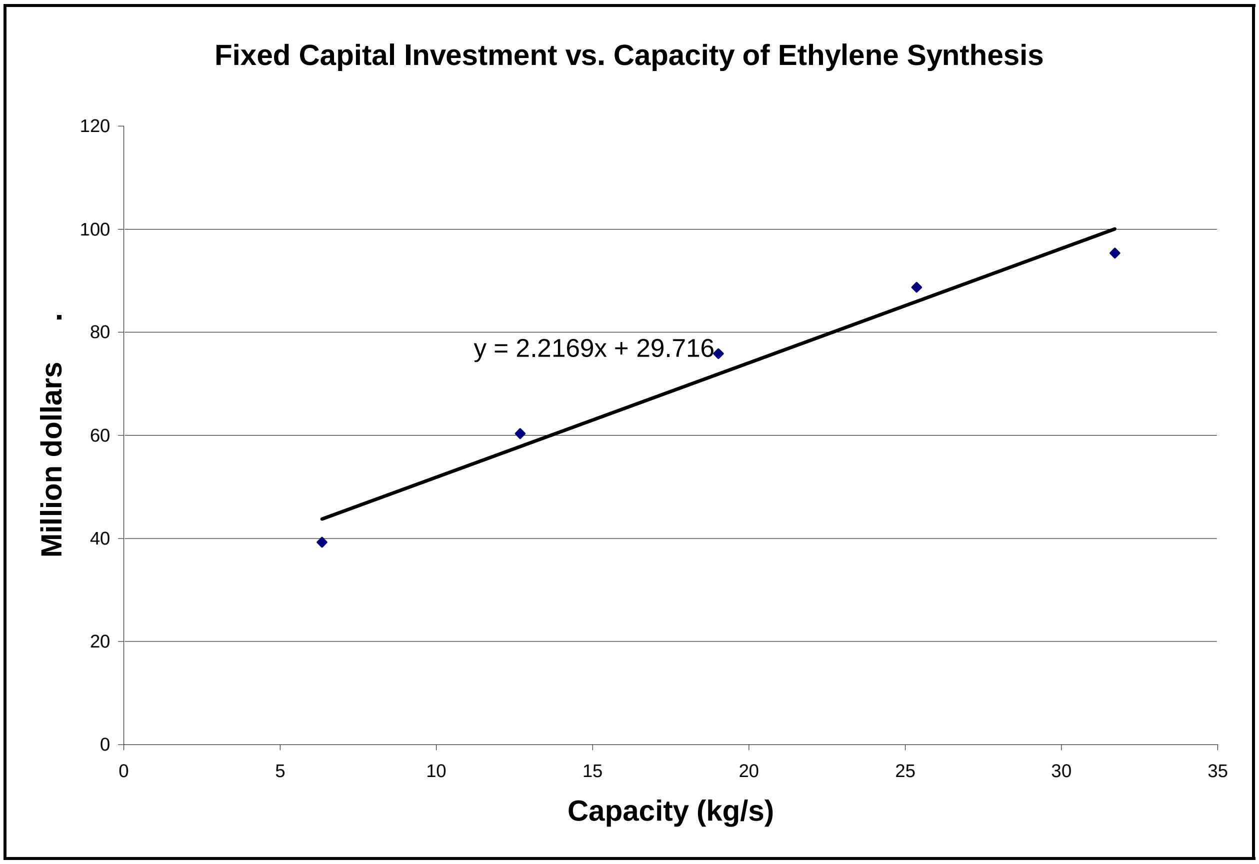


Figure 3.2.2 Ethylene Synthesis FCI versus Capacity

Since the Stone & Webster technology and Lummus technology are so similar, an estimate of the operating cost was found using the Stone and Webster technology. It should be noted that this could lead to some error. A graph of the operating cost versus capacity is shown in Figure 3.2.3. The operating cost ranges from $40 million/yr to $193 million/yr for flow rates of 6 kg/s to 32 kg/s, respectively.

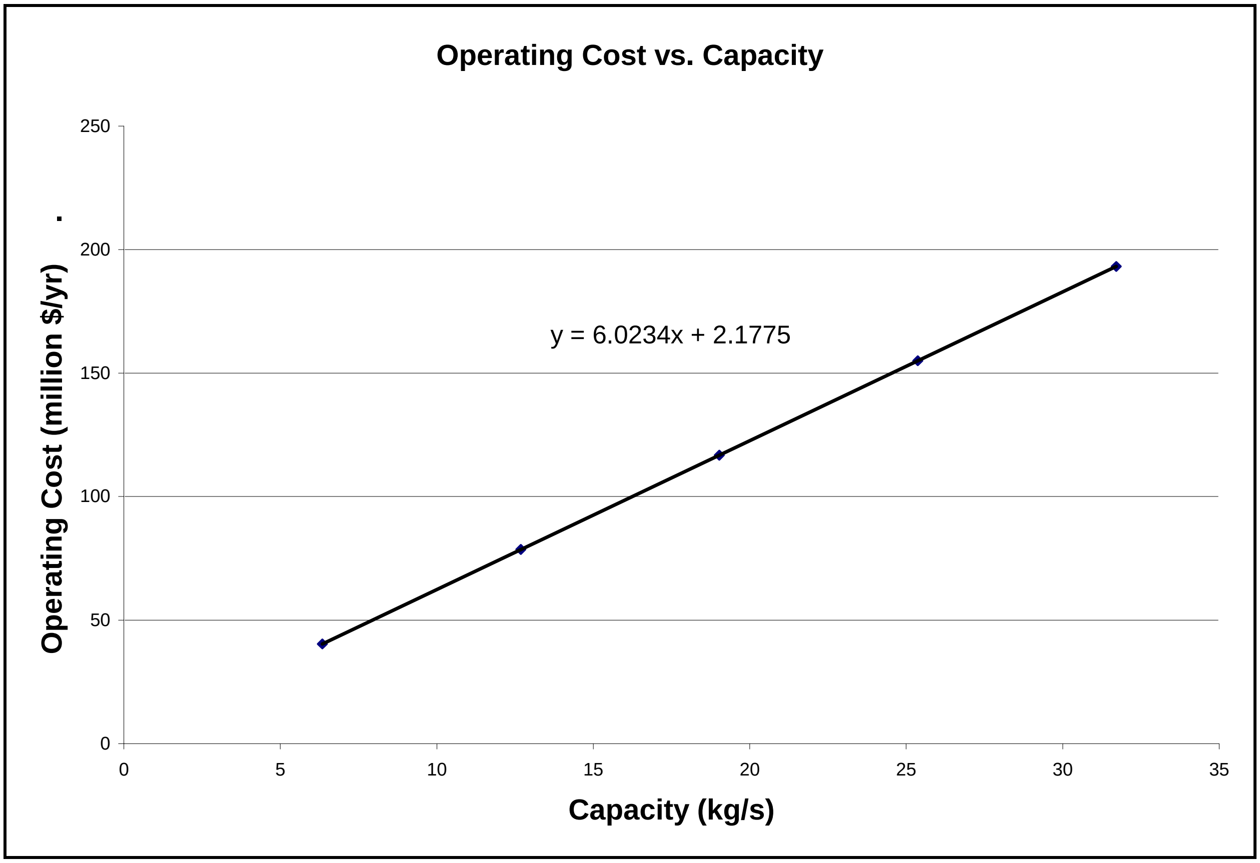


Figure 3.2.3 Ethylene Synthesis OC versus Capacity

A material balance for the ethylene synthesis is shown in Figure 2.2.1. The material balance and equations found from the FCI and OC graphs were entered into the GAMS program, and used to determine the cost of the process with respect to varying capacities.

3.3 Vinyl Chloride Process

There are five different paths to produce vinyl chloride.

1. *Acetylene hydrochlorination:*

$$C_2H_2 + HCI \rightarrow C_2H_3CI$$

2. *Ethylene direct chlorination:*

$$C_2H_4 + CI_2 \rightarrow C_2H_3CI + HCI$$

3. *Ethylene chlorination plus 1,2-dichloroethane pyrolysis:*

$$C_2H_4 + CI_2 \rightarrow C_2H_4CI_2$$

$$C_2H_4CI_2 \rightarrow C_2H_3CI + HCI$$

$$C_2H_4 + CI_2 \rightarrow C_2H_3CI + HCI(overall)$$

4. *Oxychlorination plus 1,2-dichloroethane pyrolysis:*

$$C_2H_4 + 2HCI + \frac{1}{2}O_2 \rightarrow C_2H_4CI_2 + H_2O$$

$$C_2H_4CI_2 \rightarrow C_2H_3CI + HCI$$

$$C_2H_4 + HCI + \frac{1}{2}O_2 \rightarrow C_2H_3CI + H_2O(overall)$$

5. *Chlorination plus oxychlorination pyrolysis:*

$$C_2H_4 + CI_2 \rightarrow C_2H_4CI_2$$

$$C_2H_4 + 2HCI + \frac{1}{2}O_2 \rightarrow C_2H_4CI_2 + H_2O$$

$$2C_2H_4CI_2 \rightarrow 2C_2H_3CI + 2HCI$$

$$2C_2H_4 + CI_2 + \frac{1}{2}O_2 \rightarrow 2C_2H_3CI + H_2O(overall)$$

Reaction 1 is eliminated because it is not economical due to the high price of the raw material acetylene. The price of acetylene is \$1.39/kg, while the price of vinyl chloride is \$0.45/kg. The equation above shows that one mole of acetylene will only produce one mole of vinyl chloride, which means that 26 kg of acetylene would produce 62 kg of vinyl chloride. Thus, \$36.14 of acetylene is needed for \$27.90 of vinyl chloride.

Reaction 2 is eliminated because it is not chemically feasible. When ethylene reacts directly with chlorine, the produce is almost entirely 1,2-dichloroethane (EDC), and not vinyl chloride.
A comparison table for the prices on reactions 3, 4 and 5 is shown in Table 3.3.1:

Table 3.3.1 Price of each process ($/kg)

Reactions	**3**	**4**	**5**
Cl_2	-1.13	0.00	-0.57
HCI	0.58	-0.58	0.00
C_2H_4	-0.45	-0.45	-0.45
C_2H_3Cl	1.00	1.00	1.00
O_2	0.00	-0.26	-0.13
profit	**0.09**	**0.00**	**0.04**

Reaction 4 is eliminated because of its zero excess value of product over reactants.

Reaction 3 shows a higher excess value compared to reaction 5. However, the excess value is solely dependent on the by-product HCI. Reaction 3 would produce a negative profit if HCl is not sold. Therefore, the choice between reaction 3 and 5 depends on the sales of HCI. According to research, 95% of all HCI produced in USA is from the by-product of chlorination process. Reaction 5 is the only process that proves to be profitable.

Reaction 5, which is the chlorination plus oxychlorination pyrolysis is the only commercial process in the market that produces vinyl chloride from ethylene. The process chosen to produce vinyl chloride in this plant is the Vinnolit process, since it is the only technology available.

Figure 3.3.1 shows the flow diagram of the Vinnolit vinyl chloride process, with the molar flow rate of each component at each stream. The energy needed or released at some related equipments are also shown.

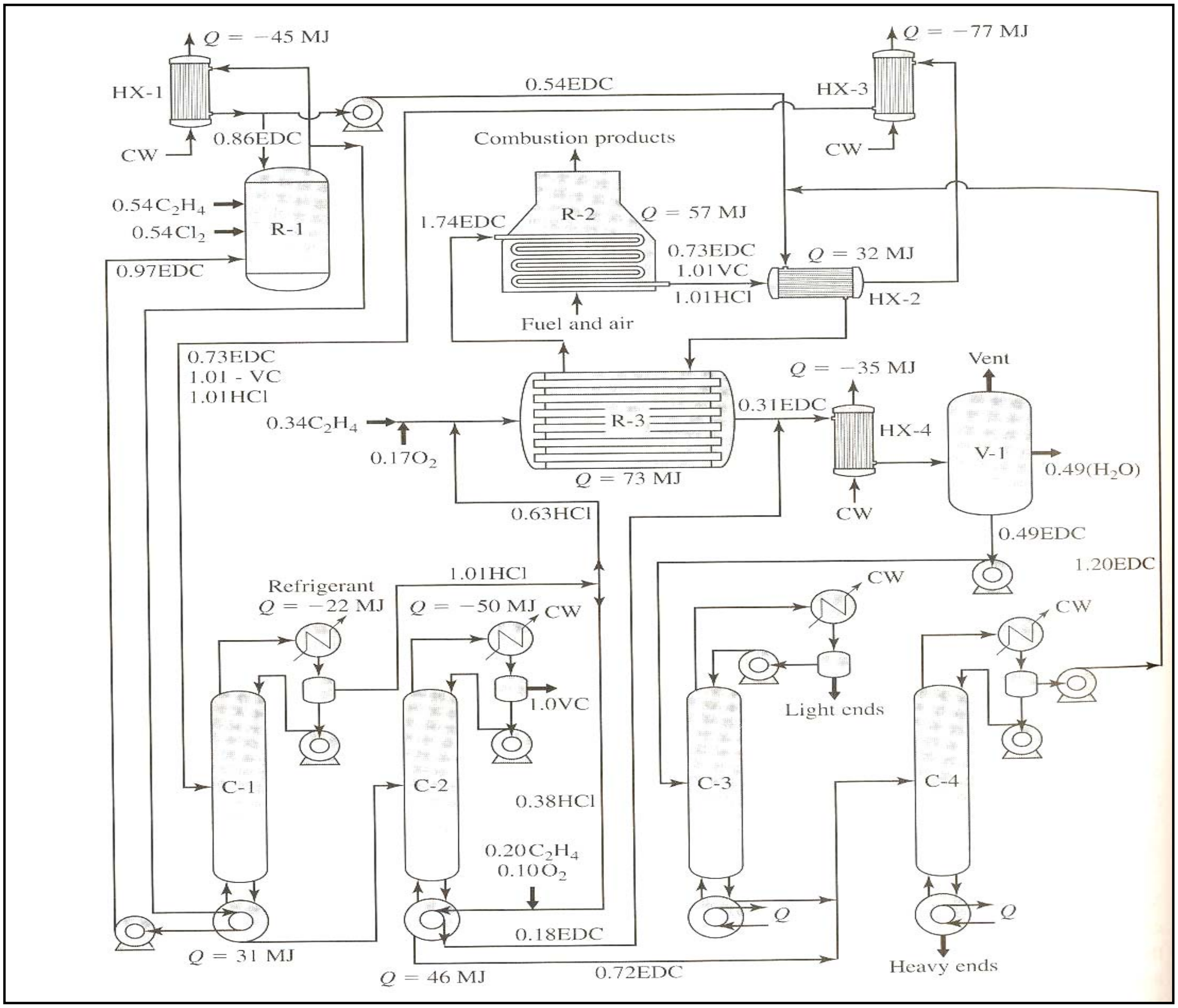


Figure 3.3.1 Flow diagram of Vinnolit vinyl chloride production

The production of vinyl chloride from ethylene and chlorine is based on two different routes: direct chlorination or oxychlorination. The direct chlorination takes place in R1, and the oxychlorination takes place in R3 (shown in Figure 3.3.1). The product 1,2-dichloroethane (EDC) is produced from both routes. EDC is then cracked in R2 to produce vinyl chloride and hydrochloride acid. The EDC from direct chlorination can be sent to the cracking unit directly, while the EDC from the oxychlorination needs to be distillated and purified before it is sent to the cracking unit.

In order to calculate the FCI of this process at different capacities, each piece of equipment in the flow diagram was sized based on the material balances of the inlet and outlet streams. After the total equipment cost was quoted, the FCI was calculated by the method outlined by PT&W.

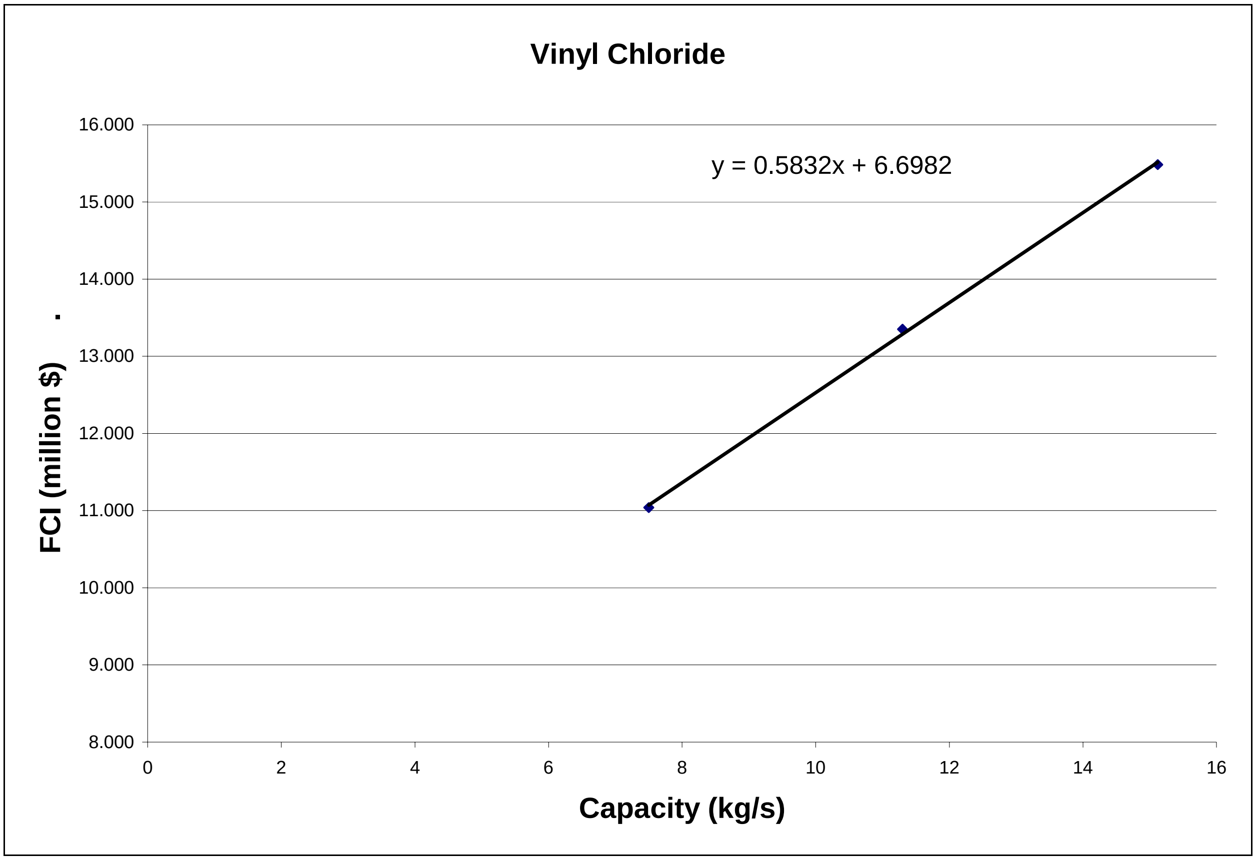


Figure 3.3.2 FCI vs. capacity for vinyl chloride process

In order to calculate the operating cost of the vinyl chloride process, the determining factors were electricity cost, cooling water cost, refrigerant cost, and fuel cost. The operating cost of each capacity was also calculated with the method outlined in PT&W with some reasonable modifications for each capacity. Figure 3.3.3 shows the trend line of the operating cost at different capacities for the vinyl chloride process.

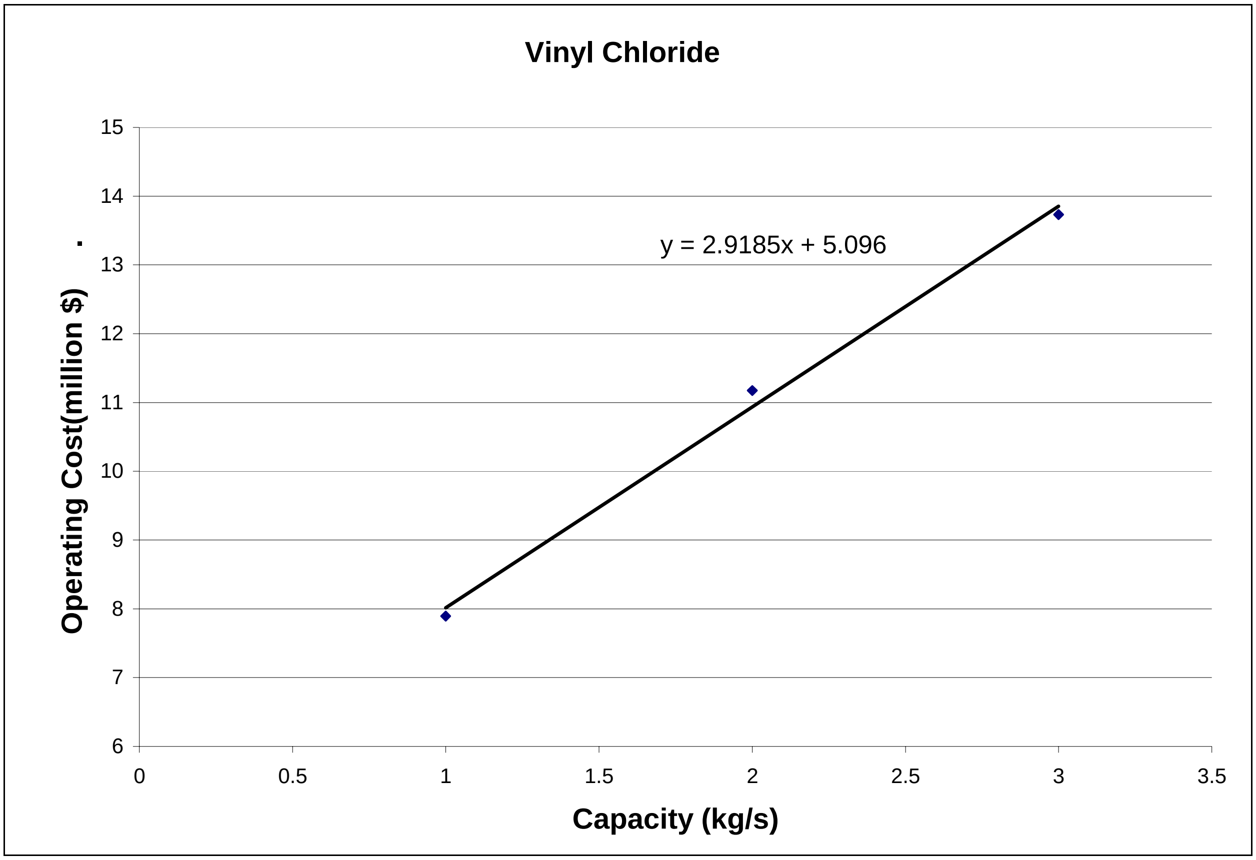


Figure 3.3.3 Vinyl Chloride process operating cost versus capacity

3.4 Polyvinyl Chloride Process

The polyvinyl chloride process is a process of "polymerization" that links vinyl chloride monomers into polymer chains. The following reaction shows the polyvinyl chloride process.

$$-nCH_2 = CHCl \rightarrow [-CH_2 - CHCl-]_n$$

There are two types of PVC that can be produced from polyvinyl chloride; these are suspension-PVC (S-PVC) and emulsion or paste-PVC (E/P-PVC). The applications of S-PVC are pipes, construction material, bottles, cables, and bags. The applications of E-PVC are flooring, coated fabrics, and wall covering.

The process that produces S-PVC is chosen because the market of S-PVC is much larger than E-PVC based on the applications listed above. Most of the manufacturing of E-PVC is dominated by Western Europe where the market is saturated. The market for PVC is shown in Figure 3.4.1. As shown, the majority of the market in Western Europe is S-PVC.

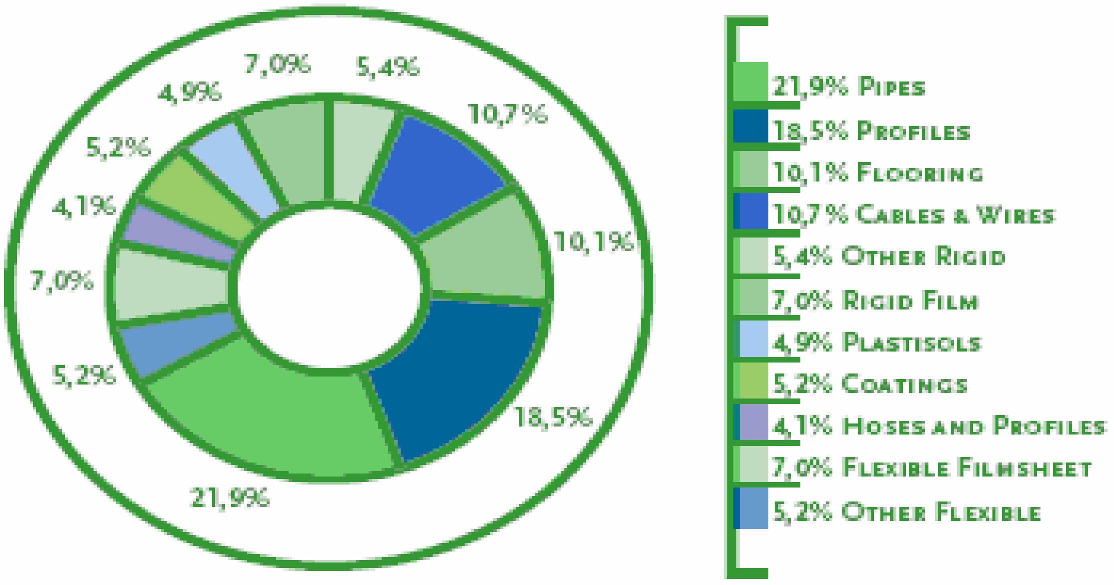


Figure 3.4.1 Consumption of PVC

The process of S-PVC is describes briefly in Figure 3.4.2.

