Syngas Production from Petroleum Coke Gasification

University of Illinois

Chemical Engineering

Senior Design Project

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# Abstract

Petroleum coke is a major byproduct that historically has been used as a substitute for coal in power production or as a fuel in cement manufacture. The decreasing quality of crude oil refined in the United States means that more petroleum coke is being produced, often with much higher metals and sulfur content. Our objective is to evaluate a better route for using low quality petroleum coke by converting it into a feed for our linked acetic acid production team while capturing all of the sulfur, metals and most of the CO2 from combustion. Since petroleum coke is linked to the refining of crude oil, it is available at much lower cost and in much larger quantities than bio-feeds. In addition, because petroleum coke is a byproduct, and not directly extracted from the environment, it lacks the negative land use impacts of bio-feeds.

In our process, petroleum coke along with oxygen and steam are fed into an entrained flow gasifier to produce synthesis gas, a combination of carbon monoxide, hydrogen, carbon dioxide and hydrogen sulfide. Sulfur is a poison to downstream chemical production catalysts and must be removed from syngas to ppm levels by the Claus process. A significant advantage of our process is that unlike burning petroleum coke for conventional power, the CO2 from combustion can be captured and sent via pipeline for sequestration, or enhanced oil recovery. Aspen, a thermodynamic simulation tool, is used to establish the material and energy balance for the overall process.

# Executive Summary

We are producing 2600 tons per day of raw synthesis gas from the gasification of petroleum coke. Of that amount 1400 tons is carbon monoxide and 240 tons is hydrogen which fits the requirements of our linked acetic acid production group. They require that the syngas come with a H2:COmolar ratio of 2.33 We have chosen to put our plant in Pasadena, Texas which is located in the Gulf Coast region to be near the petroleum refineries to lessen transportation costs for our petroleum coke. This area also is known for its refining and chemical production so it should be a good fit. Another advantage of this area that could be used by us are green pipe lines that transport captured CO2­ for geological sequestration. This would be very helpful in enhanced oil recovery and would be a better way of disposing of our CO2 because we do produce 2100 tons a day.

Our entire facility cost $238,000,000 (approximate cost with some components missing) and covers the gasifier, gas clean up, the water gas shift, and sulfur removal. Petroleum coke costs approximately $75 a ton and therefore would cost $52,500,000 a year based on 350 days. It would seem this process is way too expensive for what is produced in the end. Carbon dioxide is the most produced component from this process and really serves no purpose to both our process and the acetic acid group’s production either and is a costly process to capture.

# Discussion

## 1: Introduction

In this project, 2600 tons of the syngas is produced using 2000 tons of the petroleum coke as a feedstock in a membrane walled gasifier designed by Shell Oil Company. The final syngas will have the molar ratio of 1:2.33 for CO and H2. The petroleum coke will be gasified using the shell membrane gasifier which is an entrained bed gasifier. This syngas will be used to produce industrial acetic acid, glacial acetic acid, and propionic acid by the chemical production team.

The plant will be built in Pasadena, Texas which is about 30 miles inland from the Gulf Coast. The complete gasification plant is assumed to occupy 140 acres of the available industrial lands. The major reason for this location is that it provides rail road and 35 mile long water way that is used for transportation.

The synthesis gas can also be produced from biomass and coal. One of the advantages of using the petroleum coke is that it eliminates the negative effect of land use compare to biomass. Furthermore, the petroleum coke is very compatible with coal. However, petroleum coke provides higher carbon yield during the gasification process compare to coal. In addition to the carbon output, petroleum coke is also less expensive than comparable coal.

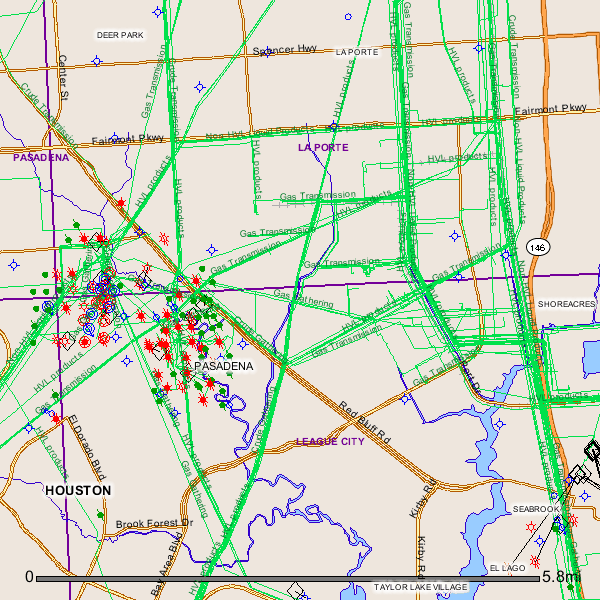
The membrane wall gasifier has an operation life span of 25 years. Thus, the process doesn’t need to shut down every six months as it is required for refractory wall gasifier. In addition, the feed does not react with any element of the membrane wall during the reaction compare to refractory wall. This gasifier can be used for all types of petroleum coke including higher grade ones. The gas is quenched using a recycle stream split from syngas which eliminates some cost on the syngas cooler by lowering the heat duty. Furthermore, the Claus process attached to the plant removes the hydrogen sulfide that is produced in the gasifier. Hence, the final syngas contains less than 2 ppm amount. This plant provides the purest synthesis gas that can be utilized to produce acetic acid and propionic acid.