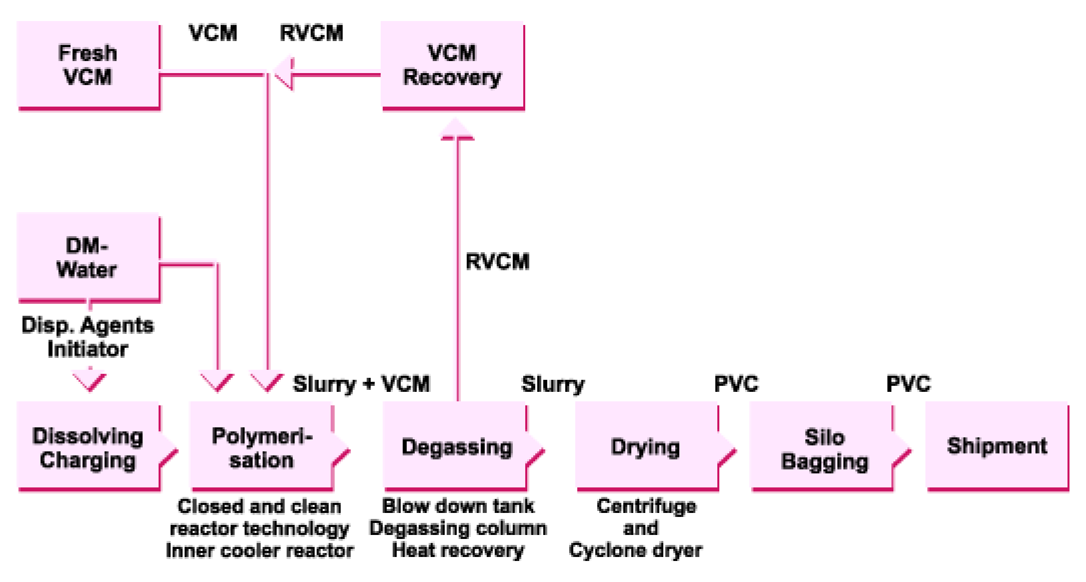


Figure 3.4.2 Flowchart of S-PVC process

Three main parts in the S-PVC process are polymerization, degassing and drying, as shown in Figure 3.4.3. Fresh vinyl chloride monomer (VCM), activators, and deionized water are sent into a reactor for polymerization. They are then sent to a blow down vessel where most of the unreacted VCM is recycled. The function of the degassing section is to remove the remaining unreacted vinyl chloride. After degassing, the PVC-slurry is separated from water in decantation centrifuges. Drying of the PVC is carried out in the cyclone drier.

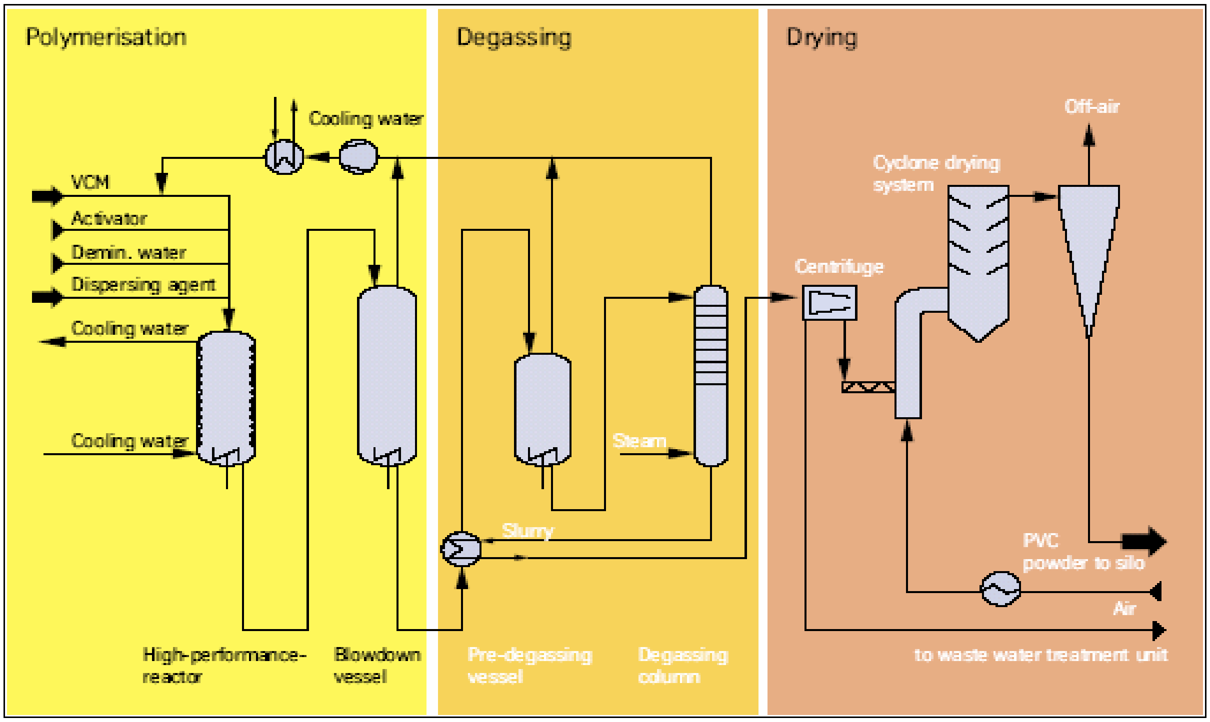


Figure 3.4.3 Flow diagram of S-PVC process

In order to calculate the FCI and operating cost of the process at different capacities, the method outlined in PT&W was used with some reasonable modifications. Each piece of equipment is quoted from the flow diagram of S-PVC process. The operating cost of each process at different capacities is a function of fuel cost, electricity cost, labor cost, cooling water cost, deionized water cost, saturated steam cost, and additives cost. Figure 3.4.4 shows the trend line of FCI versus capacity for S-PVC process.

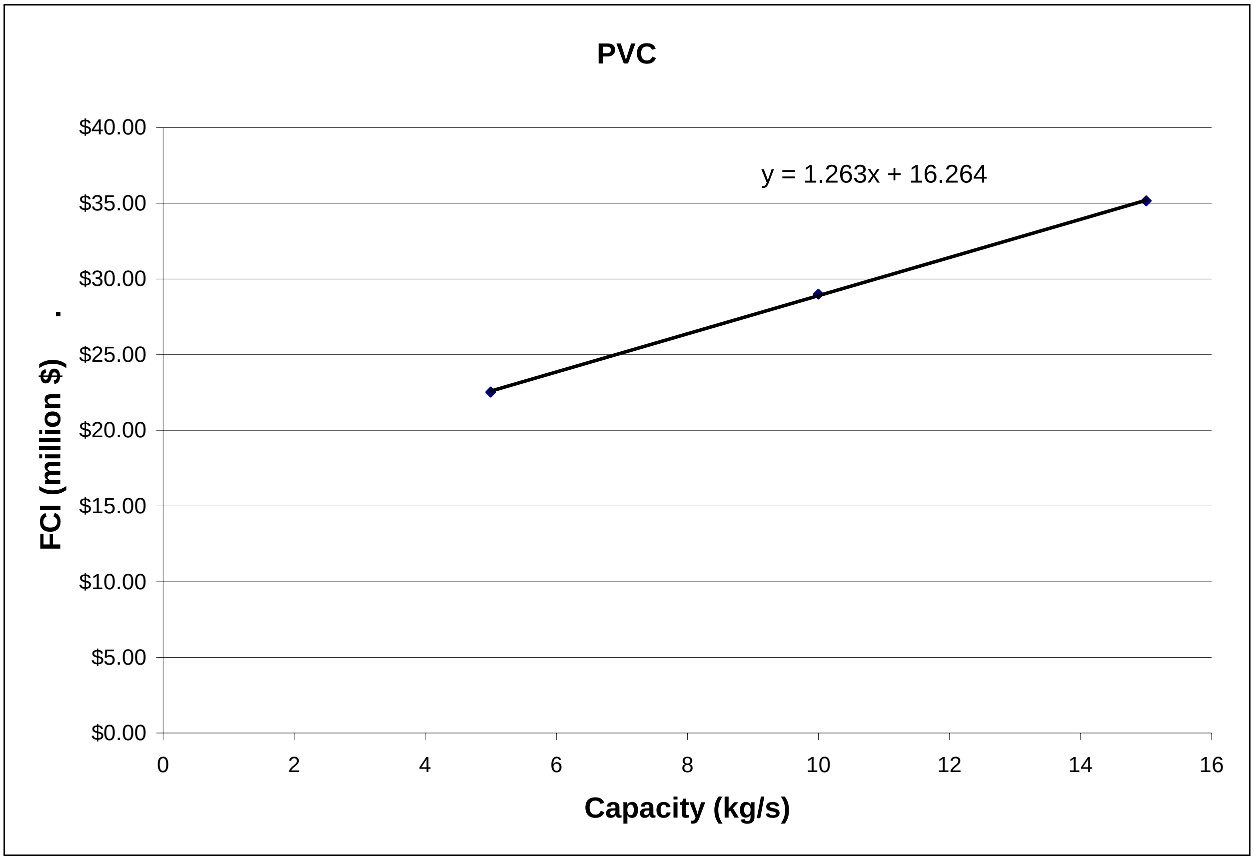


Figure 3.4.4 S-PVC FCI versus capacity

Figure 3.4.5 shows the trend line of operating cost versus capacity for S-PVC process.

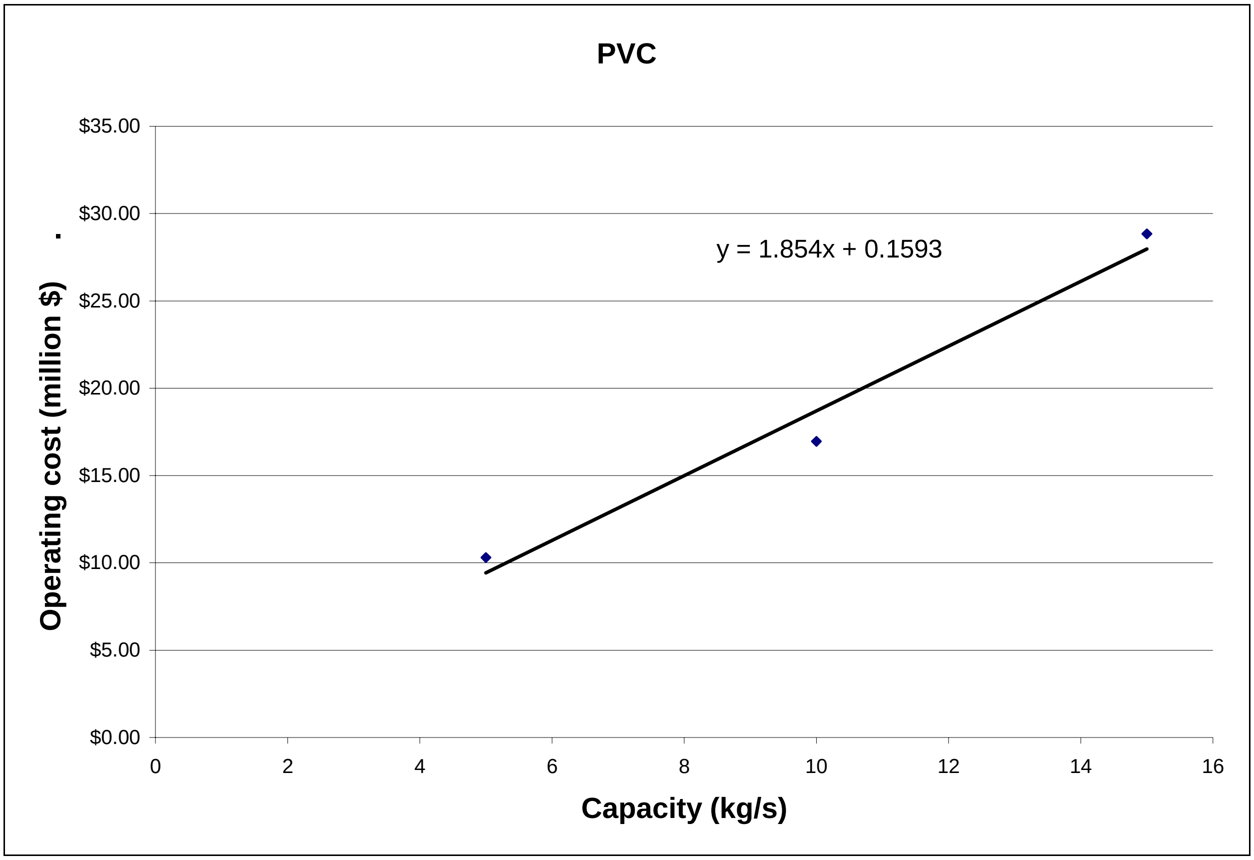


Figure 3.4.5 S-PVC Operating cost versus capacity

3.5 Ethylene Glycol Process

Most of the ethylene glycol in the world is produced by the hydration of ethylene oxide. Such an ethylene glycol process includes two separate plants. The first plant converts ethylene to ethylene oxide. The second plant converts ethylene oxide to ethylene glycol.

Figure 3.5.1 describes an ethylene oxide plant. Oxygen-based ethylene oxide plant has been proved to be more economical than air-based ethylene glycol plant. Oxygen and ethylene are mixed together and react in a multi-tube reactor. The reaction effluent goes through CO_2 removal and product upgrading.

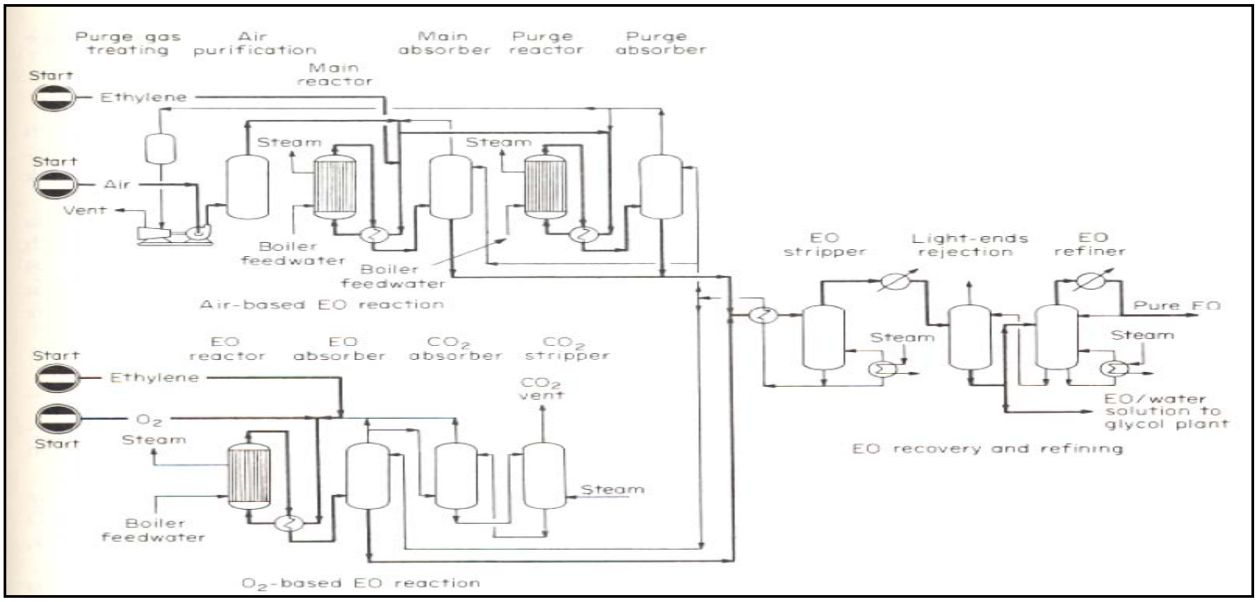


Figure 3.5.1 Schematic Ethylene Glycol Plant Flow Diagram

Figure 3.5.2 describes an ethylene glycol plant (Meyers). Ethylene oxide and water are mixed in a tank. The mixture is compressed and fed to a reactor where ethylene glycol is produced. The reaction effluent goes through a series of evaporators from medium pressure level to very low pressure where water and impurities are removed.

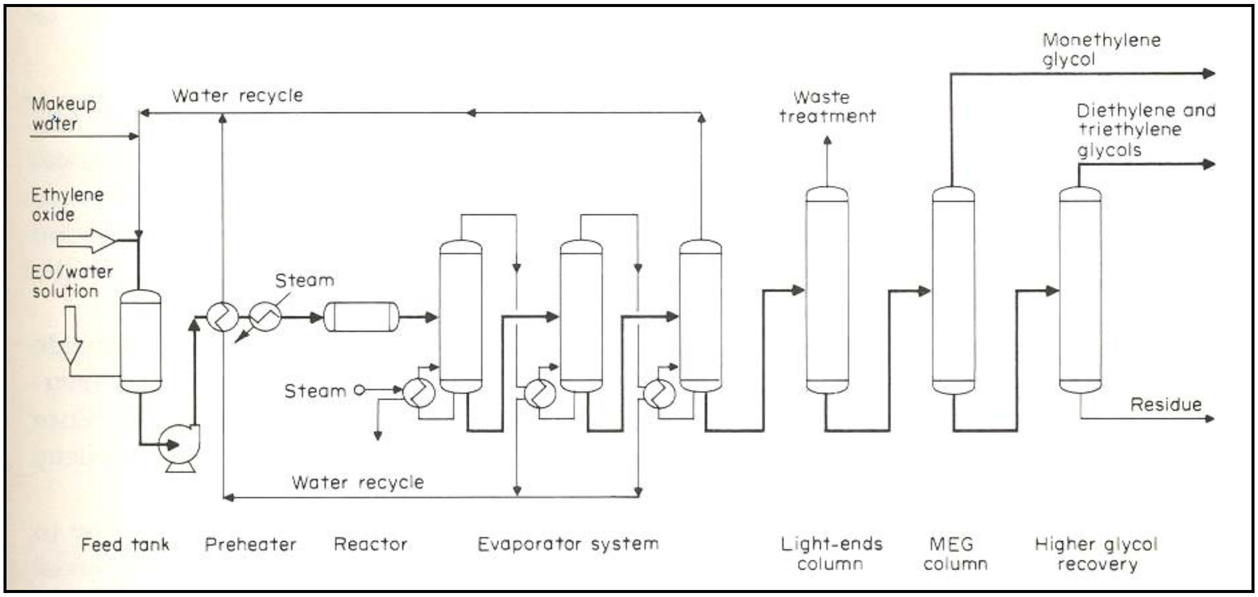


Figure 3.5.2 Schematic Ethylene Glycol Plant Flow Diagram

Equipment in the plants was sized by performing material and energy balances on the process flow diagrams. The equipment cost and FCI were calculated for the process, the combination of the two plants above. Figure 3.5.3 shows a relationship of FCI versus capacity. These costs include the control room, equipment, piping and instrumentation, and in-process surge and product rundown tankage required for normal operation. It also includes complete engineering, procurement services, and resident engineering.

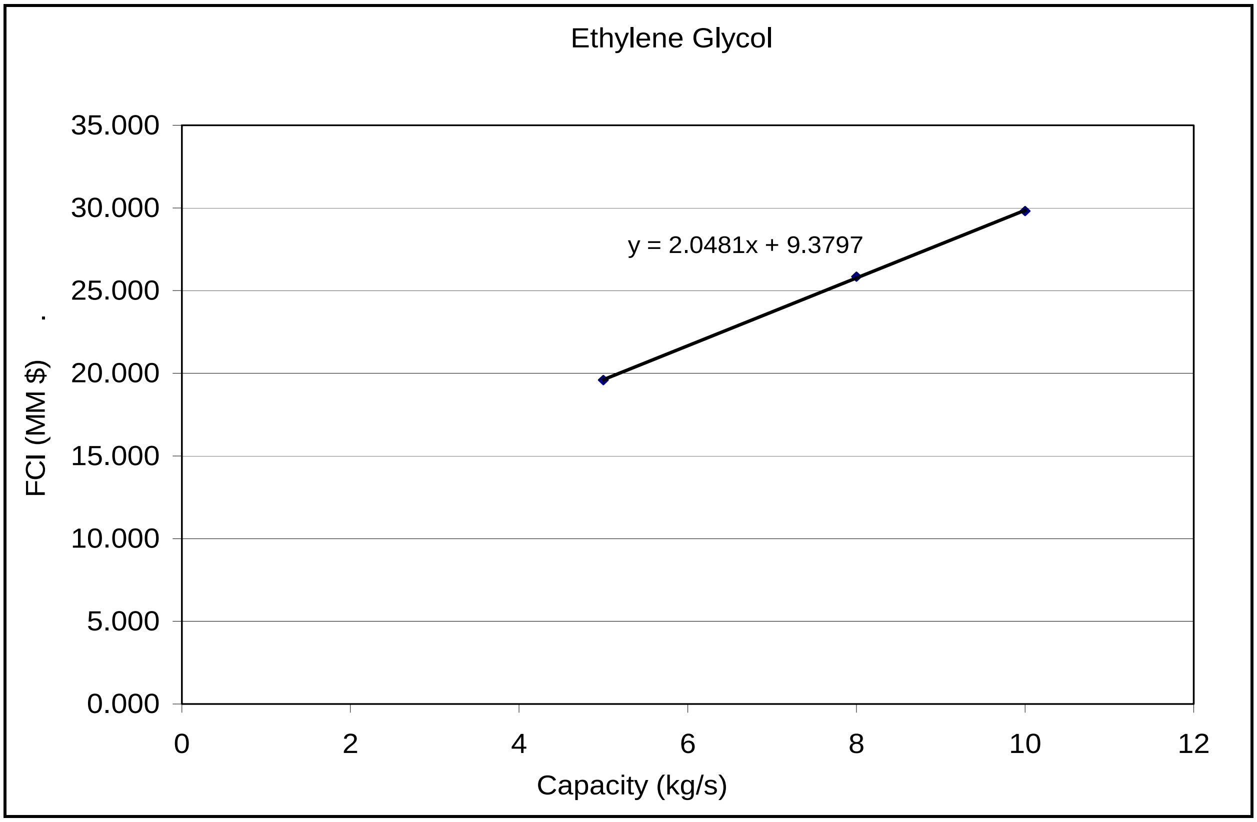


Figure 3.5.3 Ethylene Glycol FCI versus Capacity

The operating costs were calculated using the online spreadsheet from PT&W. The amount of utilities and raw material used were determined by performing energy and material balance on the system. A graph showing the relationship of operating cost versus capacity is shown in Figure 3.5.4.

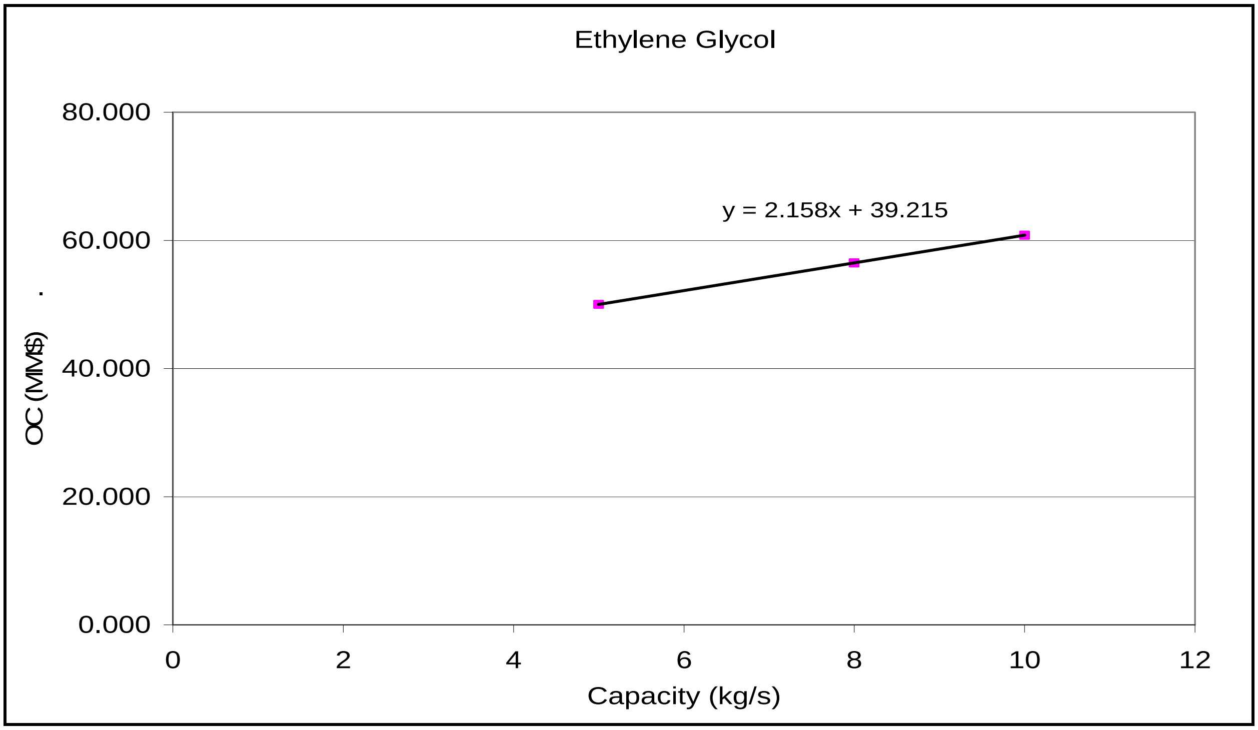


Figure 3.5.4 Ethylene Glycol Operating Cost versus Capacity

A material balance for the ethylene glycol is shown in Figure 2.2.1. This information and the equations found are entered into the GAMS program, and are used to determine the cost of the process when different capacities are used.

3.6 Polyethylene

3.6.1 Low-Density Polyethylene Process

Two processes were considered for the low density polyethylene (LDPE) plant: ExxonMobil High-Pressure Process and Polimeri Europa. The Polimeri Europa process utilizes a choice of an autoclave or tubular reactor in which the resin density is lowered to create a long-chain branching property making LDPE easier to produce. This high-pressure process, however, requires the use of special equipment and technology.

There are benefits to the Polimeri Europa process. The Ziegler-Natta catalytic system allows for flexibility in products, allowing the production of most grades. The recently developed technology of this system has allowed capacities up to 400,000 metric tons annually and is a very efficient process. Overall this process has good performance. High conversion rates and large-scale reactors, specific equipment and machinery design, and operating procedures allow low net monomer consumption and energy requirements.