Other than the acetic acid, the produced synthesis gas can also be used as building block for the production of various fuels such as methanol, acetic acid, ammonia, propionic acid, and also for the production of electricity.

### 1.1: Petroleum Coke Background

### 1.2: Gasification Technology Background

### 1.3: Location



*The above image shows the gas lines running throughout the area (green) as well as wells (green and red)*

<http://www.rrc.state.tx.us/>

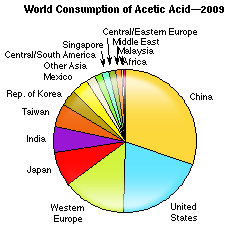
<http://www.ci.la-porte.tx.us/>

<http://www.bayareahouston.com/default/index.cfm/target-industries/chemicals-plastics/>

**Location Discussion**

Several locations were discussed before finalizing with La Porte, Texas, one of which was Victoria, Texas about 30 miles inland from the Gulf Coast, between Houston and Corpus Cristi. Victoria has a current population of about 64,000 people. Our site’s land will be bought from the Port of Victoria, which has 2000 acres of available industrial land. It is assumed that we will only need about 200 acres for our purposes. The port is a center for the chemical, construction and steel fabrication and agribusiness industries offering access to all transportation modes. Port of Victoria would have allowed us to have access to railways (Kansas City Southern, Union Pacific, Burlington Northern Santa Fe) and has a 35 mile long waterway that connects the Turning Basin in Victoria to the Gulf Intracoastal Waterway (GIWW). Roads can also be accessed from highway 59 and the soon to be established highway I-69. As attractive as the area is we wanted to be near other establishments and be in an area that has been proven, which is what we get out of Bay Port Industrial Park.

It was also discussed that the location could be in Alberta, Canada due to their excess in petroleum coke. In the short term being close to a large supply of petroleum coke would be beneficial and economically understandable, cheaper coke would lead to better profit, however, in the long run our shipping costs would be substantially higher. The United States consumes close to 20 percent of the worlds acetic acid, as shown below, which means most of our product would need to be sent internationally.



*Illustrates worlds consumption of Acetic acid. Note the substantial consumption of the United States.*

<http://www.victoriatx.org/>

<http://www.city-data.com/city/Victoria-Texas.html>

<http://www.portofvictoria.com/>

<http://www.sriconsulting.com/CEH/Public/Reports/602.5000/>

### 1.4: Research Review

## 2: Petroleum Coke Gasification Process

### 2.1: Process Overview

### 2.2: Petroleum Coke Feed

### 2.3: Petroleum Coke Gasifier

**Figure 2.3-1: Shell Gasifier System (Higman, Burgt 119)**

The ground, pressurized coke is transported along with Nitrogen gas (Because Nitrogen is an inert gas). This petcoke is supplied with 95% Oxygen and steam through nozzle of the burner on the wall. The temperature inside the reactor is about 2700 ⁰F and pressure is around 360 – 650 psi which speeds up the reaction. As a result, the syngas leaves the reactor from the top through the lock hopper. The steam produced during the heat exchange process leaves the annular space at medium pressure.

The slag comes down in the reactor where it is quenched in a water bath. The Boiler Feed Water (BFW) supplied to annular wall of the gasifier is used as water bath. The huge temperature drop due to water bath results into hardening of the slag. This slag is ground by slag crusher. The granulated slag leaves the reactor through the lock hopper and the Boiler Feed Water (BFW) supplied to cool the slag, moves to heat exchanger. Again, the BFW is supplied through another source to liquefy the slag.

The Syn-gas goes into the heat exchanger where it’s cooled by supplied Boiler Feed Water (BFW). As a result, Syn gas moves down the heat exchanger and water leaves the heat exchanger as High Pressure Steam. Additional Boiler Feed Water (BFW) is supplied from the bottom nozzle of the heat exchanger that cools the syn gas even more. (Higman, Chris, and Maarten 120) This water comes out of the heat exchanger as medium pressure and the syn gas leaves the heat exchanger at approximately 530 ⁰F and passes a candle filter unit where the solids from the gas are removed. About half the gas is then recycled via recycle gas compressor as quench gas and the other half is cooled in water scrubbing system. The gas is quenched by recycle in Shell Gasifier shown above in Figure 1. Quenching by recycle of cooled syngas is applied in the Shell gasifier. After particle removal in the candle filter, about half of the syngas flow which has a temperature around 600°F is recompressed and recycled to the gasifier outlet. By mixing the 2700°F hot syngas with the recycle stream, a cooling down to around 1650°F is achieved. Heat is then recovered in a convective syngas cooler. (Maurstad 5)

### 2.4: Solids Removal

### 2.5: Hydrogen Sulfide and Carbon Dioxide Removal

After the syngas has had any particulates removed it is then sent to the section of the plant to remove H2S and CO2. This is a pretty lengthy processes involving multiple absorbers and stripping columns to ensure that the final syngas is primarily H2 and CO­2. The most important component in this system is the absorption solvent, Selexol, which is physical solvent. Which means it does not react with the components it is removing as compared to a chemical solvent, like the commonly used MDEA. Selexol is a mixture of dimethyl ethers of polyethylene glycol with the empirical formula of CH3(CH2CH­2O)nCH3 where n is between 3 and 91. Another key difference between chemical and physical solvents is their relationship with partial pressure and their solvent loading capacity. Chemical solvents tend to plateau off at higher partial pressures unlike chemical solvents which just increase linearly with increasing partial pressure1. In the case of our system it is necessary to use a physical solvent such as Selexol to ensure that the final syngas has as little sulfur as possible and also because our process is on such a large scale where the partial pressure will be factor in the overall loading capacity. Selexol will also insure that our acid gas to the Claus plant is greater than 45% H2S, which is necessary for the Claus process to perform properly.