The ExxonMobil process has high operating efficiencies, control over product properties, outstanding product quality, conversion rates up to 40 percent, excellent reactor designs, and high capacity capabilities. ExxonMobil has continued to develop and enhance the basic process technology because it remains one of the acknowledged industry leaders in the production of LDPE. Although the Polimeri Europa and ExxonMobil processes are both proven to be excellent, the ExxonMobil process was chosen because it is more predominant in the world. The ExxonMobil High-Pressure Process is shown in Figure 3.6.1.1.

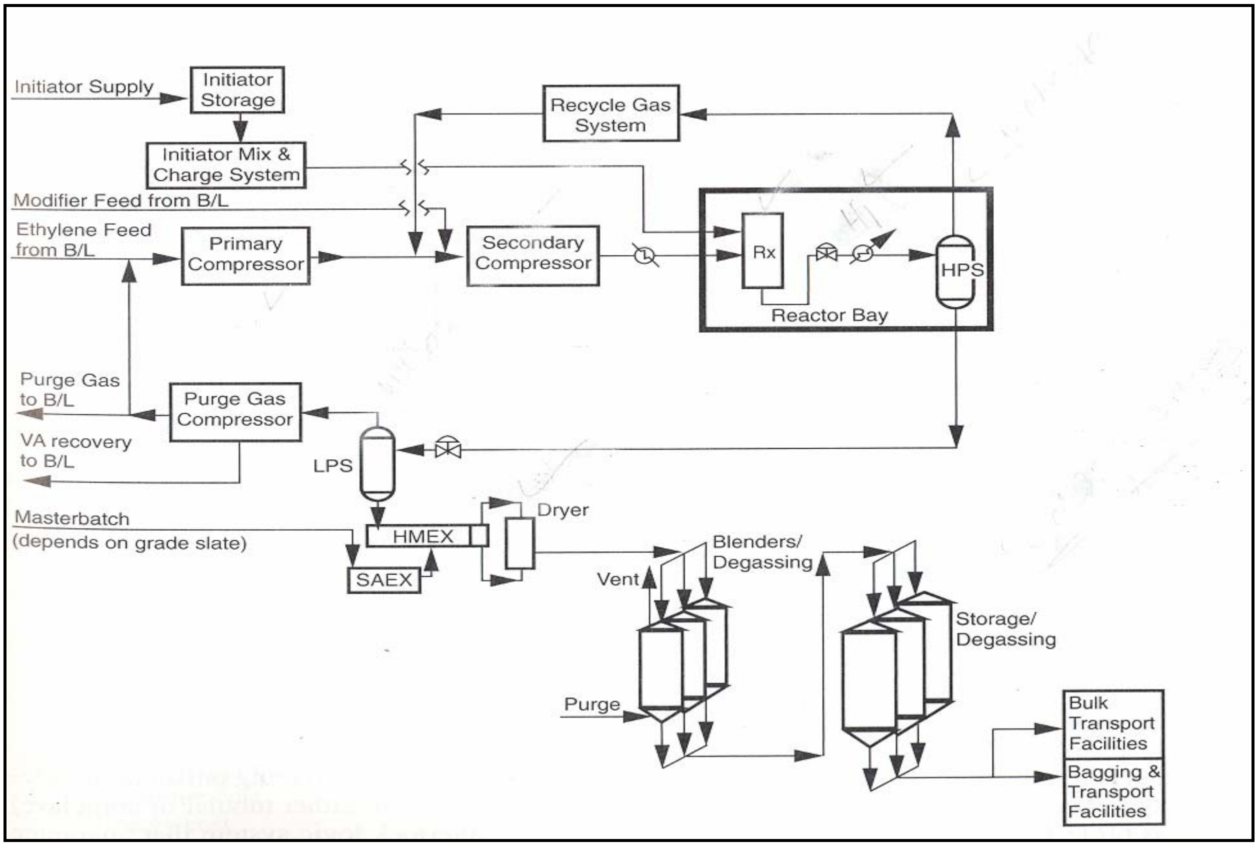


Figure 3.6.1.1 ExxonMobil Low Density Polyethylene Process

The equipment cost was calculated by finding the price of each apparatus. The most costly piece of equipment is the reactor. The two types of reactors to choose from are autoclave and tubular. The autoclave reactor is lower in cost and therefore, was used in the plant design.

The overall polymerization reaction occurring in the reactor is shown below.

$$n(CH_2{=}CH_2) \rightarrow (-CH_2\text{-}CH_2\text{-})_n$$

The cost estimator function from the PT&W website and a description of the ExxonMobil process were used to estimate the cost of each apparatus. The methods used to size each piece of equipment are described below.

The residence time is 20-40 seconds for the most powerful autoclaves. The volume of autoclaves usually ranges between 1000 to 2000 L. For this reason the lowest flow considered, 10 kg/s, was assumed to have a 1000 L autoclave reactor. Using this information, the reactor volume and cost was found.

Organic peroxide solutions (to create oxidation) are injected at multiple points into the reactor to initiate the free-radical exothermic polymerization reaction. The amount of organic peroxides needed was found using a mole balance of the system. This was then used to find the mixing tank volume required. The initiator volume was found from the total inlet flow of the organic peroxides and ethylene.

The pump type was found using the volumetric flow rate and Figure 12-25 in PT&W. A description of the conditions of both pumps was given in Handbook of Petrochemicals Production Processes.

Changes in temperatures across the length of the heat exchangers were given in Handbook of Petrochemicals Production Processes. Using the inlet and outlet temperatures, the heat exchangers were sized and priced.

It was decided from Table 15-16 in Plant Design and Economics for Chemical Engineers and the U-value that a fluid bed dryer would be the most appropriate heat exchanger. The fluid bed is good for fine, free-flowing powders, very small granules or fibrous solids. It has a high capacity, low cost, high heat transfer, and uniform internal temperatures. It is limited to solids that can be fluidized and not fractured in high-velocity gas streams. Figure 15-34 in Plant Design and Economics for Chemical Engineers was used to estimate the cost. This was done using the solids velocity and L/D ratio. According to Plant Design and Economics for Chemical Engineers the conversion of ethylene to polymer can be roughly calculated close to 20 percent per pass. A solids velocity of 0.01 m/s was assumed. The minimum L/D value for a fluidized bed is 3-15 according to table 15-16 in PT&W. It is therefore estimated that the fluidized bed L/D is equal to the average, 7.65.

The recycle gas system was sized using the conditions provided by the Basell Hostalen Technology for Bimodal HDPE production. In this process, the gas recycling unit consists of a two-step distillation process. The first step contains an evaporator and the second step contains a distillation column. A column diameter was assumed for the distillation column ranging from 2m to 4m, with 2m being the lowest flow rate. This assumption was also used when finding the surface area used to size the evaporator.

The conversion of ethylene to polymer can be roughly calculated by the adiabatic temperature difference across the reactor and is close to 20% per pass. Therefore, the flow of the feed entering the purge gas compressor is roughly 80% of the initial flow rate. The purge-gas compressor boosts the low-pressure separator offgas to the suction pressure of the primary compressor. The low-pressure separator offgas was estimated to be about 0.7 bar according to the Handbook of Petrochemicals Production Processes. This information was used in the PT&W website to estimate the compressor cost.

The breakdown of the equipment cost may be found in the appendix. The trend line relating the fixed capital investment with respect to capacity is shown in Figure 3.6.1.2. The fixed capital investment ranges from $55 million to $234 million with flows of 10 kg/s to 50 kg/s.

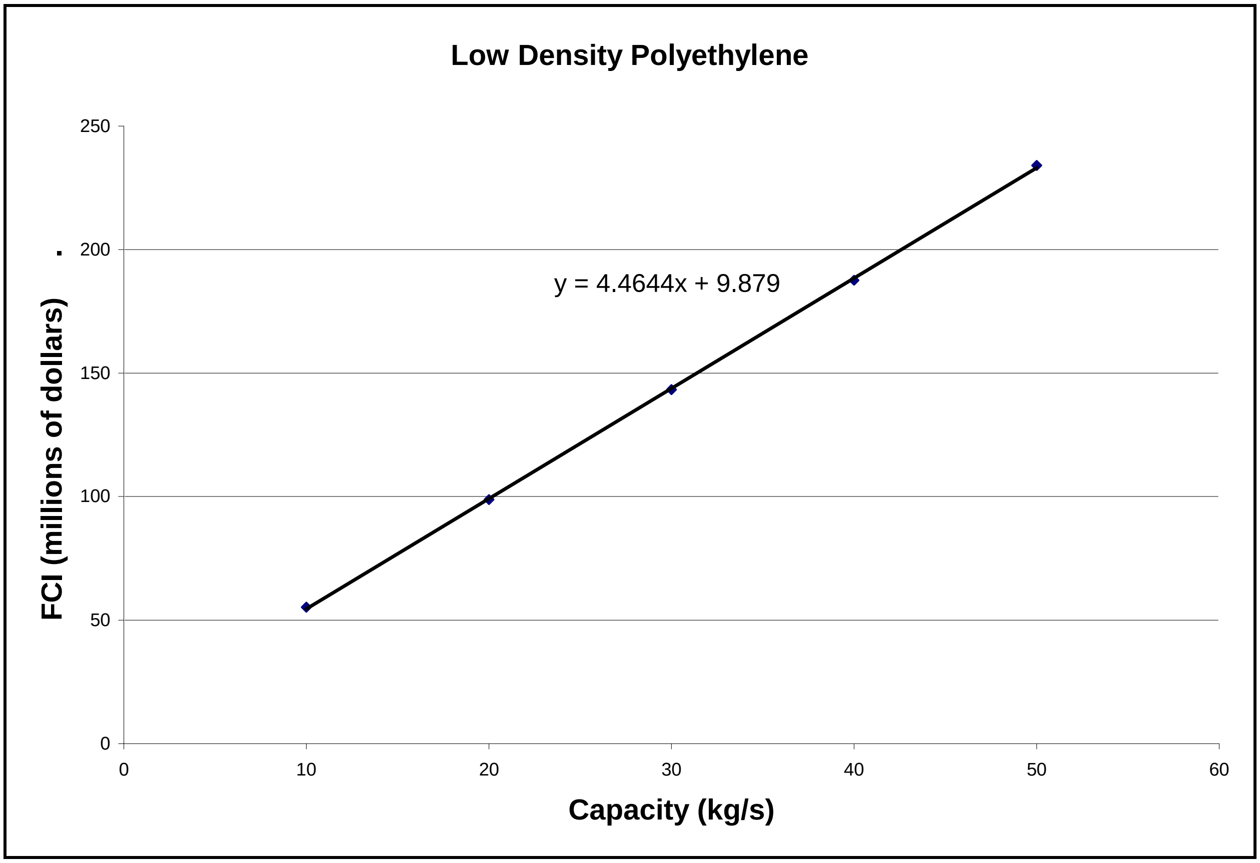


Figure 3.6.1.2 Low Density Polyethylene FCI versus Capacity

Due to similarities to the ExxonMobil process, the Imhausen LDPE process was used in order to calculate the operating cost. It should be noted that this estimation could lead to error. The operating cost trend line is shown in Figure 3.6.1.3.

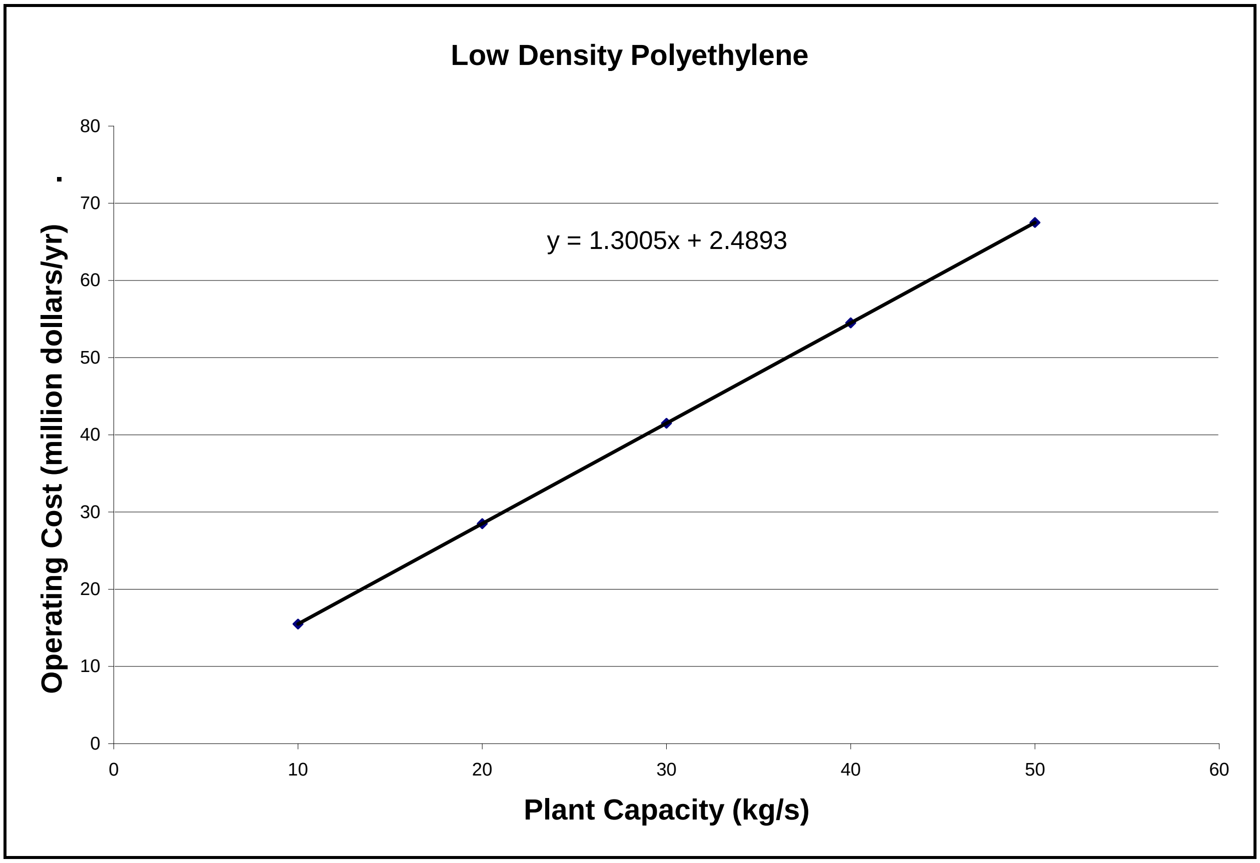


Figure 3.6.1.3 Low Density Polyethylene OC versus Capacity

A material balance for the low density polyethylene process is shown in Figure 2.2.1. The material balance and equations found from the FCI and OC graphs were entered into the GAMS program, and used to determine the cost of the process with respect to varying capacities.

3.6.2 High-Density Polyethylene

The process economics used to find the operating costs and capital investments come from the Handbook of Chemicals Production Processes. There are two different processes discussed: Hoechst and Phillips. These two processes were studied in order to find the one that provides the highest profit.

The Hoechst process is broken down into 3 steps. In the first step a batch of Ziegler-Natta catalyst is prepared. This mixture is fed along with the gas mixture of ethylene, comonomer, and hydrogen to the reactor, where polymerization takes place. In the second step, the diluent is separated and the polymer powder is dried. The third step involves palletizing of the powder and adding special stabilizers or dyes.

Some advantages of these processes are that the Ziegler-Natta catalysts are relatively easy to find and fairly inexpensive, the molecular weight of the polyethylene can be controlled, the frequent purification is not necessary, and optimal steady-state process behavior is able to be achieved (due to hexane being used as a diluent). The investment costs are also low due to the simplicity of the process. Small quantities of catalyst and optimization of utilities make the energy costs low. Although the Hoechst process operation is simple, ethylene conversion is just 90%.

The Phillips process, however, was chosen because it is more similar to the predominant technologies used today. Hoechst process uses technology that is not used as frequently in industry. A flow sheet for the Phillips high density polyethylene process is shown in Figure 3.6.2.1.

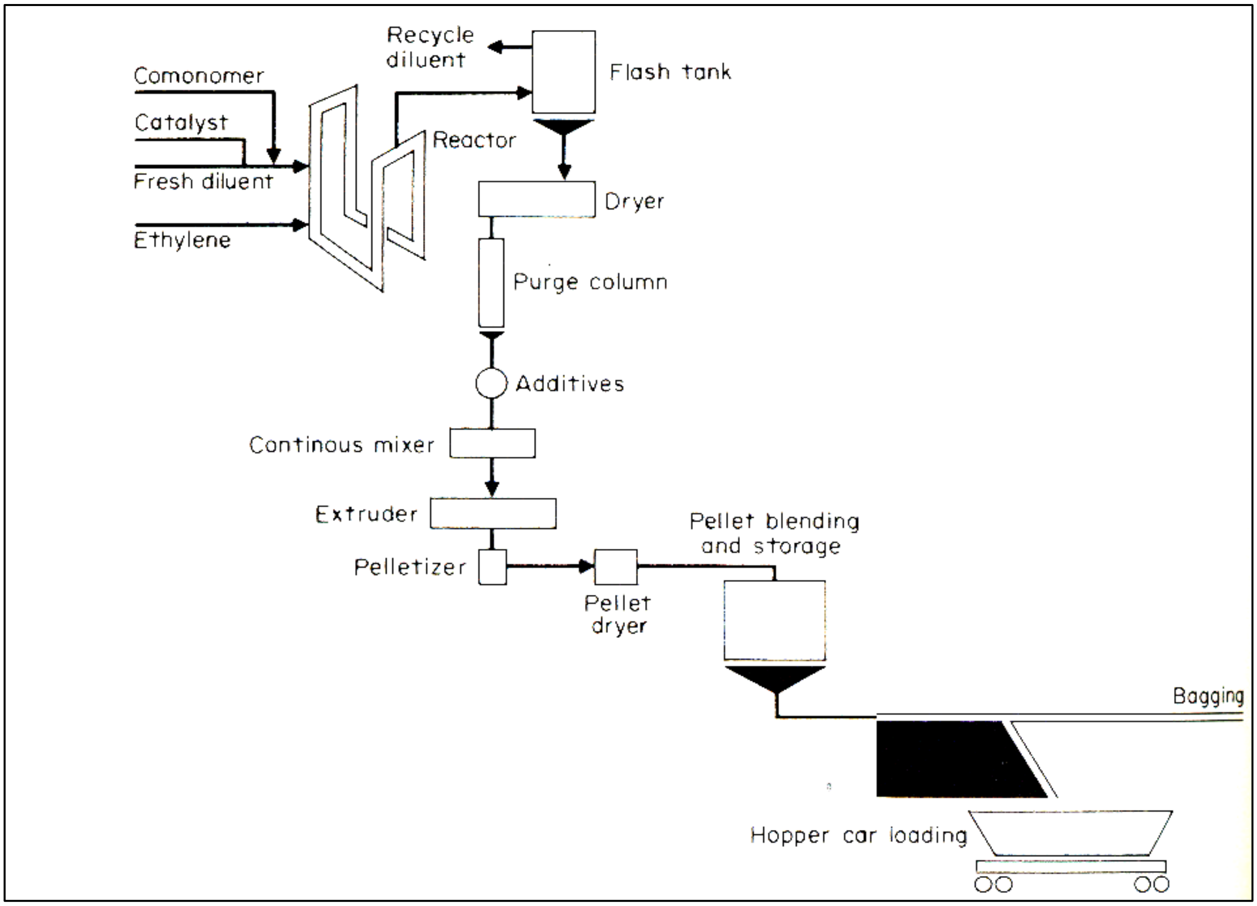


Figure 3.6.2.1 High Density Polyethylene Process Flow Diagram

The fixed capital investment was found from research already conducted by Phillips. The fixed capital investment ranges from \$29 million to \$54 million for capacities of 2 kg/s to 4 kg/s, respectively. The trend line found for the fixed capital investment is shown in Figure 3.6.2.2.

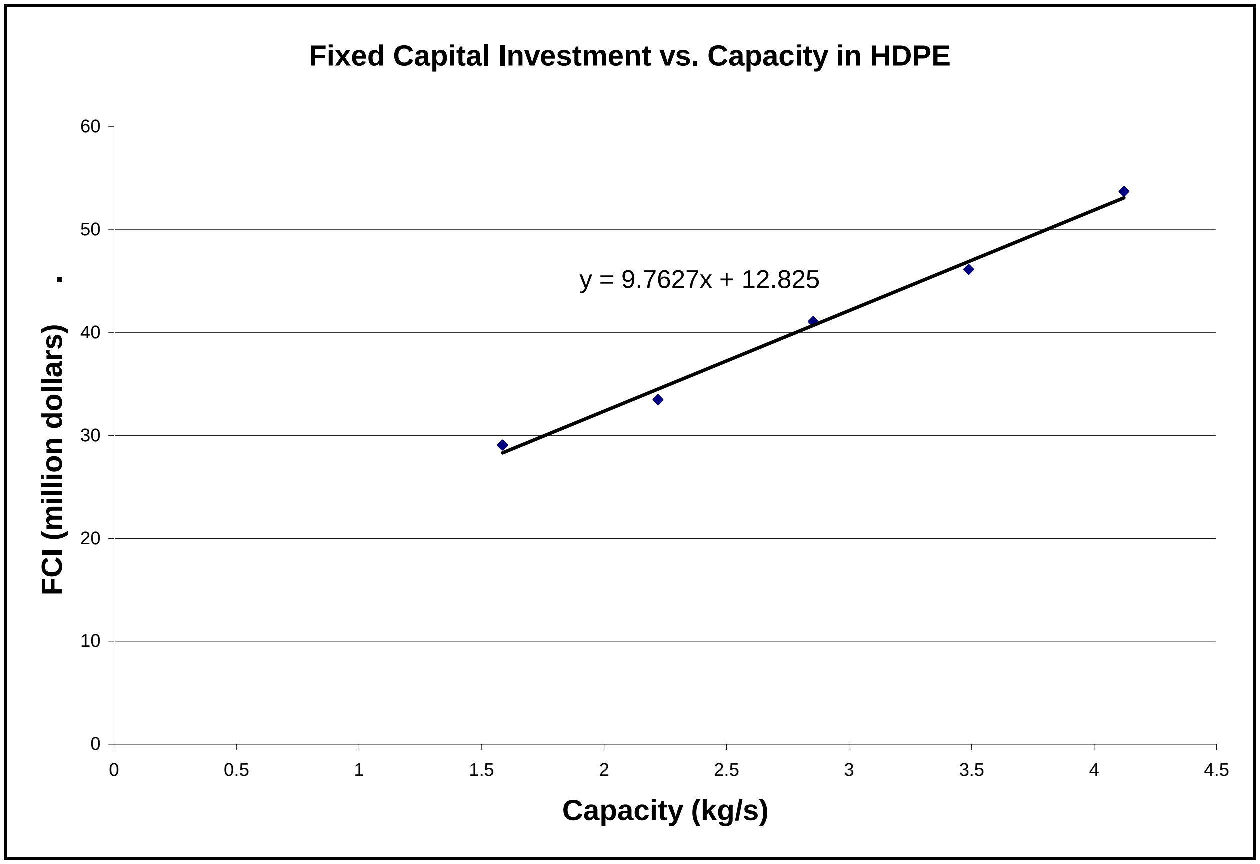


Figure 3.6.2.2 High Density Polyethylene FCI versus Capacity

The high-density polyethylene operating cost was found using the Handbook of Chemicals Production Processes. The operating cost was broken down into 3 categories: raw material, utilities, and labor. Raw material and utilities were given in terms of capacity, while the number of operators, supervisors, and laboratory technicians was estimated. The prices for raw materials were found from the Chemical Market Reporter and Plant Design and Economics for Chemical Engineers. The operating costs ranged from \$5.5 million/yr to \$10.8 million/yr for flow rates ranging from 2 kg/s to 4 kg/s. The trend line is shown in Figure 3.6.2.3.

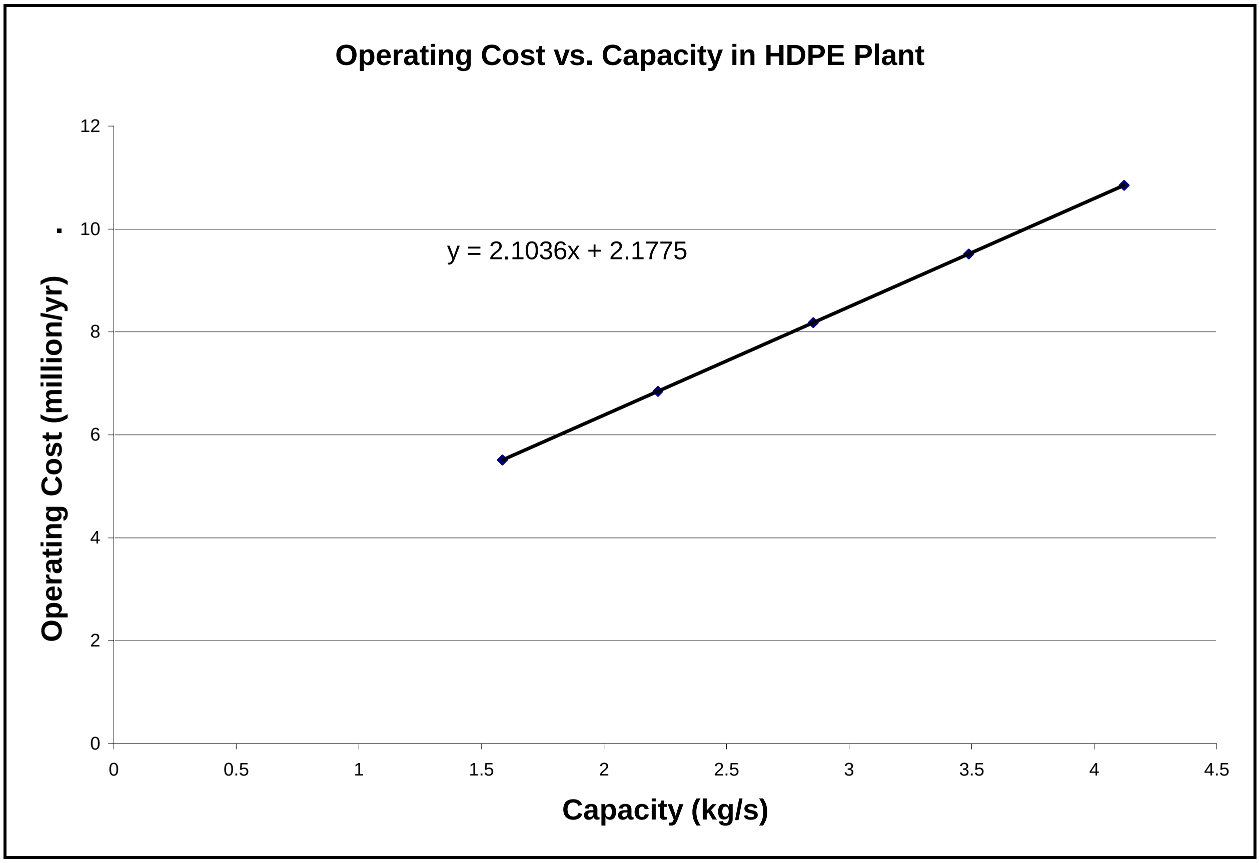


Figure 3.6.2.3 High Density Polyethylene OC versus Capacity

A material balance for the high density polyethylene process is shown in Figure 2.2.1. The material balance and equations found from the FCI and OC graphs were entered into the GAMS program, and used to determine the cost of the process with respect to varying capacities.