The process begins by sending the cooled, particulate freed syngas to the H2S absorption column where the gas is contacted with already pre-loaded Selexol which is from the CO2 absorber. The solvent is pre-loaded with CO2 ­to ensure that the CO­2 ­in the incoming gas is not absorbed on the Selexol which minimizes the temperature rise across the column1. The H­2S rich solvent then is removed through the bottom of the absorption column and sent to the regeneration cycle while the syngas and CO2 exits through the top of the absorber and is sent to the CO2 absorption process.



**Figure 2.5-1: H2S Absorption Column1**

The H2S rich solvent is then sent through a rich/lean solvent heat exchanger where the rich solvent is heated and the lean is cooled. The H2S rich solvent is then sent to a concentrator which runs at a greater temperature than the H2S absorber to remove the CO­2 to be recycled back into the feed stream for the H2S absorber1. The concentrated H2S rich solvent is then flashed to a lower pressure where the flash gas contains a higher proportion of CO2 which is then recompressed and sent back to the feed of the H2S absorber. The concentrated H2­S rich solvent then leaves the flash and is sent to the top of the H­2S stripping column where the solvent will be regenerated. The stripper removes the H2S from the Selexol where it exits through the top of the column as acid gas containing >45% H2S which will then be used in the Claus plant deal with the sulfur1. The now lean solvent leaves the stripper through the bottom and is sent to same rich/lean solvent exchanger as before where it is cooled and then cooled even more by NH3 refrigeration.



**Figure 2.5-2: H2S Regeneration Process1**

This lean solvent is then fed into the top of the CO2 absorption column where it removes the CO2 from the partially treated gas from the H2S absorber. The final syngas then exits the top of this absorber and is almost ready for use in chemical production. CO2 rich solvent comes out of the bottom where the stream is split so some is sent to the H2S absorber as the CO2 ­saturated solvent feed and the other is sent to the CO2 solvent regeneration.



**Figure 2.5-3: CO2 Absorption Column1**

This CO2 solvent regeneration is comprised of three flash vessels. The first flash vessel is the same pressure as the one used in the H2S regeneration step so that only one compressor is required that could handle the loads of both streams1. The other objective of this first flash is to recover and H2 that has been absorbed in the Selexol. The second flash takes off most of the CO2 where it is then sent to the purification process to make an end product that is 99% pure CO21. The final flash sends any remaining CO2 to the atmosphere and then sends the semi-lean solvent back to be used in the CO2 absorber.



**Figure 2.5-4: CO2 Flash Regeneration1**

The final step in this cleaning process is CO2­ purification. The raw CO2 feed gas is sent over a zinc oxide catalyst where any COS present is converted to H2S and then absorbed onto the ZnO catalyst. These reactions are the following:

COS + H2O 🡪 CO2 + H­2S

H2S + ZnO 🡪 ZnS + H2O

The gas leaving the ZnO bed is then sent over a platinum oxidation catalyst where the following reactions take place:

CO + ½ O2 🡪 CO2

H2 + ½ O2 🡪 H2O

CH4 + 2 O­2 🡪 CO2 + 2 H2O

This then leaves the remaining gas with 99% pure CO2 to be used where ever seems necessary.



**Figure 2.5-5: CO2 Purification1**

### 2.6: Claus Process (Sulfur Removal)

The Claus process takes place after the H2S removal steps to convert the H2S into elemental sulfur so it can be easily managed. The acid gas that is produced off of the stripping column in the H2S regeneration step is fed into the Claus furnace where it is combusted at 2000°F3. This combustion reaction then coverts roughly half of acid gas into sulfur and the remaining into a mix of H2S, SO2, and water vapor. The following reaction takes place in the furnace:

H2S + 1½ O2 ←→ SO2 + H2O

Upon leaving the furnace the gas is passed through a waste heat boiler to produce steam at approximately 482°F. The gas is then cooled in a condenser to pull the elemental sulfur out with the condensate. Next the gas is then reheated and passed over an alumina catalyst where the following reaction takes place in the Claus reactor:

2 H2S +SO2 ←→ 2 H2O+ 3/8 S8

Once again on leaving the reactor the gas is then condensed to remove any of the elemental sulfur. This catalytic reactor process will be repeated three times to ensure the greatest removal of sulfur. This process can be seen in the figure B-6 in appendix B. The overall reaction in the Claus process is the following:

3 H2S + 1½ O2 ←→ 3 H2O+ 3/8 S8

Another catalyst can be implemented in the final reactor to increase the overall conversion of the remaining acid gas to sulfur called the Superclaus. This catalyst oxidizes H2S to sulfur with 85% efficiency4. However it is mentioned that this high selectivity and conversion comes at a modest cost.

### 2.7: Water Gas Shift Reaction

A water as shift reaction is a reversible exothermic reaction where the reactants are carbon monoxide and steam and the desired products are hydrogen gas and carbon dioxide. The reaction is show below.