3.6.3 Linear-Low Density Polyethylene

Linear-low density polyethylene fixed capital investment may be assumed to equal that of high density polyethylene, as assumed in petrochemical industries. The Borstar PE process is very similar to the Phillips process. The process flow diagram is shown in Figure 3.6.3.1.

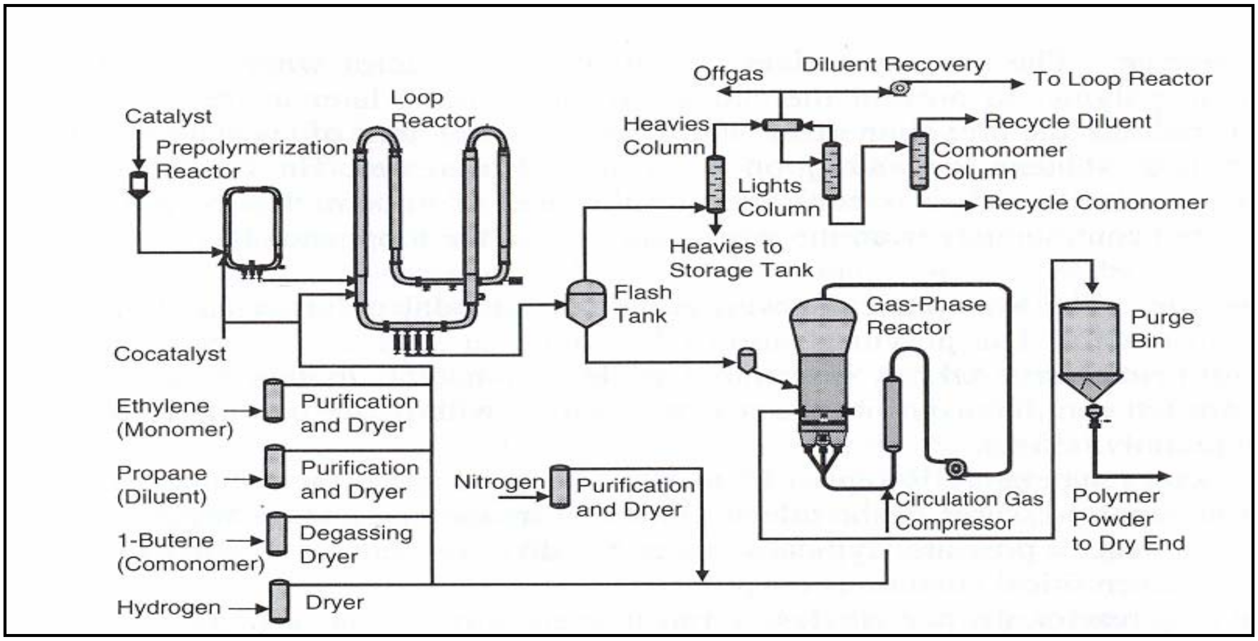


Figure 3.6.3.1 Borstar linear-low density process

This process was used to find the operating cost of the linear-low density polyethylene process. The utility and raw material consumption and number of workers was given in Handbook of Petrochemicals Production Processes. The operating cost ranged from \$37 million to \$92 million for capacities of 1 kg/s to 4 kg/s, respectively. The trend line relating operating cost to capacity is shown in Figure 3.6.3.2.

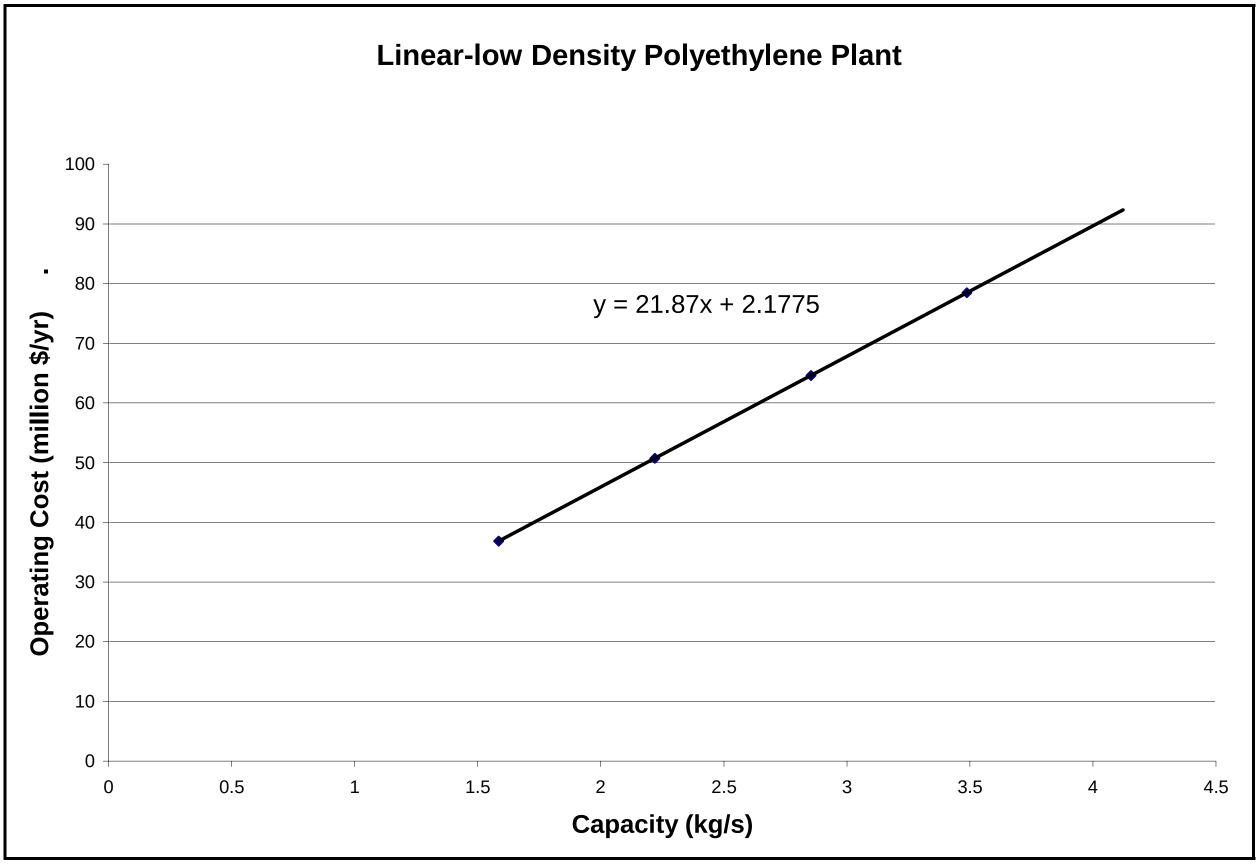


Figure 3.6.3.2 Linear-low density FCI versus Capacity

3.7 Polypropylene

The process economics used to find the operating costs and capital investments comes from the Handbook of Chemicals Production Processes. The two processes considered were the BASF Novolen Polypropylene process and the Phillips Polypropylene process.

The BASF process involves three parts: polymerization, powder discharge, and finally extrusion. A disadvantage of the BASF process is that the catalyst preparation is a crucial and rather complex process.

An advantage of the BASF process is that it uses a specially designed helical stirrer to produce an excellent homogeneity in the powder bed. Space and time yields of this process are also higher than other technologies due to the density of the powder bed. The BASF Novolen process produces all types of polypropylene homopolymers needed for a wide variety of applications, though the Novolen process is most suitable for the production of high impact strength resins.

The information available to Phillips indicates that the Phillips Polypropylene process is the simplest and most efficient for producing polypropylene, homopolymer, and ethylene propylene random copolymers in the world. The Phillips process, therefore, was used.

The components of the inlet of the Phillips Polypropylene process are continuously fed to a pipe loop reactor. A polymer in propylene monomer slurry flows out of the reactor, the catalyst is deactivated and separated using a simple extraction system. This stream is then flashed to remove hydrocarbons from the polypropylene solid particles. The particles are dried and transferred to the extruder to create pellets. The unreacted monomers are recycled after purification.

The trend line relating FCI to capacity is shown in Figure 3.7.1. The fixed capital investment ranges from $54 million to 111 million for flows of 2.2 kg/s to 6.7 k/s, respectively.

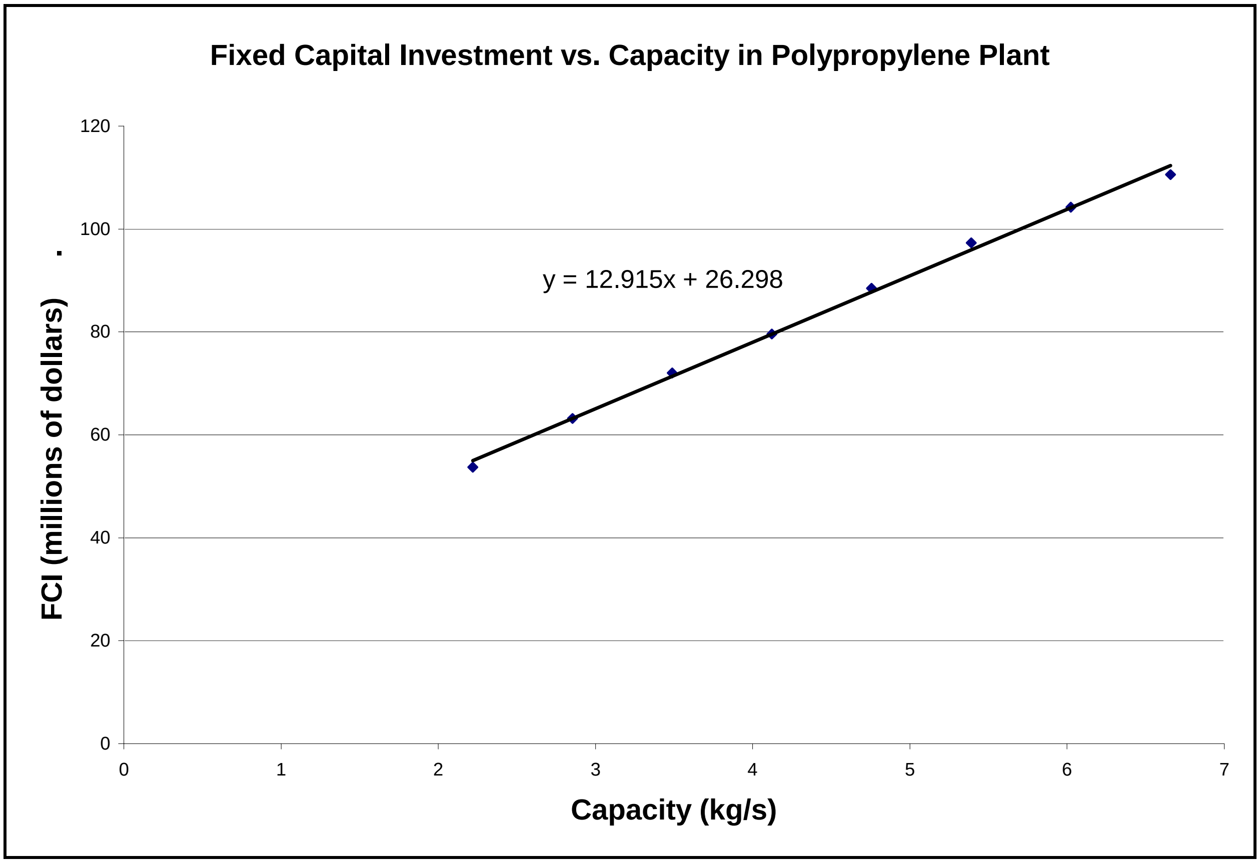


Figure 3.7.1 Polypropylene FCI versus Capacity

The operating cost was broken down into 3 unknown categories: utility cost, labor cost, and raw material cost. The raw materials required to produce polypropylene are propylene, metollocene catalyst, methyl aluminoxane co-catalyst, nitrogen steam, and stabilizers. An in depth analysis of the raw materials needed was given by *Applications for Selective Adsorbents in Polymer Production Processes*. The prices of the raw materials were provided by Chemical & Engineering News, Chemical Market Reporter, and the Great Lakes Chemical Corporation. The operating costs were found to range from \$6 million dollars/yr to \$14 million dollars/yr for flow of 2 kg/s to 7 kg/s. This relationship can be seen in Figure 3.7.2.

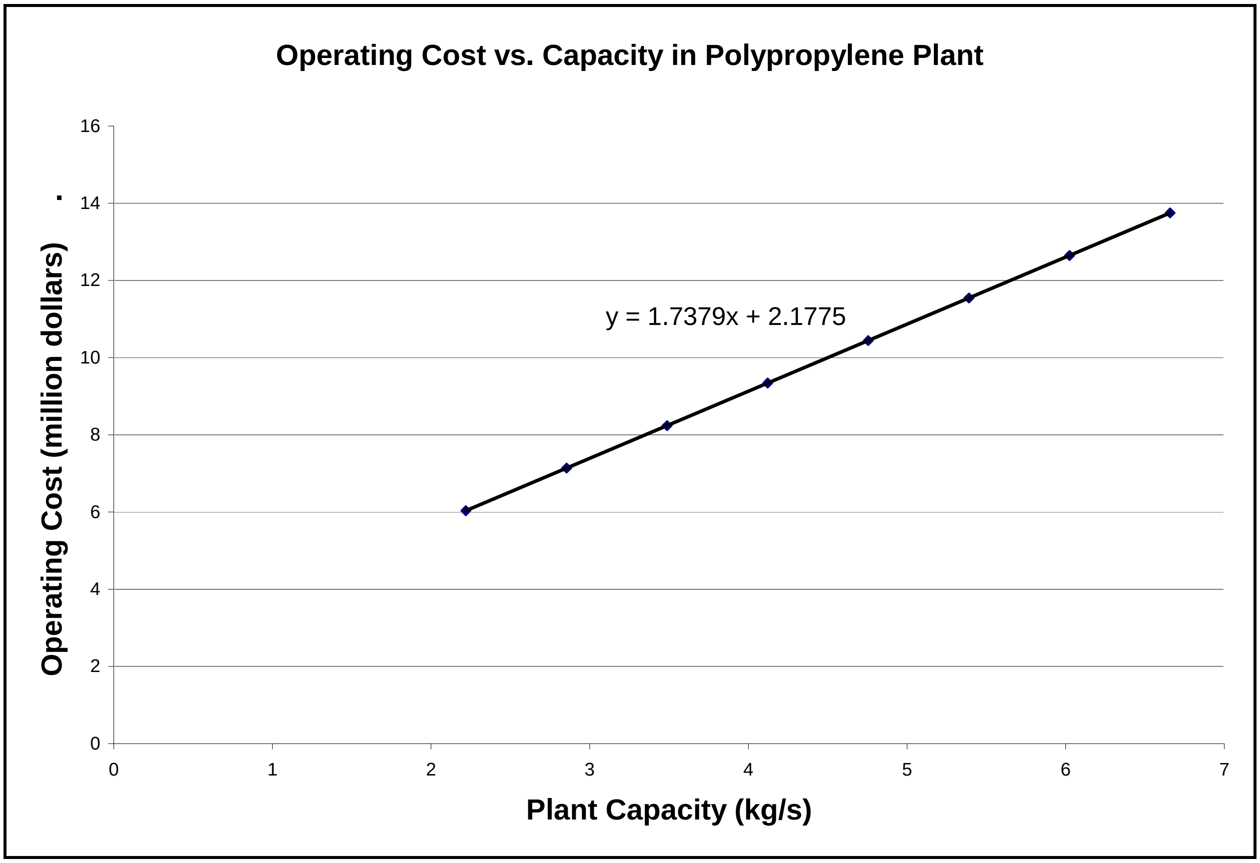


Figure 3.7.2 Polypropylene Operating Cost versus Capacity

A material balance for the polypropylene process is shown in Figure 2.2.1. This information and the equations found are entered into the GAMS program, and are used to determine the cost of the process when different capacities are used.

3.7 Thermal Coupling-Oligomerization

The equipment used for thermal coupling and oligomerization was found from equipment descriptions from Nonoxidative Methane Conversion to Acetylene over Zeolite in a Low Temperature Plasma and Reaction Unit and Product Analysis. The primary apparatuses used to find the equipment cost were two reactors and two furnaces. All of the equipment was assumed to be stainless steel. The equipment cost was found for 10 kg/s, 20 kg/s, and 30 kg/s. The reactor was assumed to have an operating pressure of 50 psia and capacity of 10 m^3, 20m^3, and 30m^3, respectively. The furnace heat duties for each flow were assumed to be 5,000 kW, 10,000 kW, and 15,000 kW, respectively. The total equipment costs were found to be $0.8 million, $1.45 million, and $4.1 million for flows of 10 kg/s, 20 kg/s, and 30 kg/s, respectively. The fixed capital investment found was approximately \$4.4 million, \$5.9 million, and \$22.6 million. The FCI trend may be seen in Figure 3.8.1.

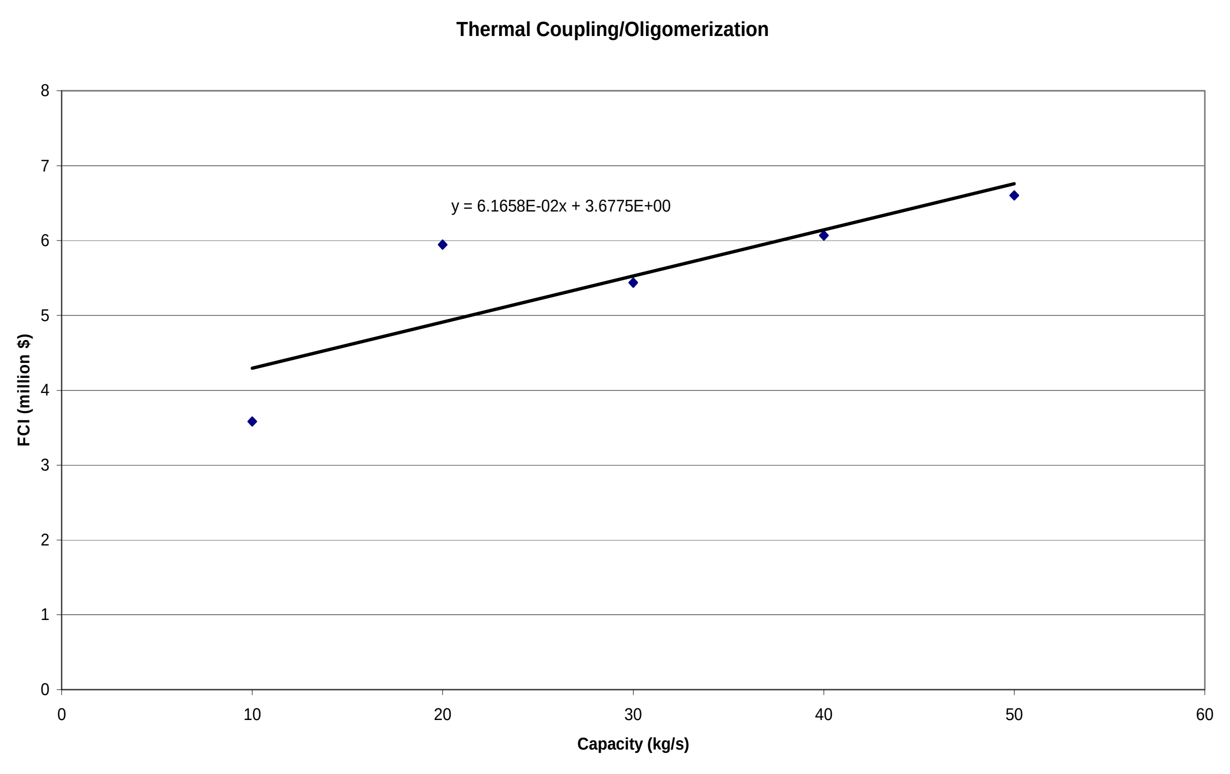


Figure 3.8.1 Thermal Coupling FCI versus Capacity

The unknown parameters of the operating cost are: utilities, labor, and raw material costs. The catalyst cost was found from Synthesis of Catalytic Zeolites. The cost of the zeolite catalyst is \$.76/kg. The mass of the catalyst needed per year was estimated using the weight hourly space velocity of .04 kg cat/mol s given in “Low Temperature Alkane Activation Over Zeolites”.. The number of operators was estimated as 3 people, regardless of process. The utility cost was assumed, as approximated in PT&W to be 10% of the total product cost. Using ratios given by PT&W, the operating costs were found ranging from \$15.8 million per year to \$44.5 million dollars per year for flows ranging from 10 kg/s to 30 kg/s. The trend line may be seen in the Figure 3.8.2.

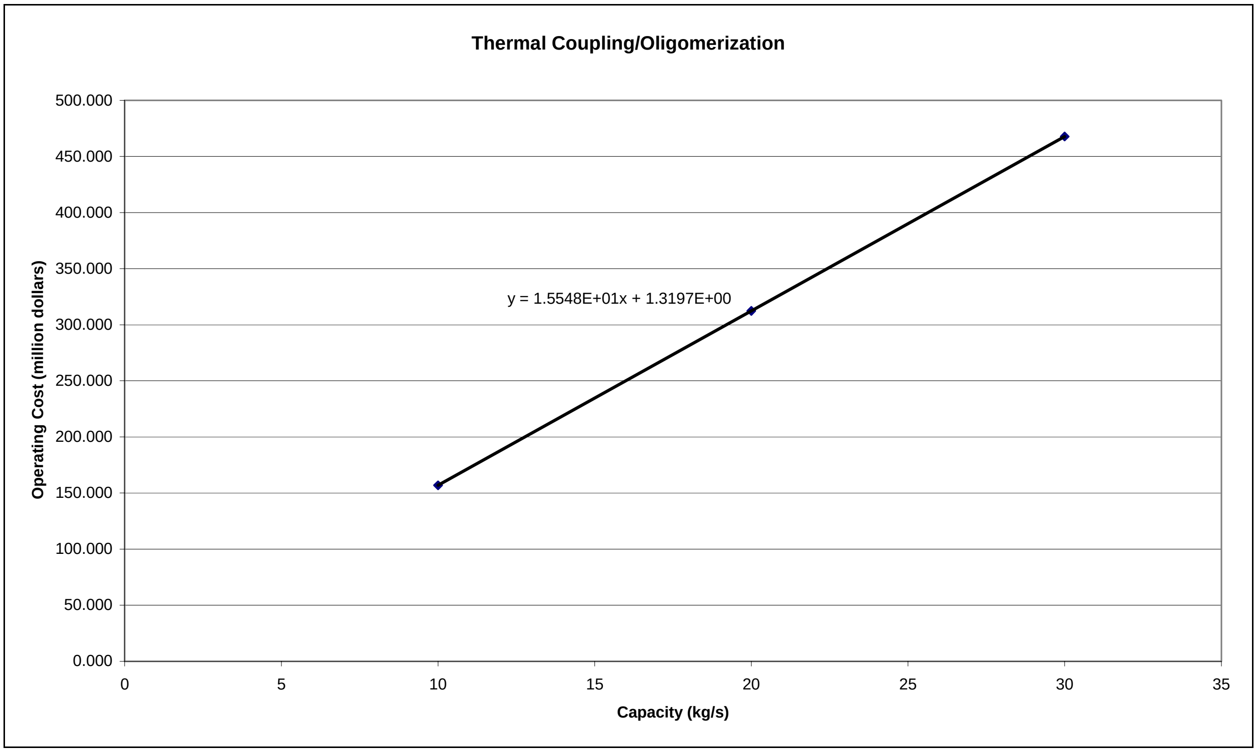


Figure 3.8.2 Thermal Coupling Operating Cost versus Capacity

A material balance for the thermal coupling process is shown in Figure 2.2.1. This information and the equations found are entered into the GAMS program, and are used to determine the cost of the process when different capacities are used.

Although initial economic calculations for FCI and operating cost were completed for this process, this was process was eliminated from consideration due to lack of literature. It is recommended that when industrial scale data for this process is available, that the thermal coupling and oligomerization processes be added to infrastructure of the entire process flow diagram and be considered as a possible process for the natural gas from the Camisea reservoir.

3.9 Methanol Synthesis

There are a few key technologies used in methanol synthesis. Low Pressure Methanol (LPM) technology uses low pressure reformer. LPM is being used to produce 60% of the methanol of the world. Gas Heated Reformer (GHR) technology enables the manufacture of greater volumes of methanol, thus reducing the cost of production relative to the conventional LPM. Leading

Concept Methanol (LCM) technology brings together LPM and GHR. It has been proved to be one of the most economical choices for large-scale methanol manufacturing.

For the purpose of calculating the FCI and operating cost for the methanol synthesis process, the flow diagram of a methanol plant using LCM technology (Figure 3.9.1) was used (Coogee). Equipments were sized by performing mass and energy balances on the flow diagram. As shown in Figure 3.9.1, natural gas entering the plant at 15 bar is compressed to 45 bar and purified. Methane leaving the purifier is mixed with steam in the saturator. The steam and methane mixture is fed to the gas heated reformer where the synthesis gas, a mixture of CO, CO_2, and H_2, is generated. The synthesis gas leaving the gas heated reformer is fed to the methanol converter where methanol is produced.

Table 3.9.1 shows the equipment cost break down and the total equipment cost for a capacity of 10 kg/s. Figure 3.9.2 shows the relationship between the FCI calculated and the capacity. The operating costs were calculated using the PT&W online spreadsheet. The amount of utilities and raw material used were determined by performing energy and material balances on the system. A graph showing the relationship of operating cost versus capacity is shown in Figure 3.9.3.

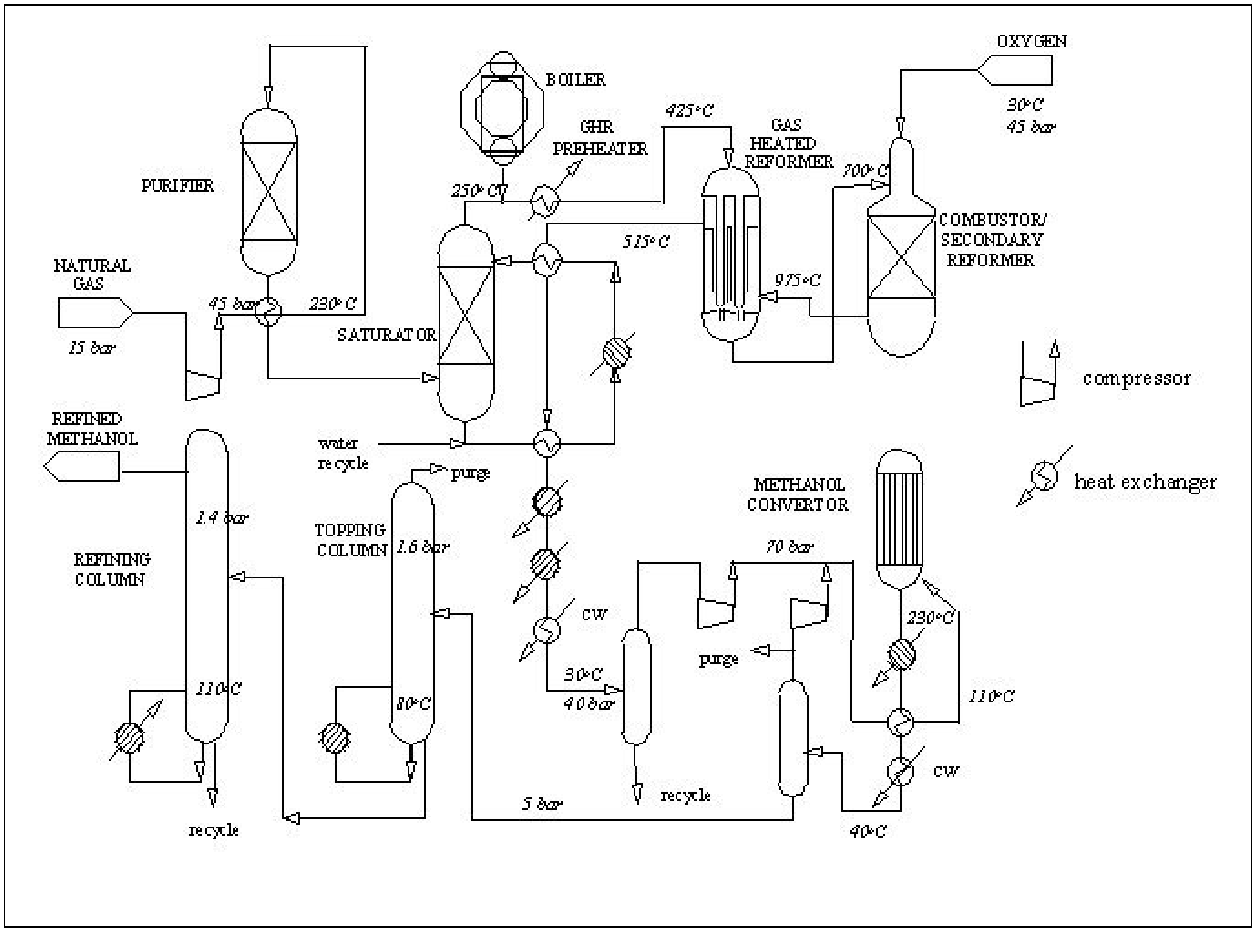


Figure 3.9.1 Methanol Synthesis Process Flow Diagram

Table 3.9.1 Equipment Cost of a methanol plant with a capacity of 10 kg/s

2 Steam Reformers	10,235,000
3 Compressors	1,241,400
Saturator	450,200
Combustor	258,000
Methanol Converter	412,000
Two flashes	464,800
Two columns	248,000
7 Heat Exchangers	490,000
Total Equipment Cost	$13,799,400

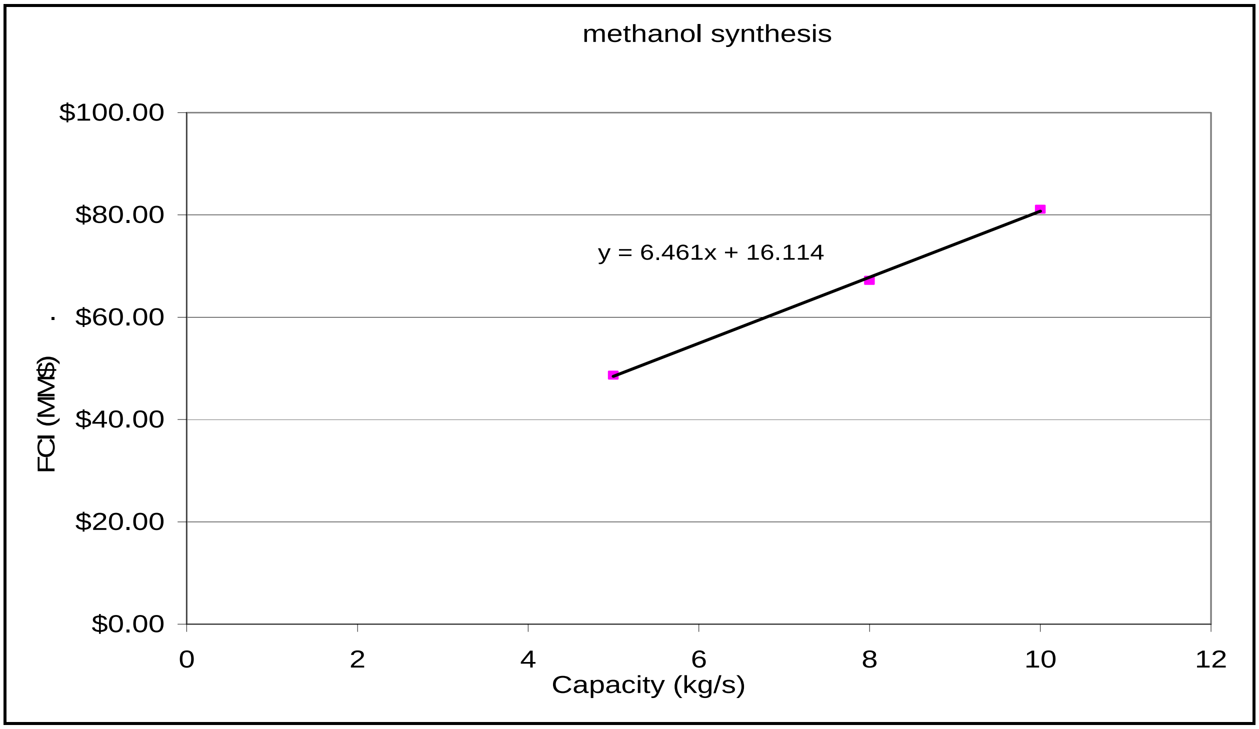


Figure 3.9.2 Methanol Synthesis FCI versus Capacity

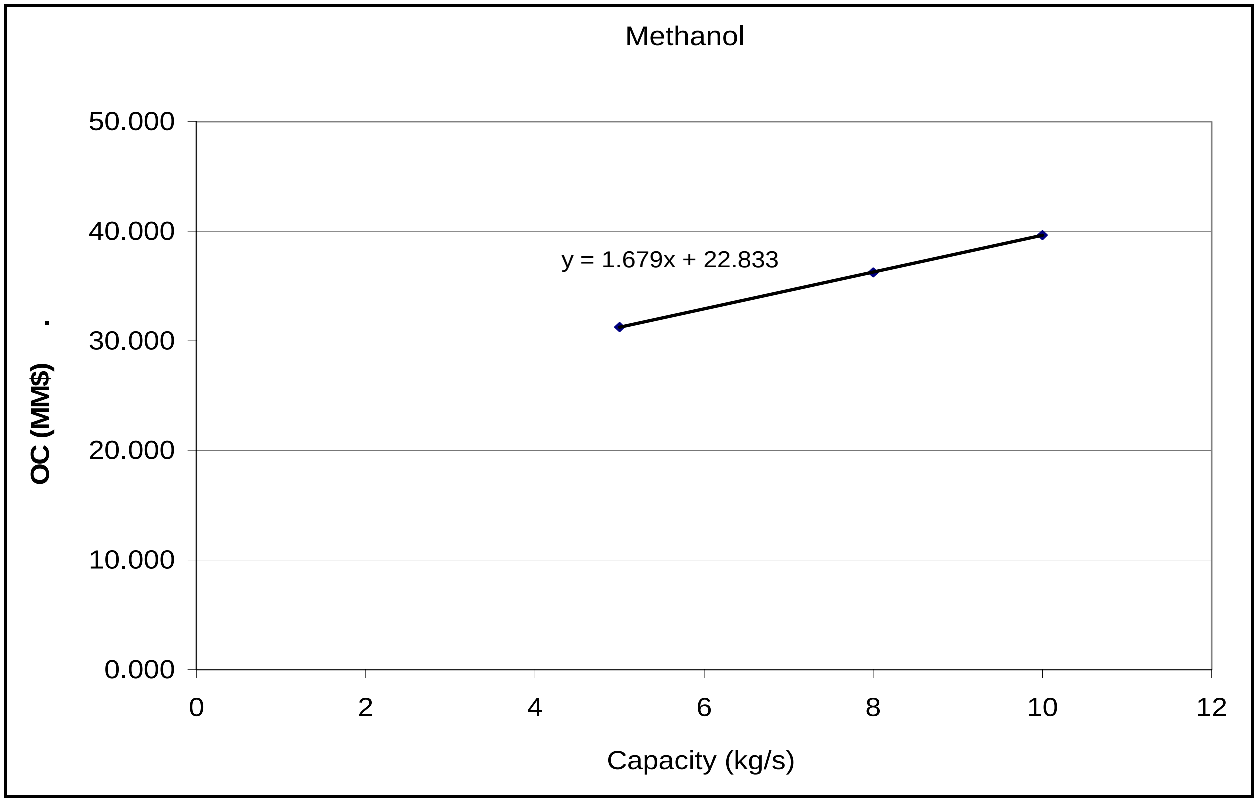


Figure 3.9.3 Methanol Synthesis Operating Cost versus Capacity

A material balance for the methanol synthesis is shown in Figure 2.2.1. This information and the equations found are entered into the GAMS program, and are used to determine the cost of the process when different capacities are used.

3.10 Ammonia Synthesis

Ammonia can be formed from nitrogen and hydrogen under the following conditions:

- Medium temperature (~500^oC)
- Very high pressure (~250 atmospheres, ~351kPa)
- Catalyst present

The following equation shows the reaction:

$$N_2(g) + 3H_2(g) \rightarrow 2NH_3(g)$$

There are four available technologies in the market that are producing ammonia:

1. ICI process
2. Kellogg Brown & Roots Advanced Ammonia plus Process (KAAPplus^{TM})
3. Haldor-Topsoe process
4. Uhde ammonia process

These technologies are almost similar equipment, but different catalysts are used. ICI process is licensed by Johnson-Matthey Catalyst company, and uses 7 different types of Johnson Matthey catalyst for the whole process. Haldor-Topsoe process uses 2 types of Topsoe catalyst. All these catalysts consist of 90% iron, while KAAPplus^{TM} uses ruthenium.

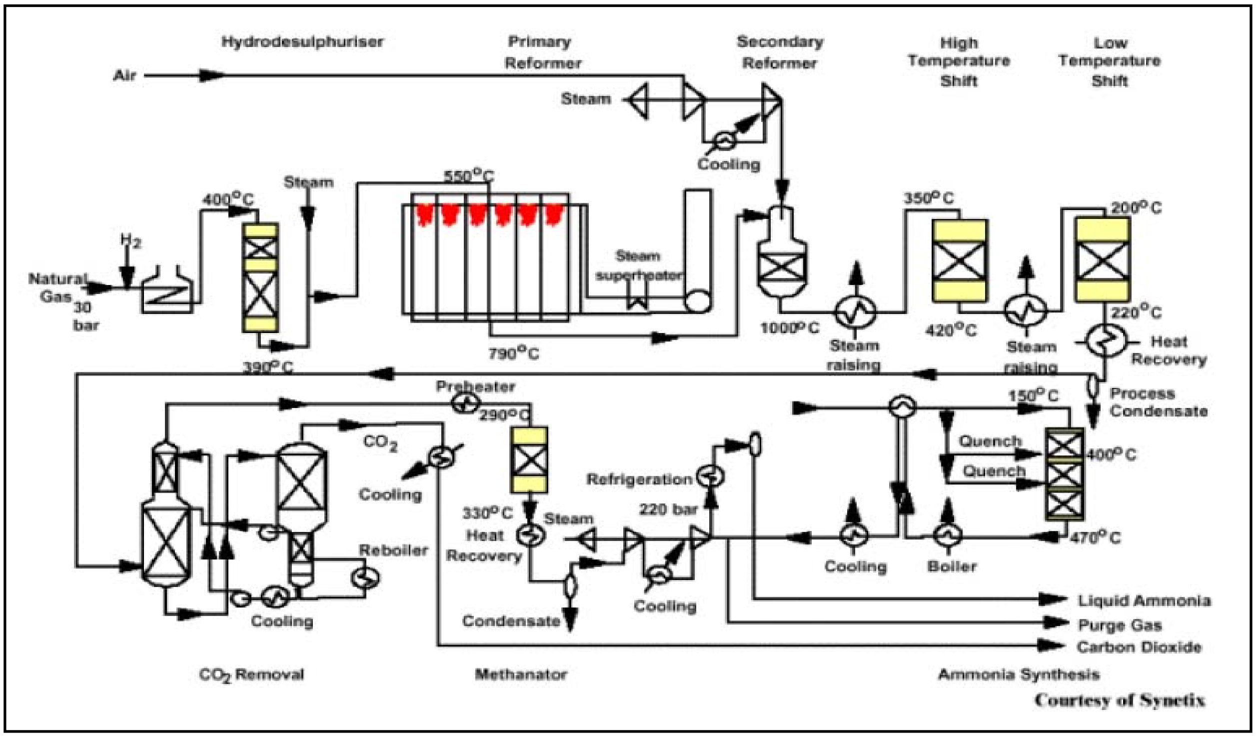
KAAPplus^{TM} is the combination of KAAP^{TM} (Kellogg Brown & Roots Advance Ammonia Process), KRES^{TM} (Kellogg Reforming Exchanger System), and Purifier^{TM} (Kellogg Purifier system). KAAPplus^{TM} is chosen to produce ammonia from methane. This process is chosen because of lower capital cost, improved reliability, reduced operating cost, and lower energy consumption.

KAAPplus^{TM} saves 10-20% of the capital cost compared to the other ammonia plants. It is more reliable because ruthenium is 20 times more active and 40% more effective than iron. The uses of ruthenium can reduce the operating pressure for 90 bar, and thus this can reduce operating cost dramatically. Besides, this process saves 0.1 Gcal to 0.2 Gcal per ton of ammonia produced.

KRES™ is used to substitute direct fired primary reformer. Direct fired primary reformer is eliminated because it is energy-intensive, high capital cost, and high maintenance operation. KRES has following benefits if compared to direct fired primary reformer:

- Lower capital cost
- Improved energy efficiency
- Reduced emissions (NO_X, CO_2)
- Reduced operator demands
- Easier maintenance
- Less plot space

Purifier™ removes the excess nitrogen, all methane, and most of the argon from the syngas. This provides a clean makeup gas to the synthesis loop. Therefore, the required synloop purge is very small, and the synloop pressure can be reduced. The synloop purge is recycled to the Purifier™, so a separate purge gas recovery unit is not needed.



Courtesy of Synetix

Figure 3.10.1 Classical plant for ammonia production

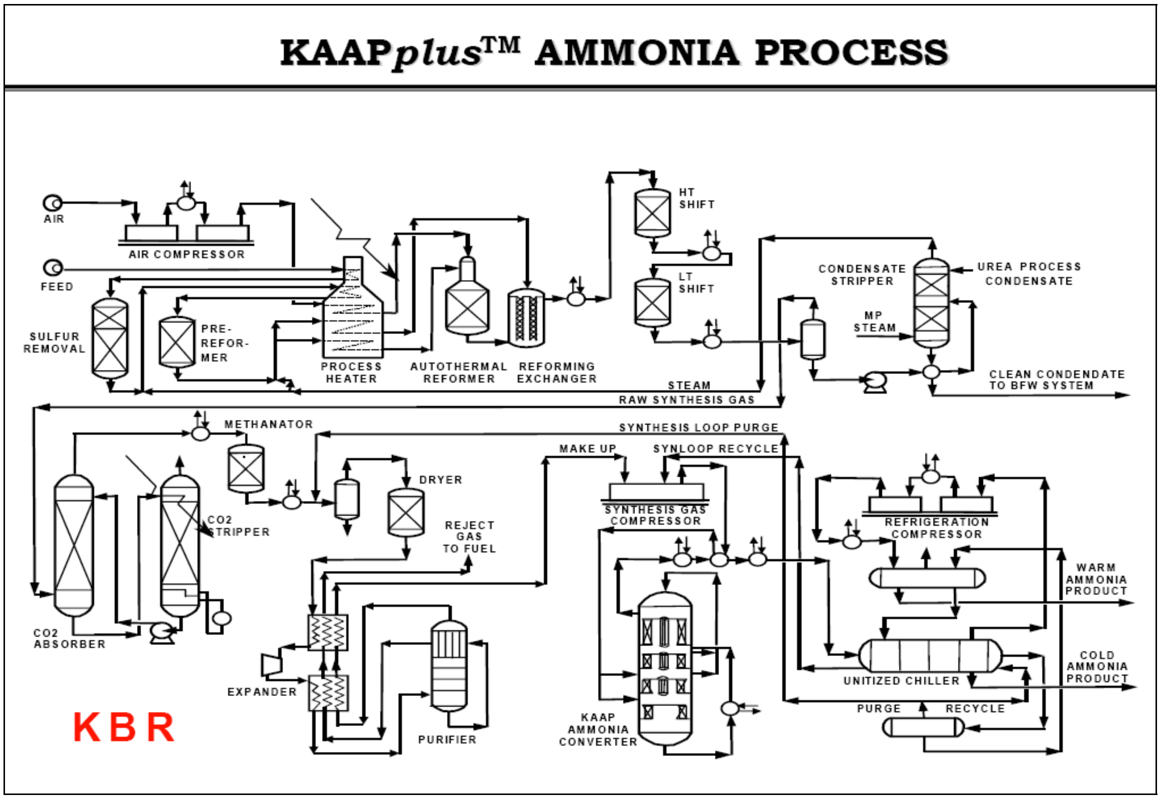


Figure 3.10.2 KAAPplus^{TM} ammonia process

The FCI for both processes are similar. The classical process, shown in Figure 3.10.1, has a higher equipment cost but lower license fees. However, KAAPplus^{TM} has a lower equipment cost but higher license fees due to this process being introduced to the market in 1999.

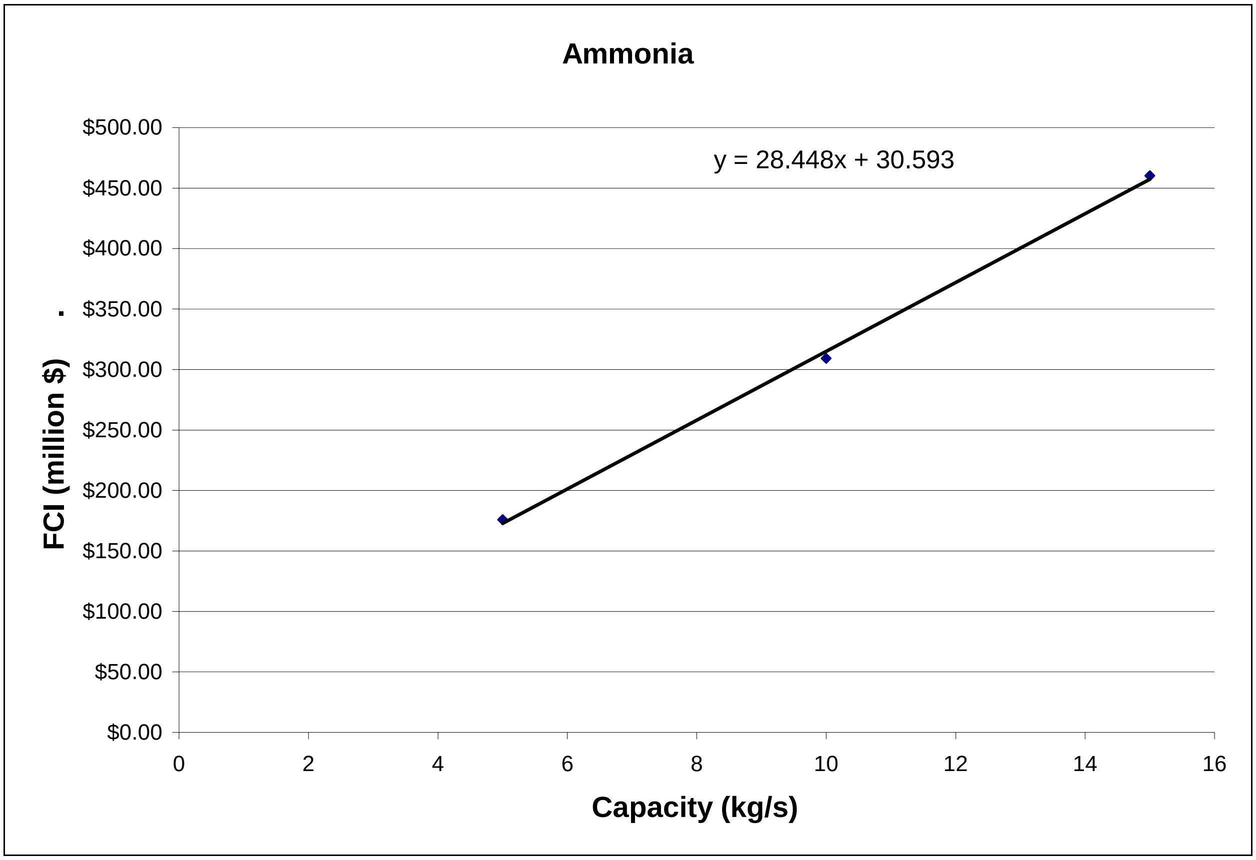


Figure 3.10.3 Ammonia synthesis-FCI

The operating cost of KAAPplus^{TM} is much lower than classical ammonia plant. However, due to it is a new process, no data is available. So, the classical plant is used to calculate the operating cost based on the energy balance, and temperature.

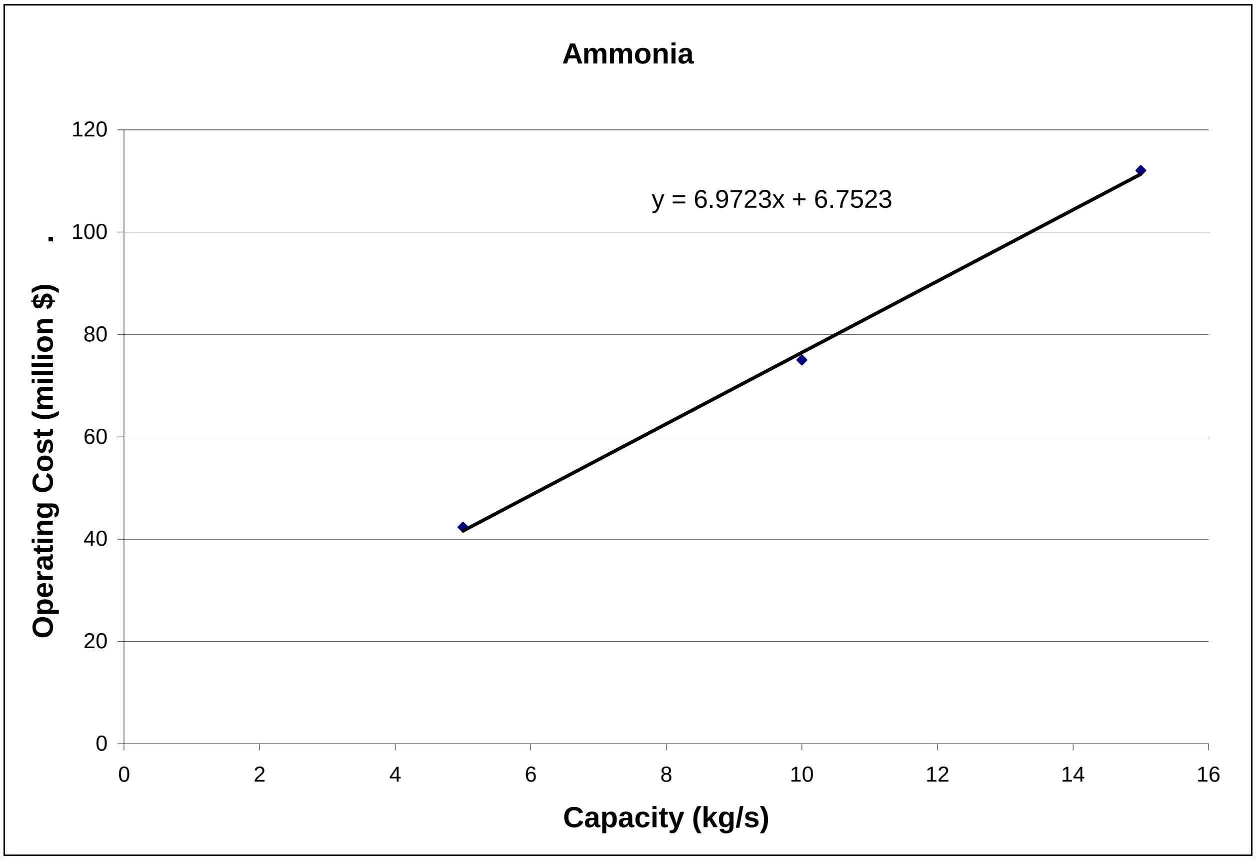


Figure 3.10.4 Ammonia synthesis – Operating cost

3.11 Ammonium Nitrate Fertilizer

Fertilizer produced in this plant in ammonium nitrate (NH_4NO_3). Ammonium nitrate is used widely in the industry of fertilization and explosion. It is also used to modify the detonation rate of other explosives, such as nitroglycerin in the so-called ammonia dynamites, or as an oxidizing agent in the ammonals.