COH2O CO2 H2 (-39.01BTU/mol)

The reason this reaction process is necessary in the process is because the syngas must be at a 2: 1 molar ratio of H2: CO. The feed stock petcoke has a large carbon content and therefore mainly produces CO and CO2 with very little H2 being produced in comparison. Since such a large amount of CO is produced the syngas stream will be split and one part will remain with its high concentration of CO while the other is feed into water gas shift reactors to convert the CO to H2. The two streams will be combined in order to produce the desired molar ratio.

The kinetic of the reaction are important in order to determine the amount of desired products produced. The two main variables in the kinetic are temperature and pressure. The pressure does not affect the reaction significantly since both sides contain the same moles of gas. The equilibrium constant (products over reactants) is highly affected by temperature and can be shown by the graph below.



**Figure 2.4-1: Shows equilibrium relation to temperature (# reference)**

The equilibrium constant is also calculated with the empirical equation shown below.



The graph and equations show a favorable leaning toward low temperature reaction settings.

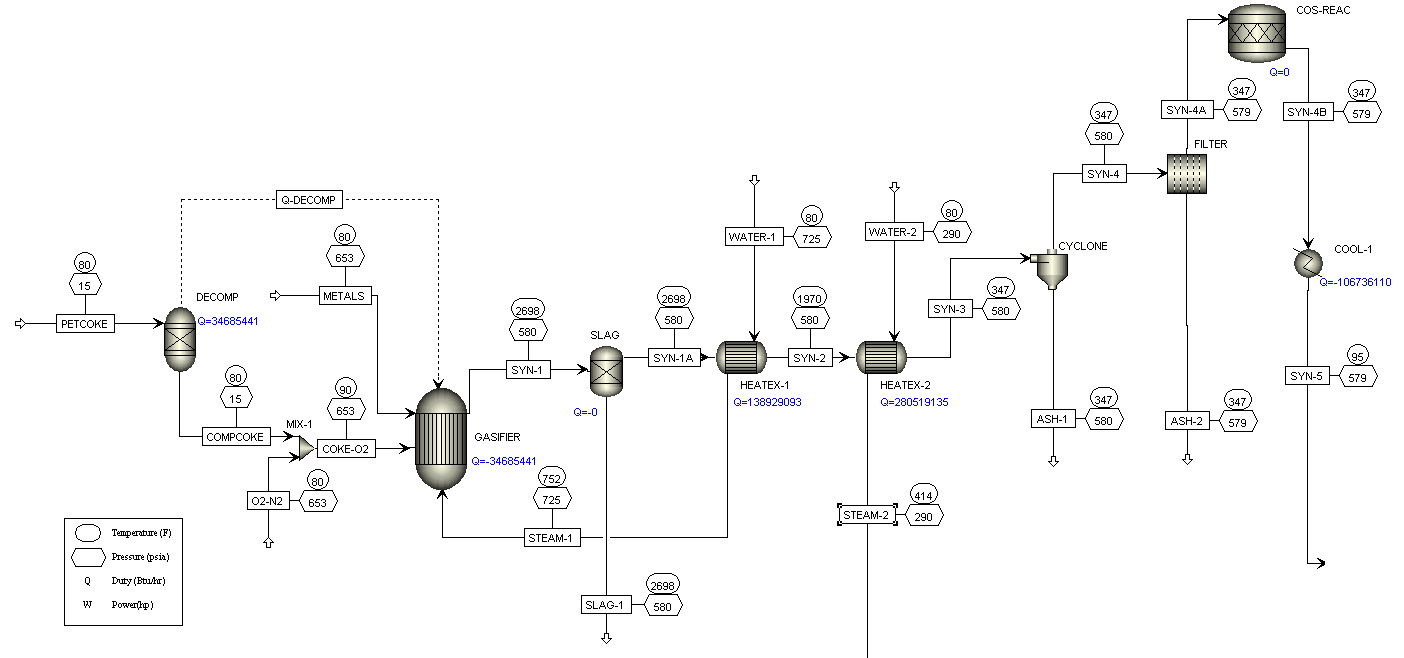
Water gas shifts are typically conducted in two stages an initial high temperature shift (HTS) then followed by a low temperature shift (LTS). Both of these reactors contain a certain amount of catalyst in relation to the feed amount. There are a variety of catalyst technologies that cater to specific shifting requirements. The following are a few of the catalyst bases available, Ferrochrome, Copper-Zinc and Cobalt – Molybdenum.