Ammonium nitrate can be produced from ammonia by first making nitric acid and then neutralizing the nitric acid with anhydrous ammonia, creating ammonium nitrate liquor. Thus, this process can be divided into 2 processes, which is shown:

1. Nitric acid formation:

$$NH_3(g) + 2O_2(g) \leftrightarrow HNO_3(aq) + H_2O(l)$$

2. Ammonium nitrate fertilizer:

$$HNO_3(aq) + NH_3(g) \leftrightarrow NH_4NO_3(aq)$$

Major licensors for nitric acid plant are listed as following:

- Krupp Uhde
- BAMAG

BAMAG is eliminated due to its operating condition. It operates at 1500kPa and 950^{o}C. There are three Uhde processes available which are medium-pressure process, dual-pressure process, and high-pressure process. Figure 3.11.1 (Uhde: Nitric Acid) shows the comparison of these three processes:

Plant type	Medium-pressure process	High-pressure process	Dual-pressure process
Operating pressure p_{abs}	5.8 bar	10.0 bar	4.6/12.0 bar
Ammonia	284.0 kg	286.0 kg	282.0 kg
Electric power	9.0 kWh	13.0 kWh	8.5 kWh
Platinum, primary losses	0.15 g	0.26 g	0.13 g
with recovery	0.04 g	0.08 g	0.03 g
Cooling water ($\Delta t = 10$ K) including water for steam turbine condenser	100 t	130 t	105 t
Process water	0.3 t	0.3 t	0.3 t

Figure 3.11.1 Consumption for one ton of nitric acid produced

Based on Figure 3.11.1, dual pressure is chosen due to milder operating temperatures, lower lost of catalyst (platinum), lower usage of electrical power. Figure 3.11.2 shows the flow diagram of Uhde dual pressure nitric acid plant:

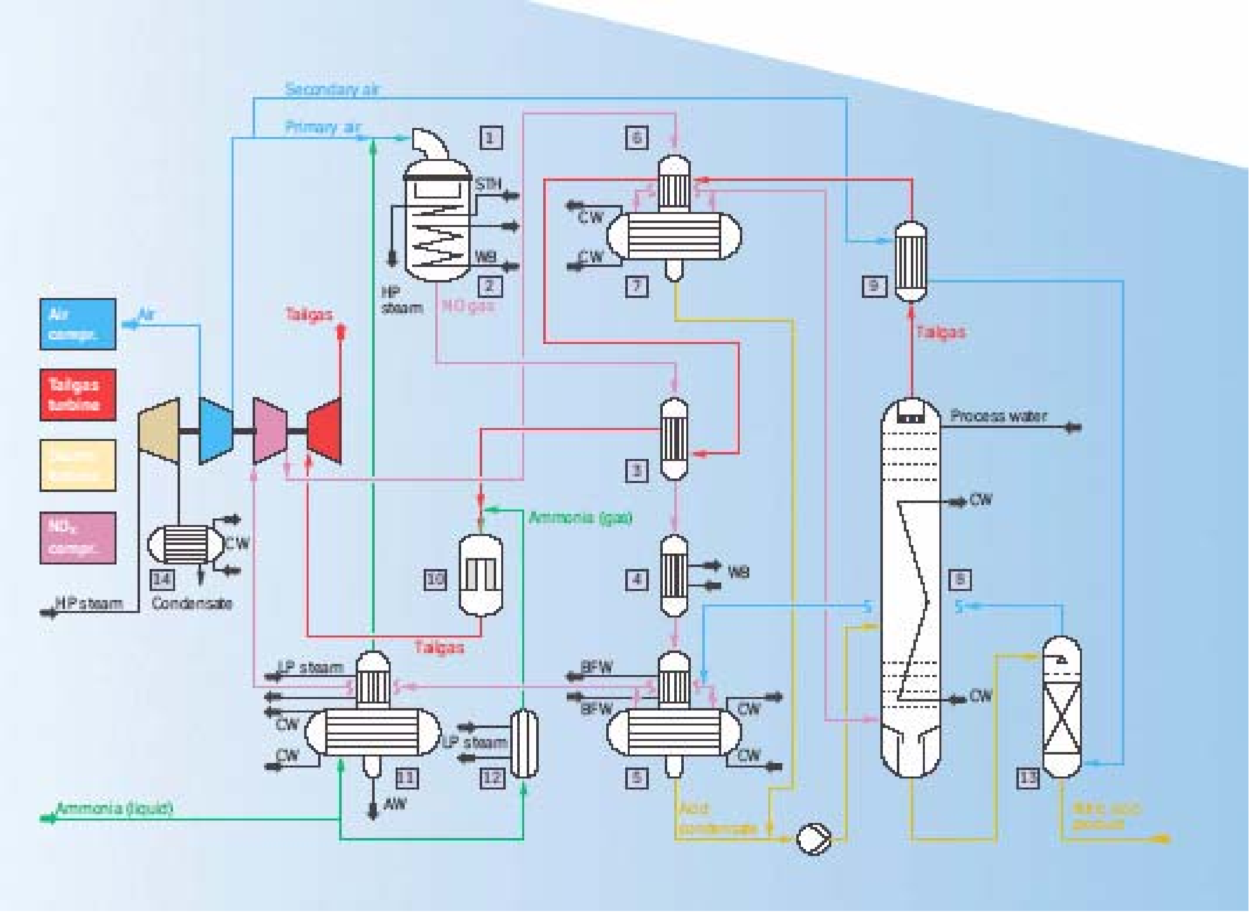


Figure 3.11.2 Uhde dual pressure nitric acid plant

After the production of nitric acid, it is sent to the ammonium nitrate process, where it reacts with ammonia to produce ammonium nitrate. However, 60% of the market uses solid ammonium nitrate and therefore, most of the ammonium nitrate solution needs to be concentrated and solidified. An additive, magnesium nitrate or magnesium oxide, may be injected directly into the melt stream. The purpose of the additive is to raise the crystalline transition temperature of the final solid product; to act as a desiccant, drawing water into the final product to reduce caking; and to allow solidification to occur at a low temperature by reducing the freezing point of molten ammonium nitrate. Figure 3.11.3 shows the flow chart of the ammonium nitrate production from ammonia and nitric acid:

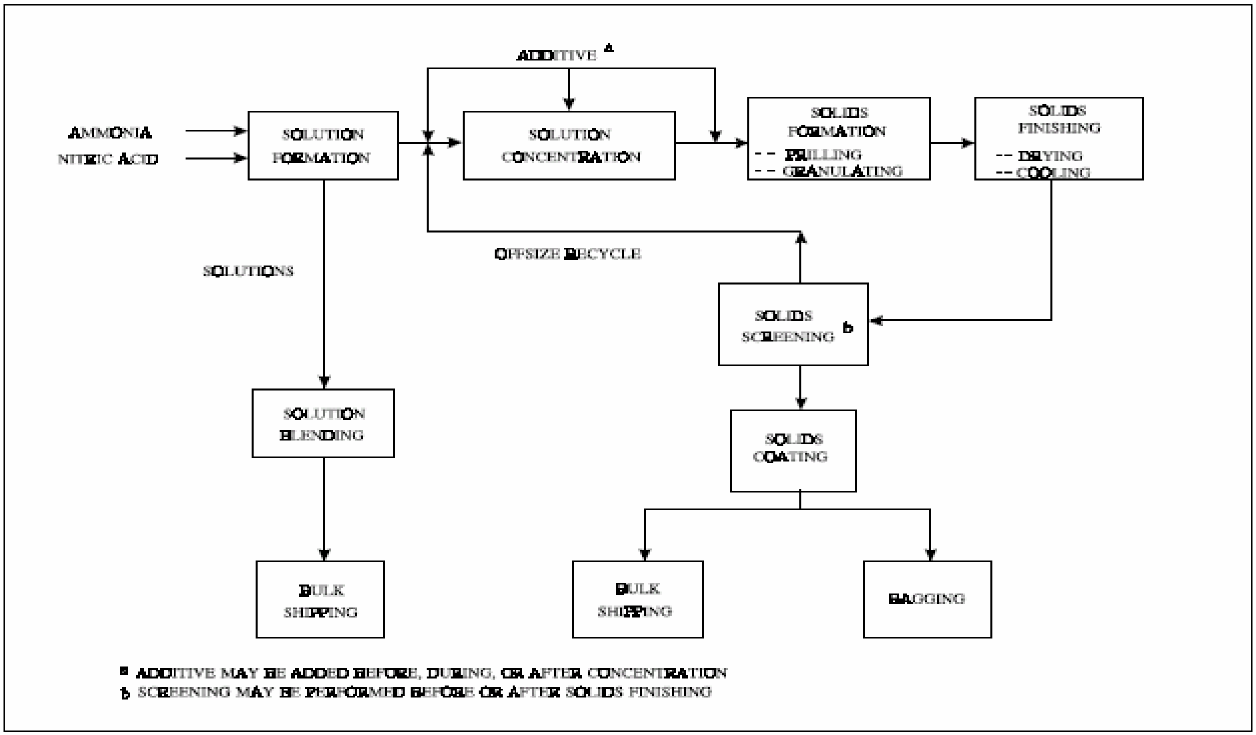


Figure 3.11.3 Ammonium nitrate manufacturing operations

The FCI and operating cost at different capacities were calculated using the PT&W tables with some modifications. The FCI and operating cost of this plant is the summation of the FCI and operating cost of nitric acid formation plant and ammonium nitrate manufacturing plant.

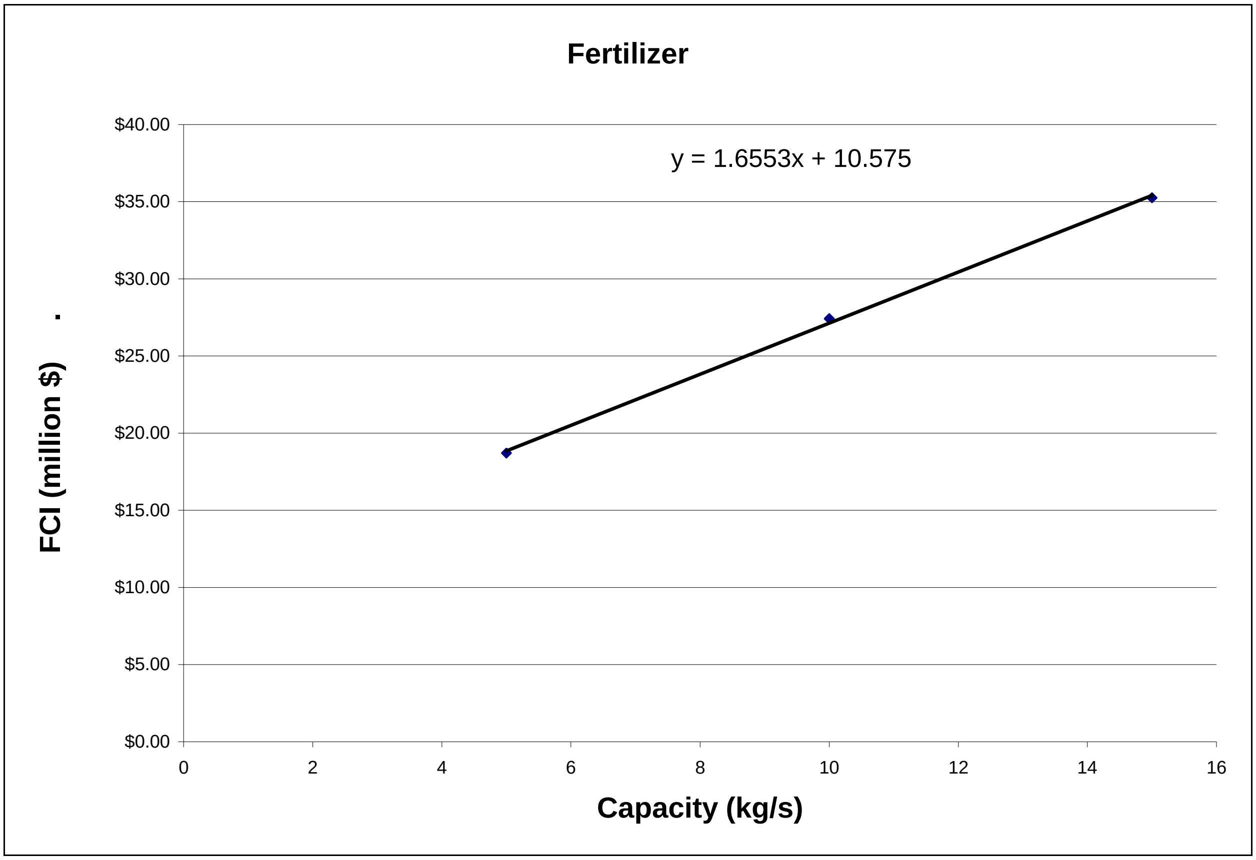


Figure 3.11.4 Fertilizer-FCI

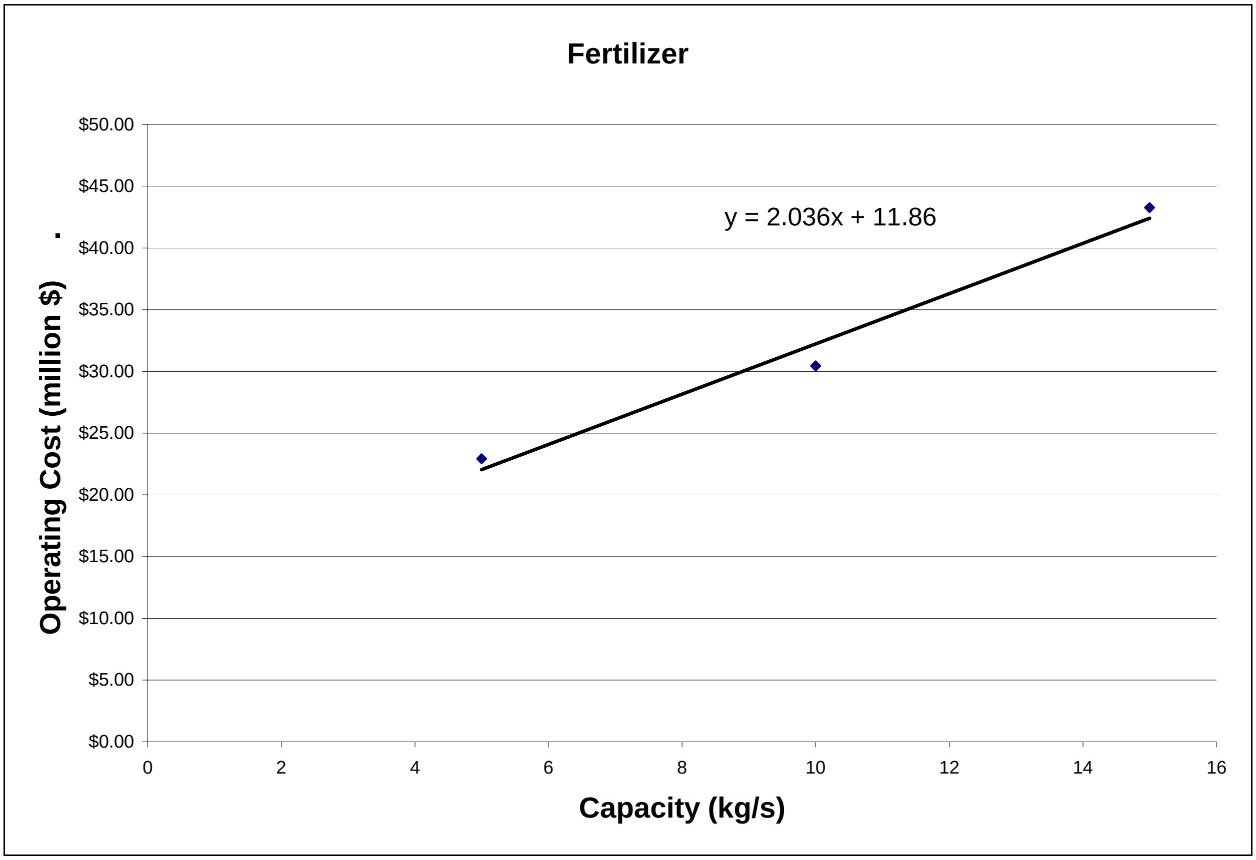


Figure 3.11.5 Fertilizer-Operating cost

3.12 Fischer-Tropsch Process

The Fischer-Tropsch (F-T) process converts natural gas into a wide range of linear hydrocarbons. Fischer-Tropsch provides alternative routes for the production of transportation fuels and petrochemical feedstock. Fischer-Tropsch requires a large capital investment.

Figure 3.12.1 shows a Sasol Fischer Tropsch plant. As shown in the flow diagram, natural gas and oxygen from the air separation unit enter the autothermal reformer where the synthesis gas, a mixture of CO and H_2, is generated. The synthesis gas enters the F-T synthesis unit where a wide range of hydrocarbons, which is called synthetic crude oil, is generated. The synthetic crude oil then enters the product upgrading system where major products (fuel gas, naphtha, and diesel) are produced. Also, the process has a small hydrogen plant to produce hydrogen for internal use and a steam turbine to produce electricity.

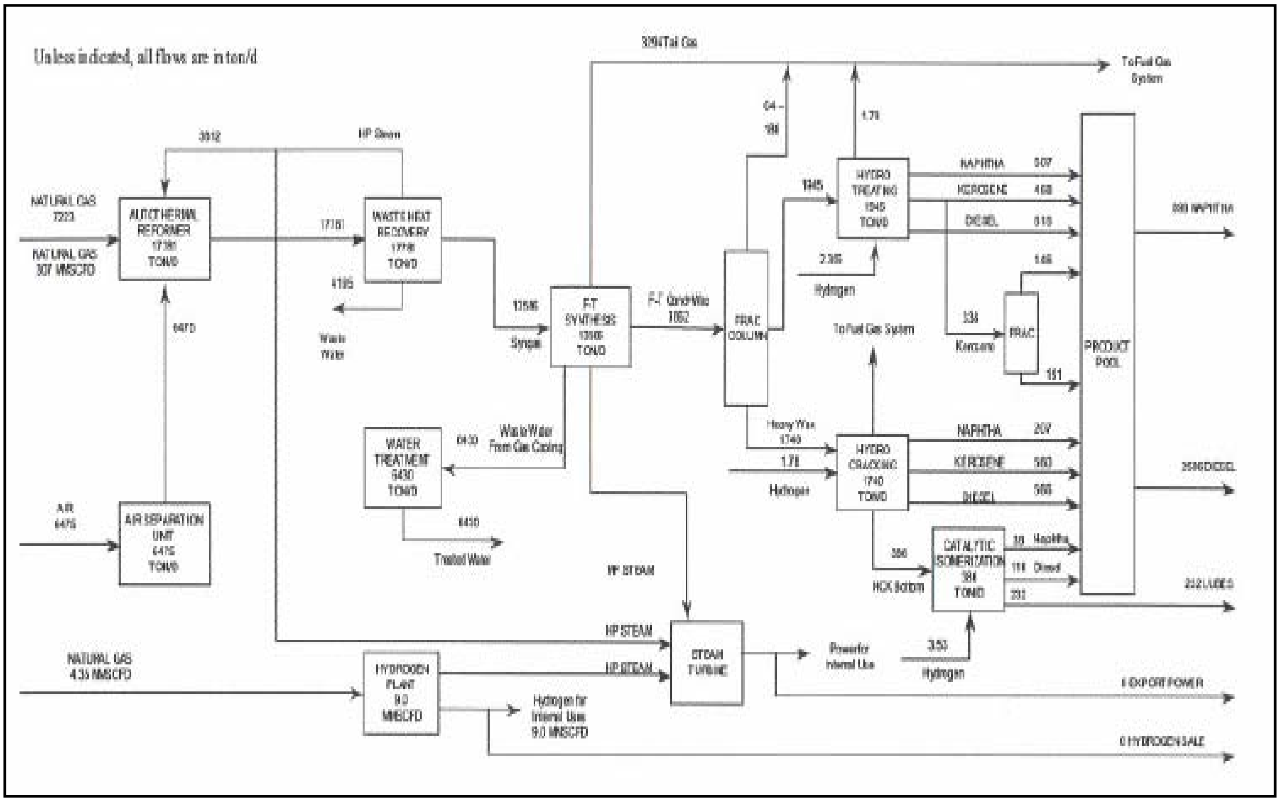


Figure 3.12.1 Fischer-Tropsch Process Flow Diagram

There are a few key technologies used for Fischer-Tropsch (F-T). This technology uses Haldor Topsoe syngas reformer, and a cobalt catalyst F-T slurry reactor. Advanced Gas Conversion (AGC-21) was developed by ExxonMobil. AGC-21 uses a circulating fluidized bed reactor for synthesis gas generation and a slurry cobalt catalytic reactor. Shell developed the Shell Middle Distillate Synthesis (SMDS) technology in 1970s. SMDS uses a non-catalytic combined reforming process and a fixed-be arge-typed reactor.

For the purpose of calculating FCI and operating cost, material and energy balances were performed on the Sasol Fischer Tropsch flow diagram. The synthesis gas generation unit, which includes the autothermal reforming unit and the air separation unit, accounts for roughly 60% of the total cost of the plant. This process also requires a cryogenic air separation unit. A plot of FCI versus capacity was shown in figure 3.12.2. The operating costs were calculated using the PT&W online spreadsheet. The amount of utilities and raw material used were determined by performing energy and material balance on the system. A graph showing the relationship of operating cost versus capacity is shown in Figure 3.12.3.

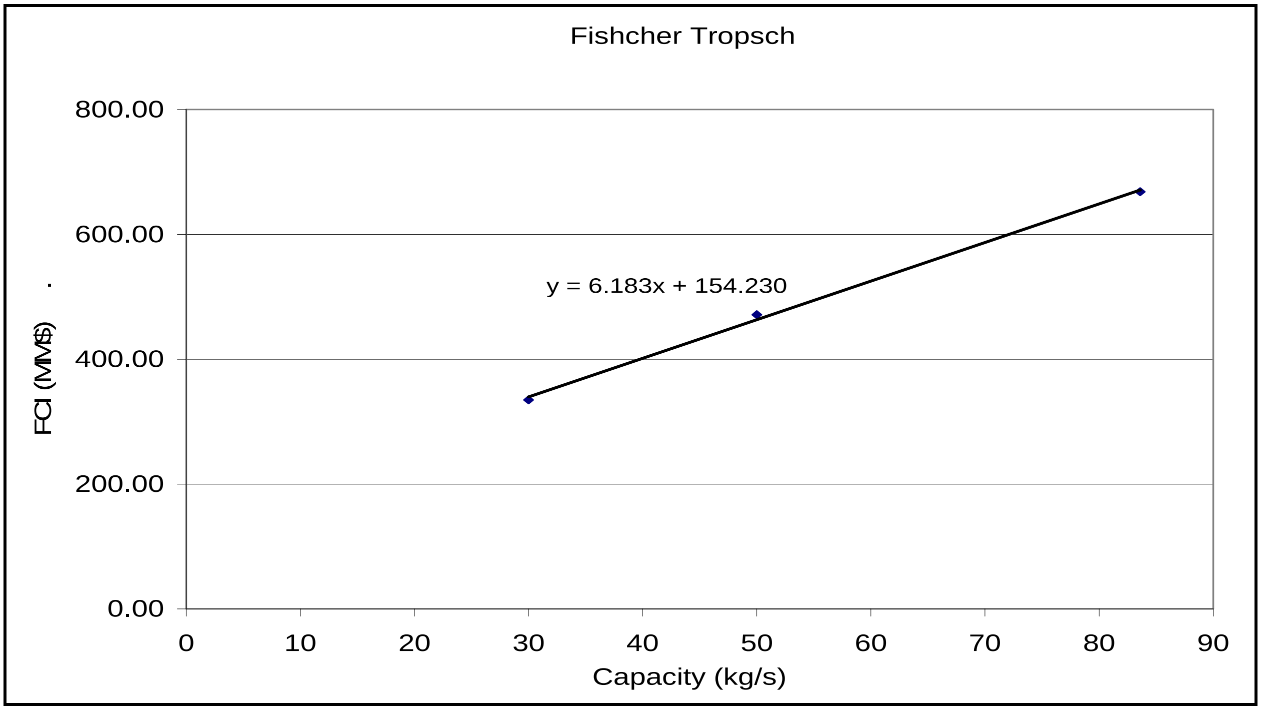


Figure 3.12.2 Fischer-Tropsch Synthesis FCI versus Capacity

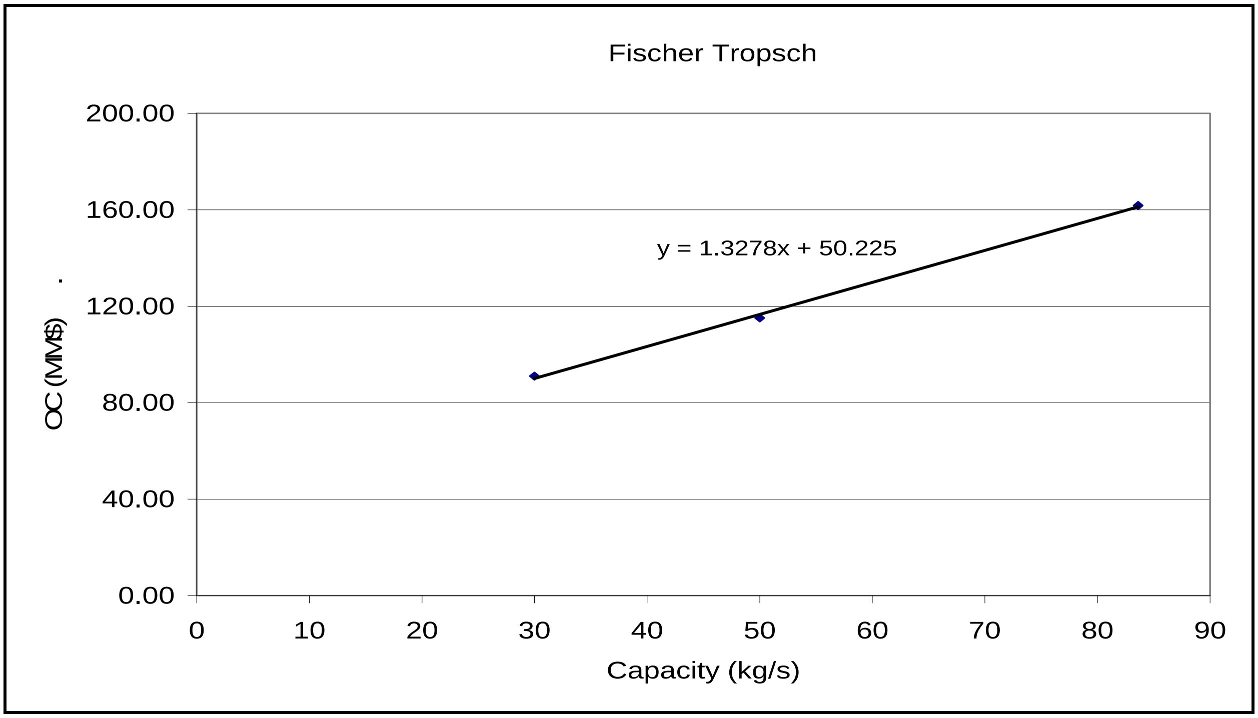


Figure 3.12.3 Fischer-Tropsch Synthesis Operating Cost versus Capacity

A material balance for the Fischer-Tropsch synthesis is shown in Figure 2.2.1. This information and the equations found are entered into the GAMS program, and are used to determine the cost of the process when different capacities are used.

3.13 Liquid Natural Gas Process

Liquid natural gas (LNG) is a process that converts natural gas to a liquid form in order to allow for shipping in tankers or trucks. For this specific region, the liquefied natural gas is sent to an LNG terminal in the western coast city of Pisco, Peru. From Pisco, the liquefied natural gas is exported to the western United States, specifically California.

The process starts when the natural gas from Camisea is condensed and contaminants are removed from the gas. The gas is dried to reduce the water content in order to prevent freezing from occurring during the liquefaction process. When the gas is liquefied, 600 cubic feet of natural gas is converted to one cubic foot of LNG. Figure 3.13.1 shows an illustration of the LNG process, storage, and tanker transportation. Figure 3.13.2 shows the process flow diagram used to analyze the actual liquefaction plant.

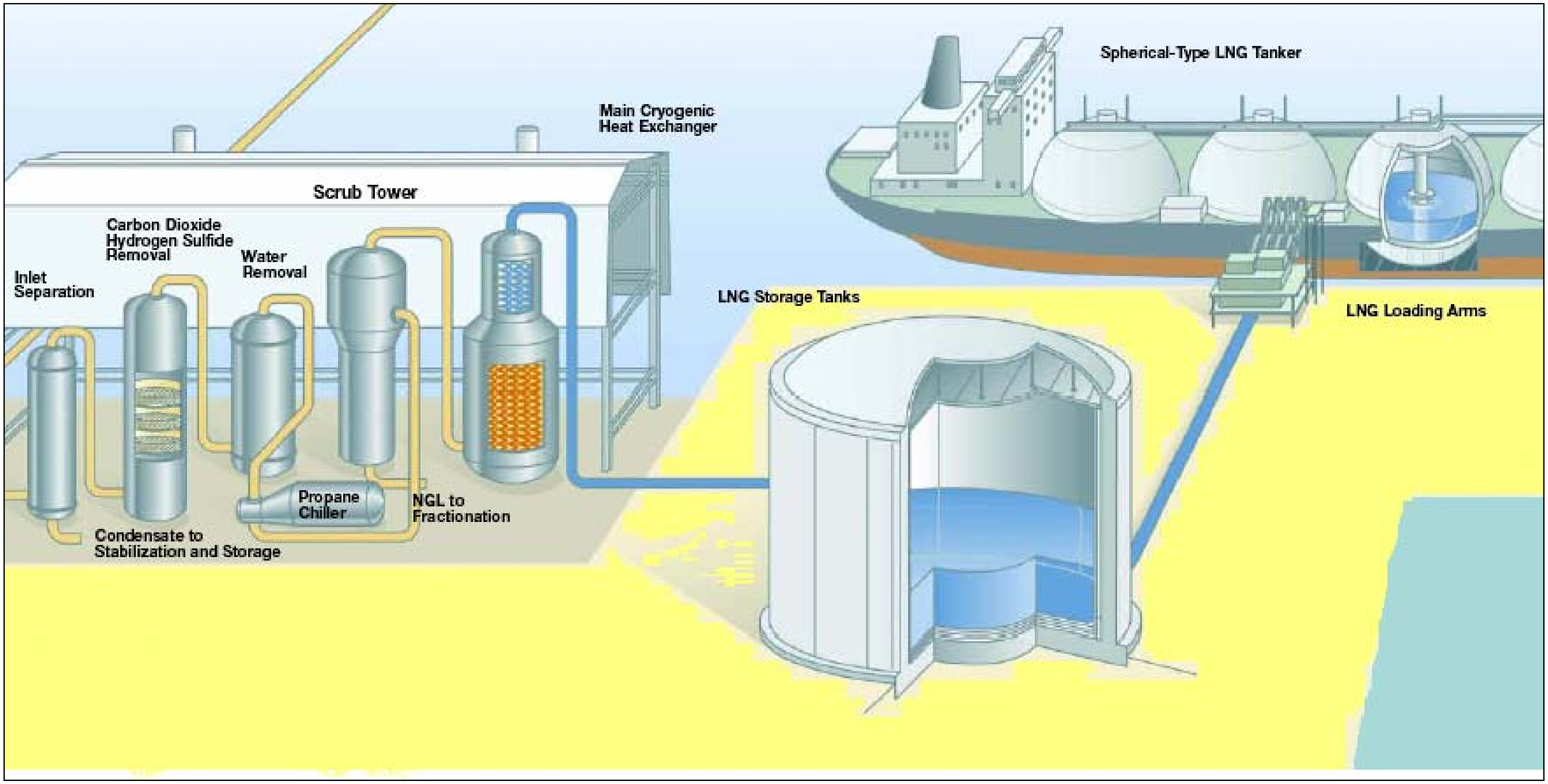


Figure 3.13.1 LNG Process Diagram

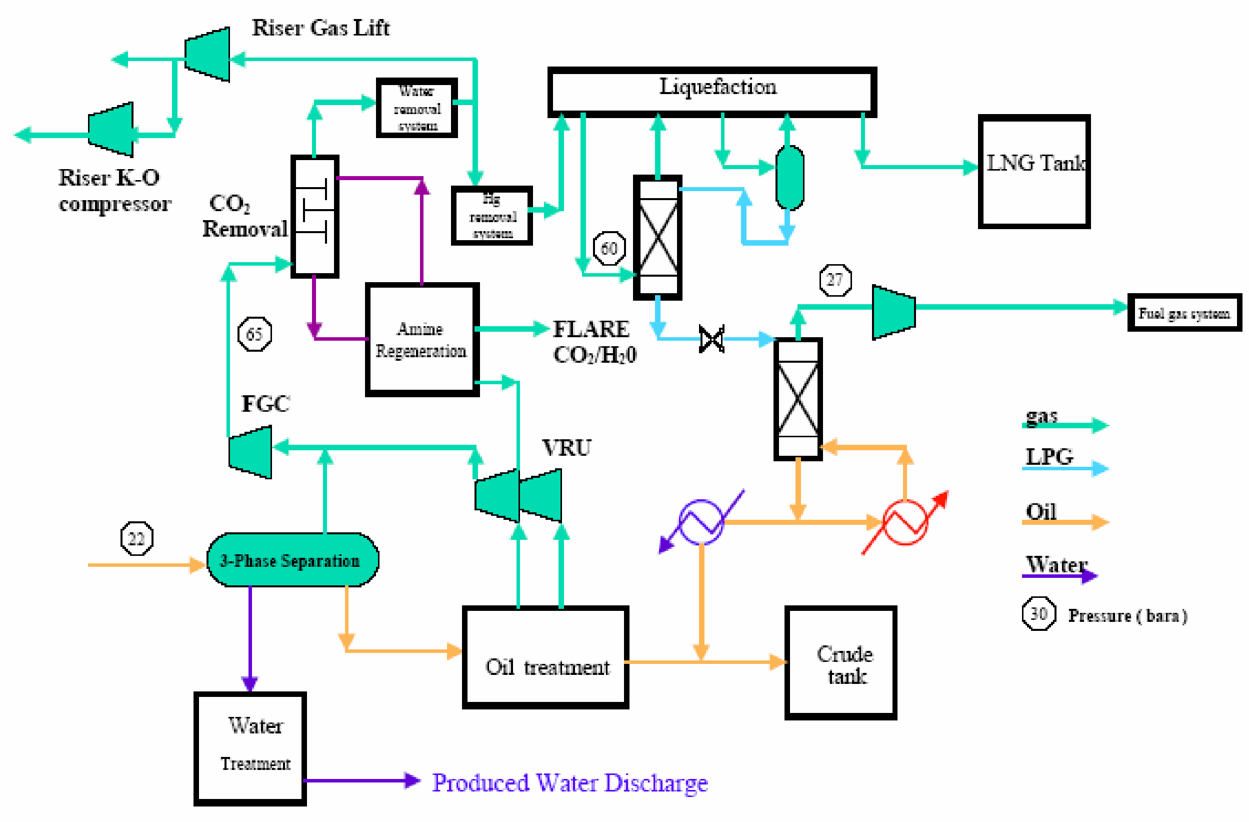


Figure 3.13.2 LNG Process Flow Diagram

The FCI and operating costs for the LNG plant were determined for the GAMS computer model. Figures 3.13.3 and 3.13.4, respectively, show the FCI and operating costs for the LNG process as a function of capacity.

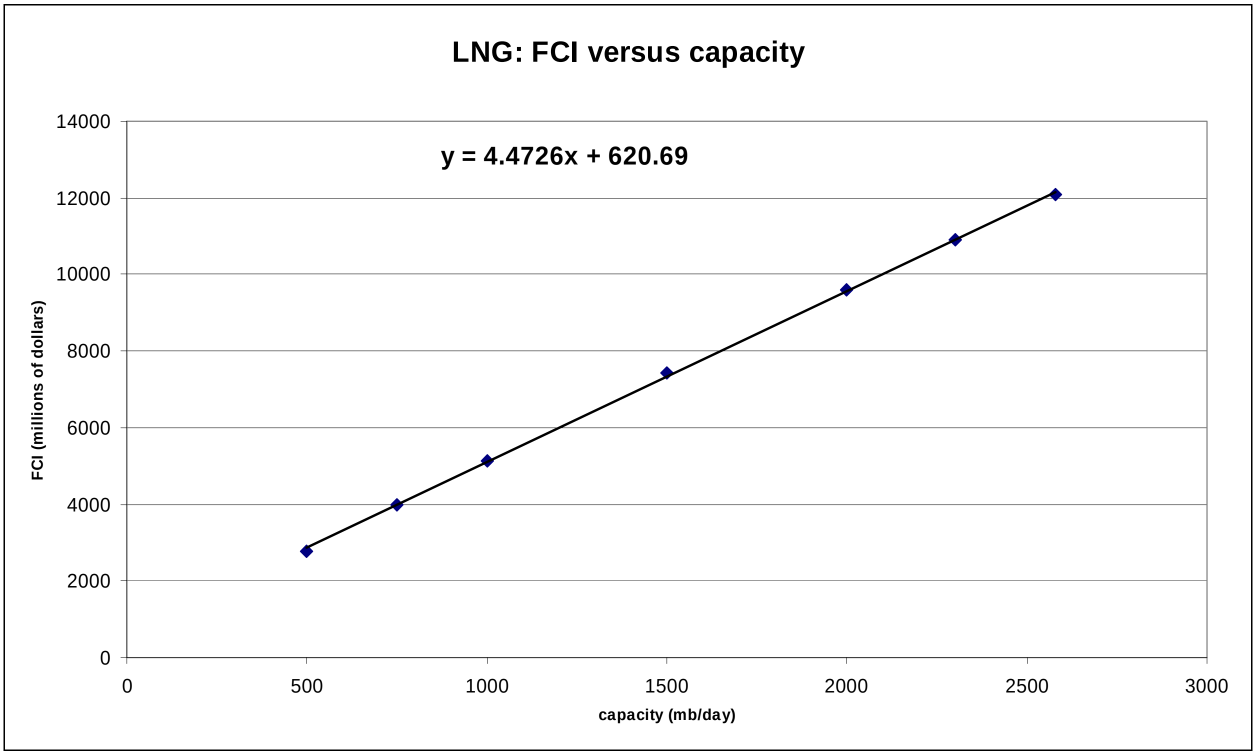


Figure 3.13.3 LNG Process FCI versus Capacity

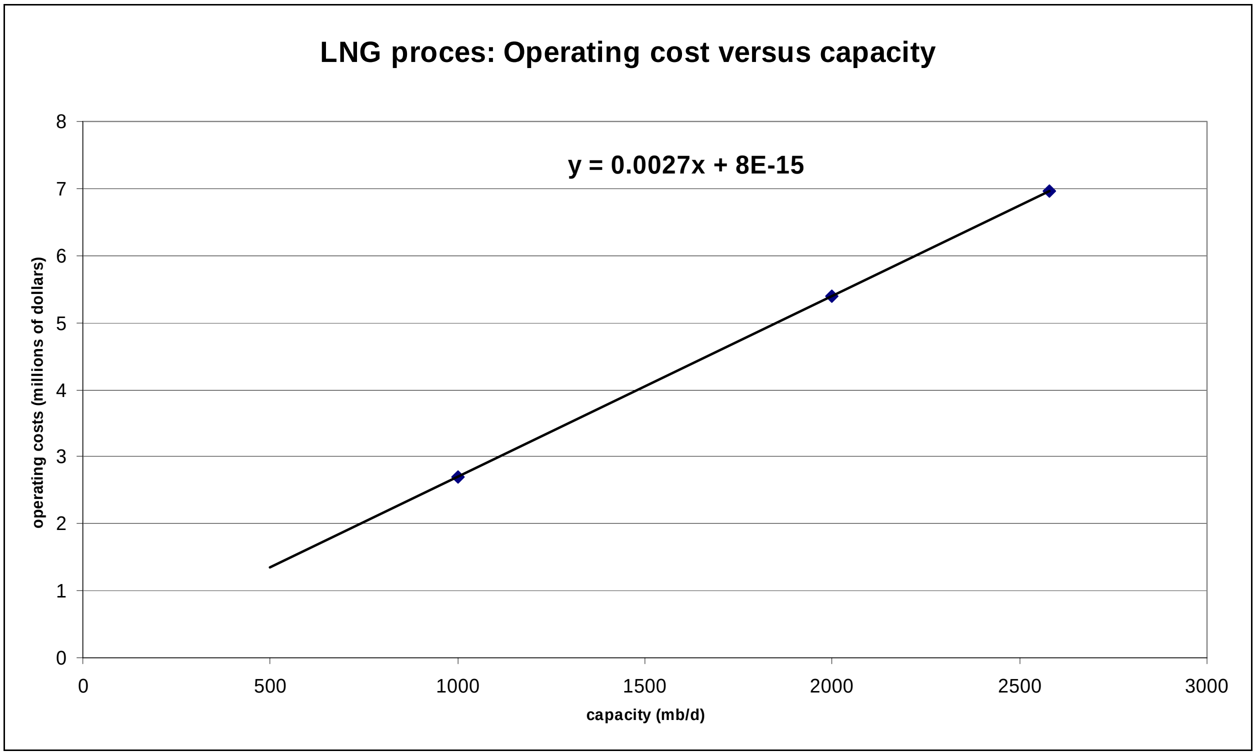


Figure 3.13.4 LNG Process versus Operating Cost

4.0 GAMS Model

4.1 Model Procedure

A linear mathematical model was created in GAMS in order to determine the maximum net present worth (NPW) of the natural gas plant. The mathematical model is used to determine which processes should be built, the year in which each process should be built, the capacity of each process, and the amount of product produced for each process based on input variables and constraints to give the highest NPW possible.

Scenarios are generated of the different input variables to make it possible for the model to analyze the input variables and determine the optimum output variables. A complete analysis of each input and output variable will be made for the next phase of the project. A material balance, the FCI and the operating costs of each process can be found in the Appendix. The model is modular so new processes can be added to the model without having to redesign the model.

4.2 Objective Function

The objective function of the model is the determination of the net present worth.
The net present worth is calculated by the equation:

$$NPW = \frac{\Sigma profit}{(1+i)^n} - Initial\ Investment - \mathrm{Re}\, investment$$

In order to determine the best NPW, the flow rate from the Camisea field is varied in order to determine the maximum and minimum flow rates through the natural gas plant. The minimum feasible flow rate is 10,000 ft^3/day. The maximum feasible flow rate is 50,000,000 ft^3/day. On either side of these flow rates, the model becomes infeasible by requiring flow rates that are smaller than the minimum size necessary to determine the FCI of each process with any degree of accuracy. At 50,000,000 ft^3/day, the size of the plants and the demand on the environment become too large to build. The capacity of each process is determined in kg/s.

4.3 Model Constraints

The model relies on several constraints and assumptions to determine the NPW of the natural gas processing plant. Each process is allowed to be built only once and the amount of feed processed is not allowed to increase once the plant is built. The minimum investment required for the plant is $6 Billion in order to account for the cost of building the two pipelines which have an estimated cost of $4.5 Billion. The minimum processing rate for the system is 90 thousand ft^3/day while the maximum has been set at an estimated realistic maximum of 10 million ft^3/day. The processing rate for natural gas distillation is assumed to be the maximum flow rate of the system in order to determine the maximum FCI of the natural gas distillation process. The money balance at the end of each year must be greater than zero in order for more processes to be added to the existing plant. The money balance was determined by investing a 1/3 of the net revenue back into the plant itself.

4.3 Tables

Table 4.3.1 FCI and Operating Cost Equations with Respect to Capacity for Each Process

Process	**Fixed Capital Investment**	**Operating Cost**
Ethylene Synthesis	$FCI = 2.22Q + 29.72$	$OC = 6.02Q + 2.18$
Low Density Polyethylene	$FCI = 4.46Q + 9.88$	$OC = 1.30Q + 2.49$
High Density Polyethylene	$FCI = 9.76Q + 12.83$	$OC = 2.10Q + 2.18$
Linear-Low Density Polyethylene	$FCI = 9.76Q + 12.83$	$OC = 21.87Q + 2.18$
Polypropylene	$FCI = 12.92Q + 26.30$	$OC = 1.74Q + 2.18$
Vinyl Chloride	$FCI = 0.58Q + 6.70$	$OC = 2.92Q + 5.10$
Polyvinyl Chloride	$FCI = 1.26Q + 16.26$	$OC = 1.85Q + 0.16$
Ammonia Synthesis	$FCI = 28.45Q + 30.59$	$OC = 6.97Q + 6.75$
Fertilizer	$FCI = 1.66Q + 10.58$	$OC = 2.04Q + 11.86$
Methanol	$FCI = 6.46Q + 16.11$	$OC = 1.68Q + 22.83$
Ethylene Glycol	$FCI = 2.05Q + 9.38$	$OC = 2.16Q + 39.22$
Fischer Tropsch	$FCI = 6.23Q + 156.31$	$OC = 1.33Q + 50.23$
Liquid Natural Gas	$FCI = 4.47Q + 620.69$	$OC = 0.0027Q + 35.4$

5.0 Results

5.1 Reinvestment

The model was designed to allow part of the total revenues be invested back into the plant in order to expand the plant which will increase the NPW. Initially, the amount of reinvestment was fixed at 20% of the total revenue per year. Money was reinvested back into the plant when there was need, however, all the processes had been constructed and the money was just being filtered from the NPW calculation lowering the expected value. This was determined to be inefficient. It was determined that the most efficient manner of calculating reinvestment was to allow the model to determine the amount of money that was to be reinvested back into the plant each year. The model was set to allow the reinvestment percentage to float between 0% and 100% which means that the value of reinvestment can range between 0% and 100% of the total revenue for each year depending on what the model determines to be the most efficient. The maximum amount of reinvestment can be lowered depending on the company's needs.

5.2 GAMS Results

The first model produced 12 processes with 10 being built in the first year and vinyl chloride and polyvinyl chloride processes being built in the fourth year. Due to the low amount of money being reinvested each year, it took several years to accumulate the capital to build the two expansion plants. In addition, money reinvested back into the plant for the next 26 years is not used to build additional processes or add capacity to the existing processes.

With the updated model, the same 12 processes were built with the same 10 built in the first year, but the vinyl chloride and polyvinyl chloride processes were built in the second year as opposed to the fourth like previously, and no additional money was reinvested. This increased the NPW from $40.5 billion to $52.5 billion. The reinvestment in the first year was 30.8%. The maximum initial investment was varied from $6 billion to $9 billion. An investment of $6 billion generated a NPW of $50.9 billion and a return on investment (ROI) of 57.6%. An actual initial investment of $8.96 billion produced the largest NPW and ROI. They were $55.0 billion and 61.4% respectively.

Table 5.2.1 NPW and ROI of Varying Maximum Initial Investments

Maximum Initial Investment (Billions)	NPW (Billions)	Actual Initial Investment (Millions)	Reinvestment (Millions)	Capital (Millions)	ROI
$6	$50.90	$6,000.00	$2,838.83	$8,838.80	57.60%
$7	$52.50	$6,504.70	$2,558.40	$9,063.10	57.90%
$8	$53.30	$6,504.70	$2,558.40	$9,063.10	58.80%
$9	$55.00	$8,963.40	$0.00	$8,963.40	61.40%

The product flow rate, process flow rate, operating costs and unit revenue for each of the processes is shown in Table 5.2.2. Two processes do not have revenue because they are intermediate products for other processes. The following results are for a maximum initial investment of $7 billion.

Table 5.2.2 Process Flow Rates, Operating Costs and Revenue at Maximum Initial Investment of $7 billion.

	Product Flow Rate (kg/s)	Process Feed Rate (kg/s)	Operating Costs (Millions)	Unit Revenue (Millions)
LNG	100.5	100.5	$240.90	$757.78
Distillation	603	603	$1,767.39	$2.26
Ethylene Synthesis	56.083	245.256	$109.69	$109.69
Vinyl Chloride	60.3	27.016	$288.96	$0.00
Polyvinyl Chloride	60.3	60.3	$288.96	$2,389.58
Ammonia Synthesis	60.3	42.766	$596.11	$0.00
Fertilizer	283.711	60.3	$1.22	$1,725.90
Methanol	120.6	60.3	$101.12	$1,205.25
Fischer-Tropsch	60.3	60.3	$88.52	$601.12
Polypropylene	1.702	1.702	$195.25	$1.53
Ethylene Glycol	48.28	29.067	$656.22	$2,014.04

All of the initial investments have the same flow rates and costs. However, the increased maximum initial investment leads to a greater NPW. All the initial investments have the same process flow rates except for the $6 billion investment which has a decreased flow rate of vinyl chloride and polyvinyl chloride. A maximum initial investment of $9 billion creates all processes in the first year but has the same process flow rates, revenue and operating costs.

5.2 Sensitivity Analysis

The prices of select products are varied to determine the effect of changing the price has on the NPW, the year in which each process is built, and the flow rate of the product. The prices of those products that are not made are increased until they are determined to be profitable by the model. This means that the product is produced in place of another product. The prices of the products that are currently being produced are lowered in order to determine the effect on the product flow rate, when the process is built, and when it is finally not worth producing that product.

Low Density Polyethylene

The price of low density polyethylene is currently $1.65/kg. The price of low density polyethylene was raised gradually until the model determined that it was most profitable to produce low density polyethylene. The profitable price of low density polyethylene is $3.50/kg. This produces a NPW of $54.1 billion. Low density polyethylene is produced under this scenario instead of polyvinyl chloride.

High Density Polyethylene

The price of high density polyethylene is $1.59/kg before it was increased to a profitable price of $3.00/kg. At this price of high density polyethylene, the NPW is $54.2 billion. Like low density polyethylene, high density polyethylene is produced instead of polyvinyl chloride.

Table 5.3.1 Price sensitivity of Low Density Polyethylene and High Density Polyethylene

	Initial Price ($/kg)	Profitable Price ($/kg)	NPW (billions)
Low Density Polyethylene	$1.65	$3.50	$54.1
High Density Polyethylene	$1.59	$3.00	$54.2

Polyvinyl Chloride

The market price for polyvinyl chloride is $1.26/kg. The price was decreased until a change in the model was realized. At $1.10/kg, the process was built in the first year instead of the second. The process flow rate is smaller so the polyvinyl chloride process FCI is less so it is built in the first year rather than the second. This decline in price leads to a NPW of $48.7 billion. If the price of the polyvinyl chloride is further reduced to a $1.00/kg, polyvinyl chloride is no longer produced and no other product is then produced to replace its ethylene flow rate. This means that less ethylene is produced which leads to a smaller natural gas flow rate into the plant.

Table 5.3.2 Price Sensitivity of Polyvinyl Chloride

Polyvinyl Chloride Price	NPW (Billions)
$1.26/kg	$52.5
$1.10/kg	$48.7
$1.00/kg	$45.80

Methanol

The price of methanol was also decreased. The price of methanol was decreased from $0.316/kg to $0.10/kg. The flow rate of methanol and the year in which methanol was built did not change with decreasing price. Lowering the price of methanol had no effect on the model. Even with the large price drop, for the given set of conditions and costs, methanol was still the most profitable product to sell. The NPW at selected prices of methanol are shown in the Table 5.3.3 below.