### 2.8: Heat Recovery and Utilization

## 3: ASPEN Plus Process Simulation

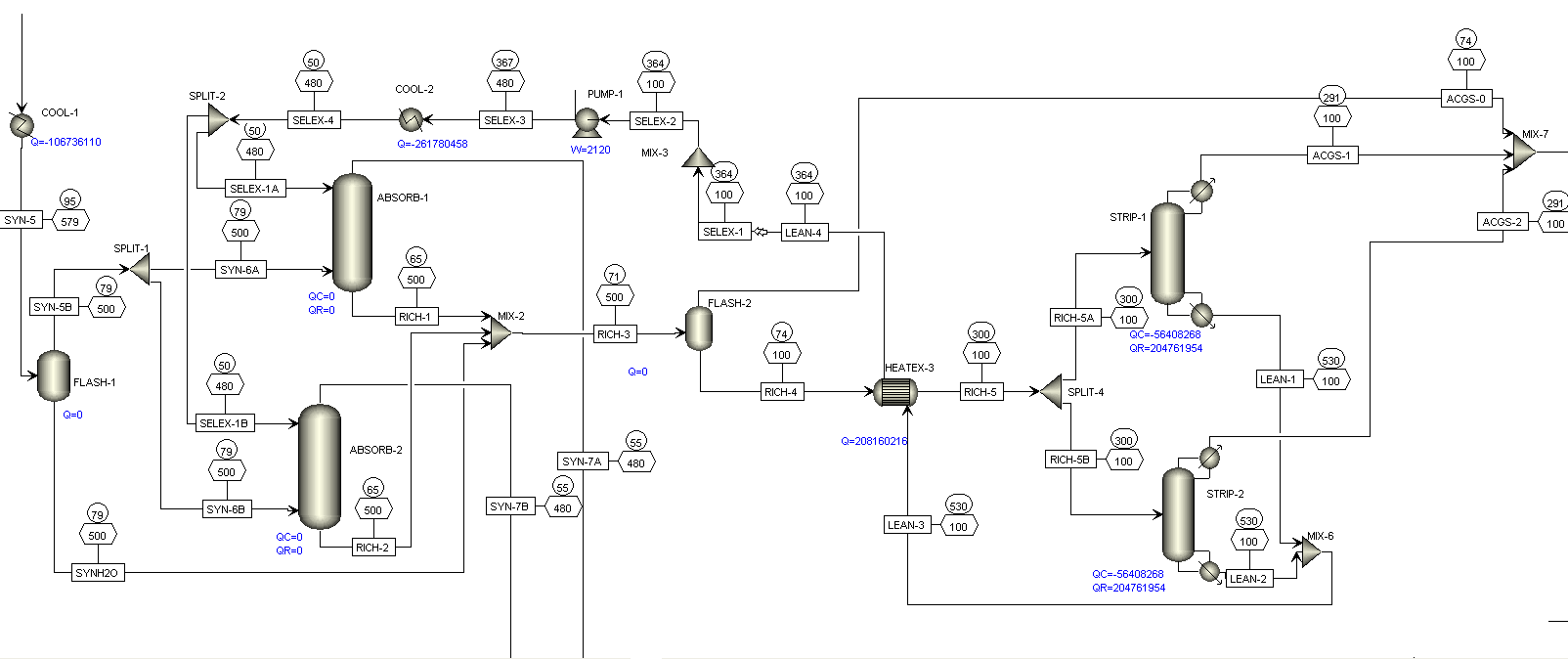
### 3.1: Simulating the Petroleum Coke Feed Stream

### 3.2: Simulating the Gasifier



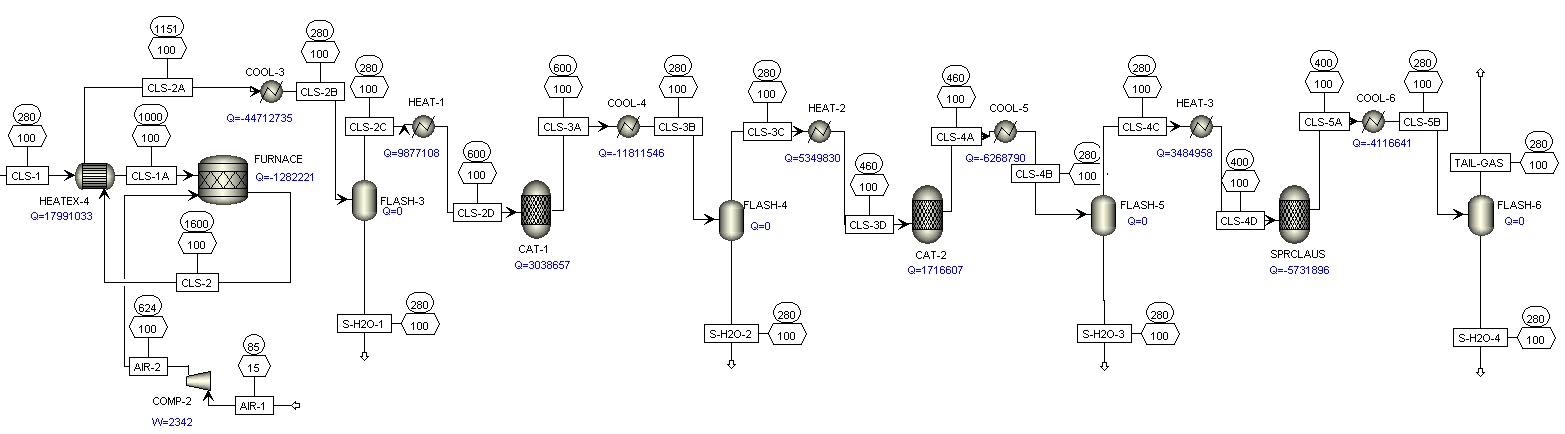
This is the gasification section. In left to right order it shows; the gasifier, two syngas heat exchangers, ash removal cyclone, ash residue bag filter, COS hydrolysis reactor, syngas cooler. After the cooler the syngas head to the sulfur removal section. This section is run or the Peng-Robinson and Peng-Robinson with Boston Mathais modification property method to account for the high temperature and pressure gas produced by the gasifier.

### 3.3: Simulating the Sour Gas Cleanup Blocks



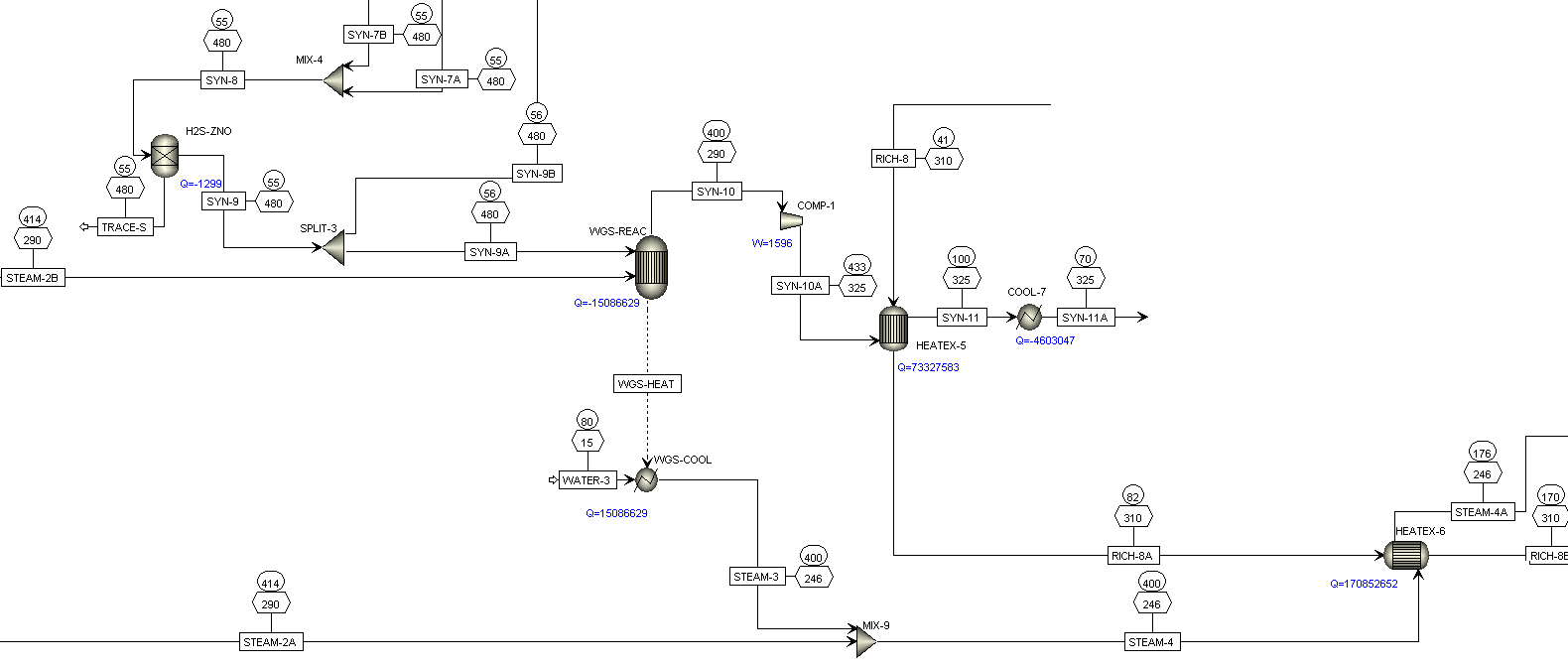
This is the hydrogen sulfide removal section. In left to right order; water knockout, two H2S scrubbers, solvent and knockout water mix, preliminary acid gas flash, heat exchanger rich and lean solvent, two strippers, lean solvent recycle. This section is run on the Schwartzentruber - Renon to account for the non ideal polar interaction of mainly H2S and H2O. The overhead gas from the absorbers heads to the water gas shift section. The acid gas from the strippers along with the flash acid gas continues on to the Claus process.

### 3.4: Simulating the Claus Process



This is the Claus process. In order from left to right; acid gas and air mix, stream heat exchanger, Claus furnace, sulfur cooler, sulfur separator, heater, first catalytic reactor, cooler, separator, heater, second catalytic reactor, cooler, separator, heater, superclause catalytic converter, cooler, separator. This process is based on the Schwartzentruber – Renon property method due to the non ideality of H2O H2S and sulfur gas. The outcomes of this process are liquid sulfur and tail gas composed of CO2 H2O and very little SO2.

### 3.5: Simulating the Water Gas Shift Reaction



This is the water gas shift along with initial CO2 sequestration section. From left to right; trace H2S-ZnO removal, syngas fraction split, water gas shift reactor, syngas ratio mixer, CO2 absorber, CO2 solvent stripper, CO2 compressor. The prepared syngas head to the chemical production team with their desired specification met. The CO2 goes toward environmentally conscience sequestration.

### 3.6: Simulating the Carbon Dioxide Capturing

### 3.7: Simulating the Heat Exchanger Blocks

### 3.8: Simulating the Compressors and Pumps

### 3.9: Simulation Results

## 4: Economics

### 4.1: Overview Summary (Written)

## 5: Environmental Review

# Recommendations

# Appendix 1: Design Basis

The objective of our process is to design a gasification system to produce synthesis gas for a downstream acetic acid production team (Team Golf). We have decided to use petroleum coke as our feedstock. Since petroleum coke, petcoke, has a high carbon content and a cheaper price. However it does have its drawbacks as well, primarily its high sulfur content. This creates a challenge when trying to clean the syngas well enough to be used for chemical production. Therefore our process contains many steps to ensure that the sulfur content is reduced to ppm levels. These steps consist of absorbing the sulfur, which is mostly in the form of H2S after the gasifier, and then converting it into elemental sulfur through the Claus process.