Table 5.3.3 Price Sensitivity of Methanol

Methanol Price	NPW (Billons)
$0.316/kg	$52.5
$0.25/kg	$48.9
$0.10/kg	$38.4

6.0 Future Work

The demand of each product currently is tied to each process. This means that the some of the lesser products do not have demand associated with them. In order to correct this, the model must be constructed in such a way that the demand of the lesser products does not dictate the flow rates of the major products such as polyvinyl chloride. Once the demand of polypropylene is met, the flow rate into ethylene synthesis must not be restricted in a way that limits the ability to produce enough polyvinyl chloride or ethylene glycol so that the demand for those products is also met.

7.0 Appendices

7.1 Sample Calculations

7.1.1 Equipment Cost Calculations

Methanol Synthesis Calculations:

Equipment Cost
Assumptions:
Mass flow rate = 1000 kg/s
Volumetric flow rate = 1250 m^3/s

Compressors:
Use mechanical energy balance to find the work needed for each compressor:

$$\left(\frac{P_2}{\rho}-\frac{P_1}{\rho}\right)=W_o$$

$$\left(\frac{4500-1500\,{}^{kN}\!/_{m^2}}{0.8\,{}^{kg}\!/_{m^3}}\right)+\left(\frac{7000-4000\,{}^{kN}\!/_{m^2}}{0.8\,{}^{kg}\!/_{m^3}}\right)+\left(\frac{7000-500\,{}^{kN}\!/_{m^2}}{0.8\,{}^{kg}\!/_{m^3}}\right)=Wo$$

$$Wo=15625kN$$

assume 80% efficiency, for 1 compressor used Table 12.28 in PT&W page 531

$$\boxed{1compressor=\$1050000}$$

Tanks:
The price for 1 tank is found using Figure 14.15 in PT&W

$$\boxed{1\tan k=\$2x10^5}$$

Heat exchangers:
Temperature differences: 190°F, 485 °F, 175 °F
U from Perry's Chemical Engineering Handbook = 600 ${}^{W}\!/_{m^2{*}C}$

$$Q=UA\Delta T$$

A = 0.06 m^2
From Figure 14.15 in PT&W, page 680 stainless steel and carbon steel shell

$$\boxed{heatexchanger=\$13200}$$

Small distillation columns:
From MSDS
Density of liquid: 681.91 kg/m^3
Density of vapor: 0.8890 kg/m^3

$$Vm = 0.3\sqrt{\frac{\rho v - \rho l}{\rho l}}$$

$$Vm = 0.3\sqrt{\frac{681.91 - 0.8890}{0.8890}} = 8.3\, {}^{ft}\!/\!{}_{s} = 2.53\, {}^{m}\!/\!{}_{s}$$

assuming 1.5 meter diameter stainless steel
small distillation column height = 2 meters
medium distillation column height = 14 meters

Table 15.16 PT&W page 796

$$\$ = \$2x10^4 / mheight$$

$$\boxed{small = \$20{,}000}$$

$$\boxed{medium = \$560{,}000}$$

Gas heated reformer (Kellogg)-stainless steel heater:
Q = 169.8 kW
From Figure 14.38 PT&W page 692

$$\boxed{heater = \$47959}$$

Combustion secondary reformer – stainless steel
From Table 14.37 PT&W page 692

$$A = 1250\, {}^{m^3}\!/\!{}_{s} * \frac{1}{100m} = 125m^2$$

heat duty

$$Q = UA\Delta T = (600W/m^2)(125m^2)(283^oC) = 2122.5kW$$

From Figure 14.37 using PT&W page 692 the price for the secondary reformer was found to be

$$\boxed{reformer = \$286500}$$

Total equipment cost for methanol synthesis plant
$2.5 million dollars

These equipment cost calculations were repeated with a different capacity, mass flow rate of 500kg/s. The total equipment cost at this capacity was found to be **$1.8 million dollars.**

For each of the remaining process, the equipment cost was calculated in the syhame manner. These equipment costs were used to determine the FCI and operating costs for each plant process

7.1.2 Tank Cost Calculations

The fixed capital investment for the tanker that will be purchased to transport natural gas and natural gas products to California was modeled after a trailership manufactured by The General Dynamic Corporation. This trailership is a twin-screw, diesel-electric light weight ship. The price was given as $150 million dollars for each tanker built. This ship has a capacity of 600 feu and 200 autos. It also includes a cargo space of 360,000 ft^2. It has the capability to reach a velocity of over 24 knots. It can hold 2.2 million kg of natural gas. Details of this ship may be found from The General Dynamics Corporation website. The FCI, calculated with respect to capacity, was found to be 6.9 dollars/kg.

The estimated cost of operating a tanker was extrapolated by information given by the US Army Corps of Engineers. The cost was found to be $19,000 per day for single hull tanker. This accounts for replacement costs, crew, lubes, maintenance, insurance, administration, and fuel. Details may be found on the US Army Corps of Engineers website. The cost is overestimated in that it assumes that the tanker will constantly be at sea. The operating cost was calculated for a distance between Los Angeles and Pisco. With the given velocity being 24 knots, the amount of time to transport natural gas and products between the two cities was found to be approximately 10 days. The operating cost is $190,000 per trip per tanker going one-way.

7.1.3 Example Material Balance Calculations

Assumed methanol synthesis produced 1 kg of product. Using the chemical reaction:
$CO + 2H_2 => CH_3OH$

Molecular weights
$CO = 28$ kg/kmol
$H_2 = 2$ kg/kmol
$CH_3OH = 32$ kg/kmol

The mass balance results in:
CO (0.875 kg) + H_2 (0.125 kg) = CH_3OH (1 kg)

Each material balance was determined in the same manner.

7.2 Price Forecasting

In order to forecast the prices of products in the 30-year life of each product, the prices of that product over the last ten years were found from some international sources. The prices were plotted versus years to find the trend line. The slope of trend line was used as the increase rate to estimate the price of the product over the next thirty years. The estimated prices of all the products over the 30-year life of the project were included in Table 7.2.1 to 7.2.6

Table 7.2.1 Price forecasting from 2005 to 2009

Products	2005 ($/kg)	2006 ($/kg)	2007 ($/kg)	2008 ($/kg)	2009 ($/kg)
Natural Gas	0.230	0.236	0.242	0.249	0.255
Diesel	0.316	0.330	0.344	0.358	0.372
Low density polyethylene	1.650	1.660	1.669	1.679	1.688
High density polyethylene	1.590	1.610	1.630	1.650	1.670
Linear low density polyethylene	1.650	1.658	1.665	1.673	1.680
Gasoline	0.757	0.794	0.830	0.867	0.904
Jet Fuel	0.941	0.994	1.048	1.102	1.156
Polypropylene	0.028	0.030	0.031	0.032	0.033
Ethylene Glycol	1.323	1.343	1.363	1.383	1.403
Methane	0.118	0.120	0.121	0.123	0.125
Methanol	0.317	0.327	0.337	0.347	0.357
Fertilizer	0.193	0.196	0.200	0.203	0.207
Butadiene	0.860	0.882	0.904	0.926	0.948
Butane	0.703	0.735	0.767	0.799	0.831
Pyrolysis Gas	1.598	1.629	1.656	1.685	1.713
Liquid Natural Gas	0.239	0.253	0.267	0.281	0.295

Table 7.2.2 Price forecasting from 2010 to 2014

Products	2010 ($/kg)	2011 ($/kg)	2012 ($/kg)	2013 ($/kg)	2014 ($/kg)
Natural Gas	0.261	0.267	0.273	0.280	0.286
Diesel	0.386	0.400	0.413	0.427	0.441
Low density polyethylene	1.698	1.708	1.717	1.727	1.736
High density polyethylene	1.691	1.711	1.731	1.751	1.771
Linear low density polyethylene	1.688	1.696	1.703	1.711	1.718
Gasoline	0.941	0.978	1.014	1.051	1.088
Jet Fuel	1.210	1.263	1.317	1.371	1.425
Polypropylene	0.034	0.036	0.037	0.038	0.039
Ethylene Glycol	1.423	1.443	1.463	1.483	1.503
Mathane	0.127	0.129	0.130	0.132	0.134
Methanol	0.367	0.378	0.388	0.398	0.408
Fertilizer	0.210	0.213	0.217	0.220	0.224
Butadiene	0.970	0.992	1.015	1.037	1.059
Butane	0.863	0.895	0.927	0.959	0.991
Pyrolysis Gas	1.742	1.772	1.802	1.833	1.864
Liquid Natural Gas	0.309	0.323	0.337	0.351	0.365

Table 7.2.3 Price forecasting from 2015 to 2019

Products	2015 ($/kg)	2016 ($/kg)	2017 ($/kg)	2018 ($/kg)	2019 ($/kg)
Natural Gas	0.292	0.298	0.304	0.311	0.317
Diesel	0.455	0.469	0.483	0.497	0.511
Low density polyethylene	1.746	1.756	1.765	1.775	1.784
High density polyethylene	1.791	1.811	1.831	1.851	1.871
Linear low density polyethylene	1.726	1.734	1.741	1.749	1.756
Gasoline	1.125	1.162	1.198	1.235	1.272
Jet Fuel	1.479	1.532	1.586	1.640	1.694
Polypropylene	0.040	0.042	0.043	0.044	0.045
Ethylene Glycol	1.523	1.543	1.563	1.583	1.603
Mathane	0.136	0.138	0.139	0.141	0.143
Methanol	0.418	0.428	0.438	0.448	0.458
Fertilizer	0.227	0.230	0.234	0.237	0.241
Butadiene	1.081	1.103	1.125	1.147	1.169
Butane	1.023	1.055	1.087	1.119	1.151
Pyrolysis Gas	1.896	1.928	1.961	1.994	2.028
Liquid Natural Gas	0.379	0.393	0.407	0.421	0.435

Table 7.2.4 Price forecasting from 2020 to 2024

Products	2020 ($/kg)	2021 ($/kg)	2022 ($/kg)	2023 ($/kg)	2024 ($/kg)
Natural Gas	0.323	0.329	0.335	0.342	0.348
Diesel	0.525	0.539	0.552	0.566	0.580
Low density polyethylene	1.794	1.804	1.813	1.823	1.832
High density polyethylene	1.892	1.912	1.932	1.952	1.972
Linear low density polyethylene	1.764	1.772	1.779	1.787	1.794
Gasoline	1.309	1.346	1.382	1.419	1.456
Jet Fuel	1.748	1.801	1.855	1.909	1.963
Polypropylene	0.046	0.048	0.049	0.050	0.051
Ethylene Glycol	1.623	1.643	1.663	1.683	1.703
Methane	0.145	0.147	0.148	0.150	0.152
Methanol	0.468	0.479	0.489	0.499	0.509
Fertilizer	0.244	0.247	0.251	0.254	0.258
Butadiene	1.191	1.213	1.236	1.258	1.280
Butane	1.183	1.215	1.247	1.279	1.311
Pyrolysis Gas	2.062	2.097	2.133	2.169	2.206
Liquid Natural Gas	0.449	0.463	0.477	0.491	0.505

Table 7.2.5 Price forecasting from 2025 to 2029

Products	2025 ($/kg)	2026 ($/kg)	2027 ($/kg)	2028 ($/kg)	2029 ($/kg)
Natural Gas	0.354	0.360	0.366	0.373	0.379
Diesel	0.594	0.608	0.622	0.636	0.650
Low density polyethylene	1.842	1.852	1.861	1.871	1.880
High density polyethylene	1.992	2.012	2.032	2.052	2.072
Linear low density polyethylene	1.802	1.810	1.817	1.825	1.832
Gasoline	1.493	1.530	1.566	1.603	1.640
Jet Fuel	2.017	2.070	2.124	2.178	2.232
Polypropylene	0.052	0.054	0.055	0.056	0.057
Ethylene Glycol	1.723	1.743	1.763	1.783	1.803
Methane	0.154	0.156	0.157	0.159	0.161
Methanol	0.519	0.529	0.539	0.549	0.559
Fertilizer	0.261	0.264	0.268	0.271	0.275
Butadiene	1.302	1.324	1.346	1.368	1.390
Butane	1.343	1.375	1.407	1.439	1.471
Pyrolysis Gas	2.244	2.282	2.321	2.360	2.400
Liquid Natural Gas	0.519	0.533	0.547	0.561	0.575

Table 7.2.6 Price forecasting from 2030 to 2034

Products	2030 ($/kg)	2031 ($/kg)	2032 ($/kg)	2033 ($/kg)	2034 ($/kg)
Natural Gas	0.385	0.391	0.397	0.404	0.410
Diesel	0.664	0.678	0.691	0.705	0.719
Low density polyethylene	1.890	1.900	1.909	1.919	1.928
High density polyethylene	2.093	2.113	2.133	2.153	2.173
Linear low density polyethylene	1.840	1.848	1.855	1.863	1.870
Gasoline	1.677	1.714	1.750	1.787	1.824
Jet Fuel	2.286	2.339	2.393	2.447	2.501
Polypropylene	0.058	0.060	0.061	0.062	0.063
Ethylene Glycol	1.823	1.843	1.863	1.883	1.903
Methane	0.163	0.165	0.166	0.168	0.170
Methanol	0.569	0.580	0.590	0.600	0.610
Fertilizer	0.278	0.281	0.285	0.288	0.292
Butadiene	1.412	1.434	1.457	1.479	1.501
Butane	1.503	1.535	1.567	1.599	1.631
Pyrolysis Gas	2.441	2.482	2.525	2.568	2.611
Liquid Natural Gas	0.589	0.603	0.617	0.631	0.645

7.3 Demand Forecasting

The demand in US for all products is used to estimate the demand in Peru due to the lack of information on the demand in Peru. The demand in US and Peru is compared with the ratio of GDP (Gross Domestic Product) and populations of both countries. The growth of demand in each year is assumed to be the same as the growth of the population, which is 1.14. Table 7.3.1 to 7.3.6 shows the demand forecasting from 2005 to 2034.

Table 7.3.1 Demand Forecasting from 2005 to 2009

	2005	2006	2007	2008	2009
Products	(kg/s)	(kg/s)	(kg/s)	(kg/s)	(kg/s)
Natural Gas	14.170	14.368	14.570	14.774	14.980
Diesel	97.800	99.169	100.558	101.965	103.393
Low density polyethylene	11.700	11.864	12.030	12.198	12.369
High density polyethylene	31.700	32.144	32.594	33.050	33.513
Linear low density polyethylene	2.040	2.069	2.098	2.127	2.157
Gasoline	0.757	0.794	0.830	0.867	0.904
Jet Fuel	0.941	0.994	1.048	1.102	1.156
Polypropylene	6.865	6.961	7.059	7.157	7.258
Ethylene Glycol	7.990	8.102	8.215	8.330	8.447
Methane	0.290	0.294	0.298	0.302	0.307
Methanol	0.590	0.598	0.607	0.615	0.624
Fertilizer	10.900	11.053	11.207	11.364	11.523
Butadiene	0.050	0.051	0.051	0.052	0.053
Mixed butane	1.120	1.136	1.152	1.168	1.184
Pyrolysis Gas	3.840	3.894	3.948	4.004	4.060
Liquid Natural Gas	66.700	67.634	68.581	69.541	70.514

Table 7.3.2 Demand Forecasting from 2010 to 2014

	2010	2011	2012	2013	2014
Products	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Natural Gas	15.190	15.403	15.618	15.837	16.059
Diesel	104.840	106.308	107.796	109.306	110.836
Low density polyethylene	12.542	12.718	12.896	13.076	13.260
High density polyethylene	33.982	34.458	34.940	35.429	35.925
Linear low density polyethylene	2.187	2.217	2.249	2.280	2.312
Gasoline	0.941	0.978	1.014	1.051	1.088
Jet Fuel	1.210	1.263	1.317	1.371	1.425
Polypropylene	7.359	7.462	7.567	7.673	7.780
Ethylene Glycol	8.565	8.685	8.807	8.930	9.055
Methane	0.311	0.315	0.320	0.324	0.329
Methanol	0.632	0.641	0.650	0.659	0.669
Fertilizer	11.685	11.848	12.014	12.182	12.353
Butadiene	0.054	0.054	0.055	0.056	0.057
Mixed butane	1.201	1.217	1.234	1.252	1.269
Pyrolysis Gas	4.116	4.174	4.232	4.292	4.352
Liquid Natural Gas	71.502	72.503	73.518	74.547	75.591

Table 7.3.3 Demand Forecasting from 2015 to 2019

	2015	2016	2017	2018	2019
Products	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Natural Gas	16.512	16.743	16.977	17.215	17.456
Diesel	113.961	115.556	117.174	118.815	120.478
Low density polyethylene	13.633	13.824	14.018	14.214	14.413
High density polyethylene	36.938	37.455	37.980	38.512	39.051
Linear low density polyethylene	2.377	2.410	2.444	2.478	2.513
Gasoline	1.162	1.198	1.235	1.272	1.309
Jet Fuel	1.532	1.586	1.640	1.694	1.748
Polypropylene	7.999	8.111	8.225	8.340	8.457
Ethylene Glycol	9.310	9.441	9.573	9.707	9.843
Methane	0.338	0.343	0.347	0.352	0.357
Methanol	0.687	0.697	0.707	0.717	0.727
Fertilizer	12.701	12.879	13.059	13.242	13.428
Butadiene	0.058	0.059	0.060	0.061	0.062
Mixed butane	1.305	1.323	1.342	1.361	1.380
Pyrolysis Gas	4.475	4.537	4.601	4.665	4.730
Liquid Natural Gas	77.722	78.810	79.913	81.032	82.167

Table 7.3.4 Demand Forecasting from 2020 to 2024

	2020	2021	2022	2023	2024
Products	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Natural Gas	17.456	17.700	17.948	18.199	18.454
Diesel	120.478	122.165	123.875	125.609	127.368
Low density polyethylene	14.413	14.615	14.819	15.027	15.237
High density polyethylene	39.051	39.597	40.152	40.714	41.284
Linear low density polyethylene	2.513	2.548	2.584	2.620	2.657
Gasoline	1.309	1.346	1.382	1.419	1.456
Jet Fuel	1.748	1.801	1.855	1.909	1.963
Polypropylene	8.457	8.575	8.695	8.817	8.940
Ethylene Glycol	9.843	9.981	10.120	10.262	10.406
Methane	0.357	0.362	0.367	0.372	0.378
Methanol	0.727	0.737	0.747	0.758	0.768
Fertilizer	13.428	13.616	13.806	13.999	14.195
Butadiene	0.062	0.062	0.063	0.064	0.065
Mixed butane	1.380	1.399	1.419	1.438	1.459
Pyrolysis Gas	4.730	4.797	4.864	4.932	5.001
Liquid Natural Gas	82.167	83.317	84.483	85.666	86.865

Table 7.3.5 Demand Forecasting from 2025 to 2029

	2025	2026	2027	2028	2029
Products	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Natural Gas	18.712	18.974	19.240	19.509	19.782
Diesel	129.151	130.959	132.793	134.652	136.537
Low density polyethylene	15.451	15.667	15.886	16.109	16.334
High density polyethylene	41.862	42.448	43.042	43.645	44.256
Linear low density polyethylene	2.694	2.732	2.770	2.809	2.848
Gasoline	1.493	1.530	1.566	1.603	1.640
Jet Fuel	2.017	2.070	2.124	2.178	2.232
Polypropylene	9.066	9.193	9.321	9.452	9.584
Ethylene Glycol	10.551	10.699	10.849	11.001	11.155
Methane	0.383	0.388	0.394	0.399	0.405
Methanol	0.779	0.790	0.801	0.812	0.824
Fertilizer	14.394	14.596	14.800	15.007	15.217
Butadiene	0.066	0.067	0.068	0.069	0.070
Mixed butane	1.479	1.500	1.521	1.542	1.564
Pyrolysis Gas	5.071	5.142	5.214	5.287	5.361
Liquid Natural Gas	88.082	89.315	90.565	91.833	93.119

Table 7.3.6 Demand Forecasting from 2030 to 2034

	2030	2031	2032	2033	2034
Products	kg/s	kg/s	kg/s	kg/s	kg/s
Natural Gas	20.059	20.340	20.625	20.914	21.207
Diesel	138.448	140.387	142.352	144.345	146.366
Low density polyethylene	16.563	16.795	17.030	17.268	17.510
High density polyethylene	44.875	45.504	46.141	46.787	47.442
Linear low density polyethylene	2.888	2.928	2.969	3.011	3.053
Gasoline	1.677	1.714	1.750	1.787	1.824
Jet Fuel	2.286	2.339	2.393	2.447	2.501
Polypropylene	9.718	9.854	9.992	10.132	10.274
Ethylene Glycol	11.311	11.469	11.630	11.793	11.958
Methane	0.411	0.416	0.422	0.428	0.434
Methanol	0.835	0.847	0.859	0.871	0.883
Fertilizer	15.430	15.646	15.865	16.088	16.313
Butadiene	0.071	0.072	0.073	0.074	0.075
Mixed butane	1.586	1.608	1.630	1.653	1.676
Pyrolysis Gas	5.436	5.512	5.589	5.668	5.747
Liquid Natural Gas	94.422	95.744	97.085	98.444	99.822

8.0 References

-Choi, Kramer, Samuel S. Tam. Design Economics of a Once-Through Natural Gas Fischer-Tropsch Plant With Power Co-Production.

-Ghaemmaghami, Bahram. "GTL: Progress and Prospects-Study yields generic, coastal-based GTL, plant." Oil & Gas Journal, vol. 99, issue 11, March 12, 2001.

-Hibbard, Walter R. "A Look Ahead at Synthetic Hydrocarbon Technology." Energy Forecast. Director, U.S. Bureau of Miles, October 1967.

-Kumar, Kastanas, Barge, Zamansky, Randy A. Seeker. "Hydrogen Refueling System Based on Autothermal Cyclic Reforming." US Department of Energy, 2002.

-Liu, Mallinson, Lance Lobban. "Nonoxidative Methane Conversion to Acetylene over Zeolite in a Low Temperature Plasma." Journal of Catalysis, 1998, vol. 179, pg 326-334.

-Meyers, Robert A. Handbook of Chemicals Production Processes. McGraw-Hill, Inc., NY 1996.

-Meyers, Robert A. Handbook of Petrochemicals Production Processes. McGraw-Hill, NY, 2005.

-Peters, Timmerhaus, and West. Plant Design and Economics for Chemical Engineers. McGraw-Hill, NY, 2003.

-European Council of Vinyl Manufacturers, PVC (Suspension Polymerised), http://www.apme.org/media/public_documents/20010820_091802/26.pdf, 1999

-European Vinyl Corporations, EVC International NV Report & Account 2000, http://www.evc-int.com/publications/ar/2000AR.pdf , 2001

-ICI Australia, Ammonia, http://www.bicyclesource.com/images/ammonia.pdf, 1992

-Kellogg Brown& Roots, Ammonia, http://www.mwkl.co.uk/pdf/Ammonia_brochure_1.pdf, 2000

-Inorganic Chemistry Industry, Ammonium Nitrate, http://www.epa.gov/ttn/chief/ap42/ch08/final/c08s03.pdf, July 1993

-Inorganic Chemistry Industry, Nitric Acid, http://www.epa.gov/ttn/chief/ap42/ch08/final/c08s08.pdf, February 1998

-Kellogg Brown& Roots, Ammonia Technology for tomorrow, http://www.halliburton.com/kbr/relatedInfo/chemicals/K0670_1AmmoniaBrochure.pdf, 2004

-Uhde A Company of ThyssenKrupp Technologies, Nitric Acid, http://www.uhde.biz/cgi-bin/byteserver.pl/pdf/broschueren/Duengemittel/Nitric_acid.pdf, 15 March 2004

-Uhde A Company of ThyssenKrupp Technologies, Organic Chemicals and Polymers, http://www.uhde.biz/cgi-bin/byteserver.pl/pdf/broschueren/Organic_Chemicals_and_Polymers/OrganicChemicals_and_Polymers.pdf, 15 March 2004

-Uhde A Company of ThyssenKrupp Technologies, Vinyl Chloride and Polyvinyl Chloride, http://www.uhde.biz/cgi-bin/byteserver.pl/pdf/broschueren/Organic_Chemicals_and_Polymers/Vinyl_chloride_and_polyvinyl_chloride.pdf, April 2004

-www.chemicalmarketreporter.com

-http://www.nassco.com/cdc/orcafactsheet.html

-http://www.usace.army.mil/inet/functions/cw/cecwp/General_guidance/EGM02-06a.pdf

-http://txspace.tamu/edu/bitstream/1969/561/1/etd-tamu-2003C-CHEN-Wu-1.pdf

-http://www.dti.gov.za/econdb/raport/rapUSA3836dg.html

-http://pubs.acs.org/hotartcl/cenear/950911/art02.html.

-www.greatlakes.com

-www.almatis.com/adcats/pdf/1usa6040.pdf

-http://www.refiningonline.com/EngelhardKB/crep/TCR1_1.htm

- Energy and Energy Savings in Distillation, http://www.svce.ac.in/~parthi/DrPLS.htm

- http://www.coogee.com.au/methanol/process.htm

-http://www.naturalgas.org/naturalgas/transport.asp

-http://doc.utwente.nl/fid/1095

-Synthetic Ammonia

-Exxon Mobile LNG Marketing

- http://actamont.tuke.sk/pdf/2004/n3/26pozivil1.pdf

-Hunt.A.UK Plant Process Capacity. April 2005.
http://www.dti.gov.uk/energy/eid/FinalReprevB.pdf.

-Steynberg A.P. Dry M.E. Fischer-Tropsch Technology. Vol. 152. Elsevier. 2004

-Coogee Energy. Methanol Process Description. March 2005
http://www.coogee.com.au/methanol/process.htm

-Gas Heated Reformer. March 2005
http://www.synetix.com/methanol/technology-ghr.htm

-Synetix. Methanol Plant Technology. March 2005
http://www.synetix.com/methanol/pdfs/general/shortlcm.pdf