Our process will be on a large scale, processing approximately 2000 tons of petroleum coke a day to meet the requirements of our linked acetic acid production team. It has been specified that they need 2600 tons per day of syngas with a CO:H2 mole ratio of 0.40 and a C:H mole ratio of 0.23. They also want the incoming syngas to be at 460°F and below 725 psi. Since this is such a large operation and one that has some strict requirements we have chosen to use an entrained flow gasifier produced by Shell. This gasifier has a unique membrane wall inside to produce steam because it is run at 2900°F.

There are many environmental issues that must be addressed when running this process. One of which is the sulfur removal. After combustion in the gasifier there are large amounts of H2S produced that cannot be released to the atmosphere. This must be removed downstream in an absorber where it is then separated from the raw syngas and converted into elemental sulfur in the Claus process. Once converted to elemental sulfur it is easily manageable and can be sold off. Another environmental issue is managing what happens to all of the CO2 produced throughout the entire process. During the sulfur removal step the CO2 will also be absorbed in a separate column of Selexol and after the Selexol regeneration step the CO2 will be removed from the solvent and will be capture ready.

Like was mention before we are using a not commonly used feedstock for our process, petroleum coke. Petroleum coke is a byproduct of petroleum refining that is high in carbon content but also high in sulfur, on average 6.14% by weight sulfur. An advantage of using petroleum coke is its somewhat cheaper price as compared to coal averaging about $75 per ton2. Other benefits are its large supply do to the many refineries in the Gulf Coast and lack the negative environmental destruction such as coal or bio-feeds.

# Appendix 2: Block Flow Diagram

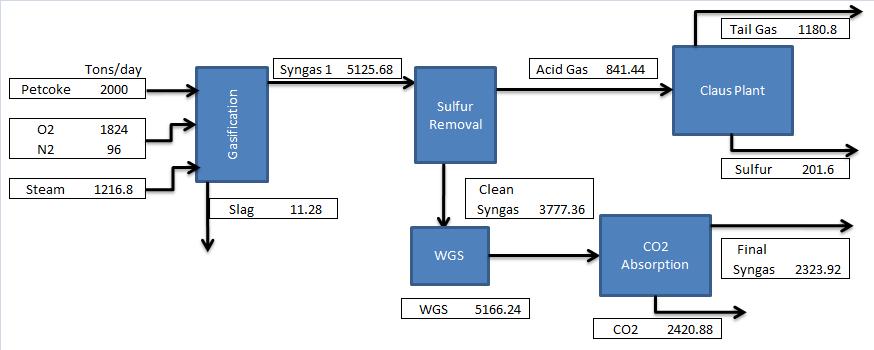


Figure A2.1: Block Flow Diagram of Process in Tons/Day

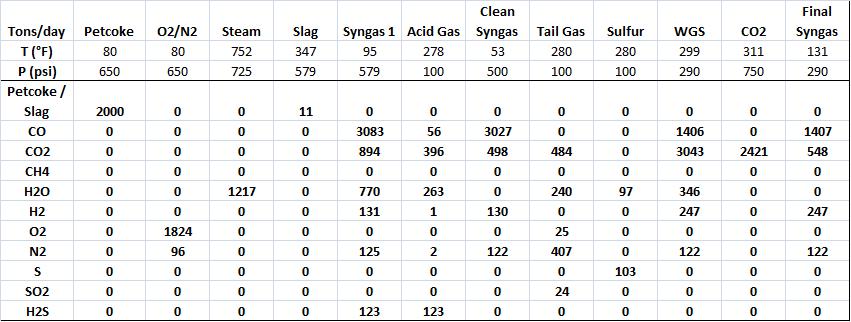


Table A2.2: Data Corresponding to Figure A2.1

# Appendix 3: Process Flow Diagram

Figure A3.1: Gasification Section

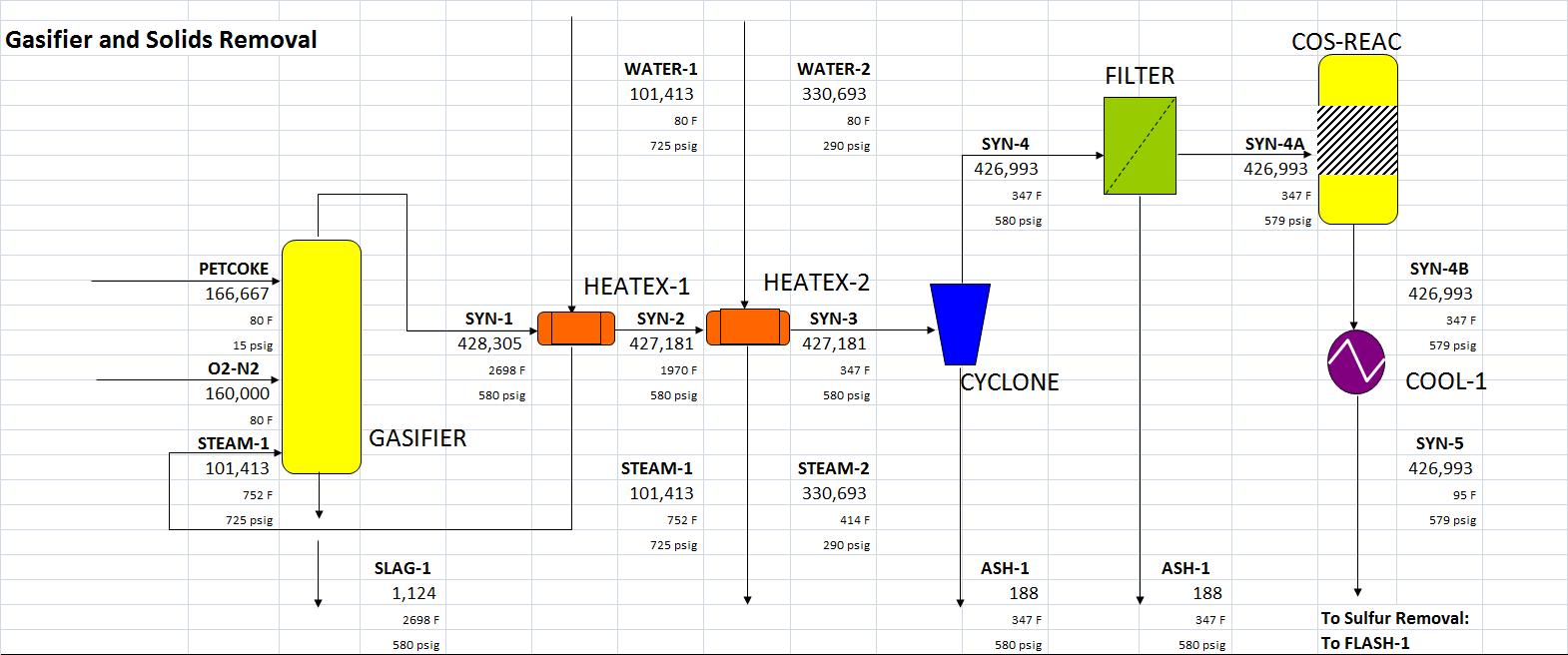




Table A3.1: Gasification Section Simulation Data

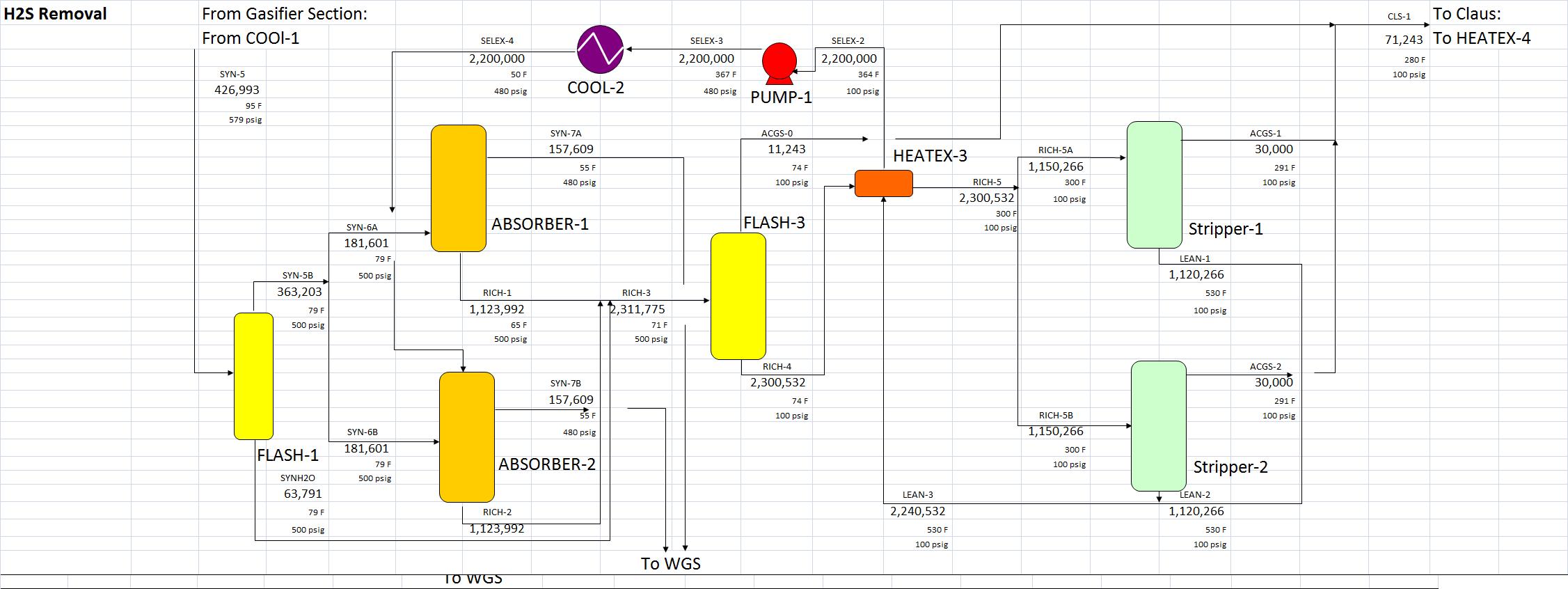


Figure A3.2: Hydrogen Sulfide Removal



Table A3.2: Hydrogen Sulfide Removal Simulation Data pt1/2



Table A3.3: Hydrogen Sulfide Removal Simulation Data pt2/2

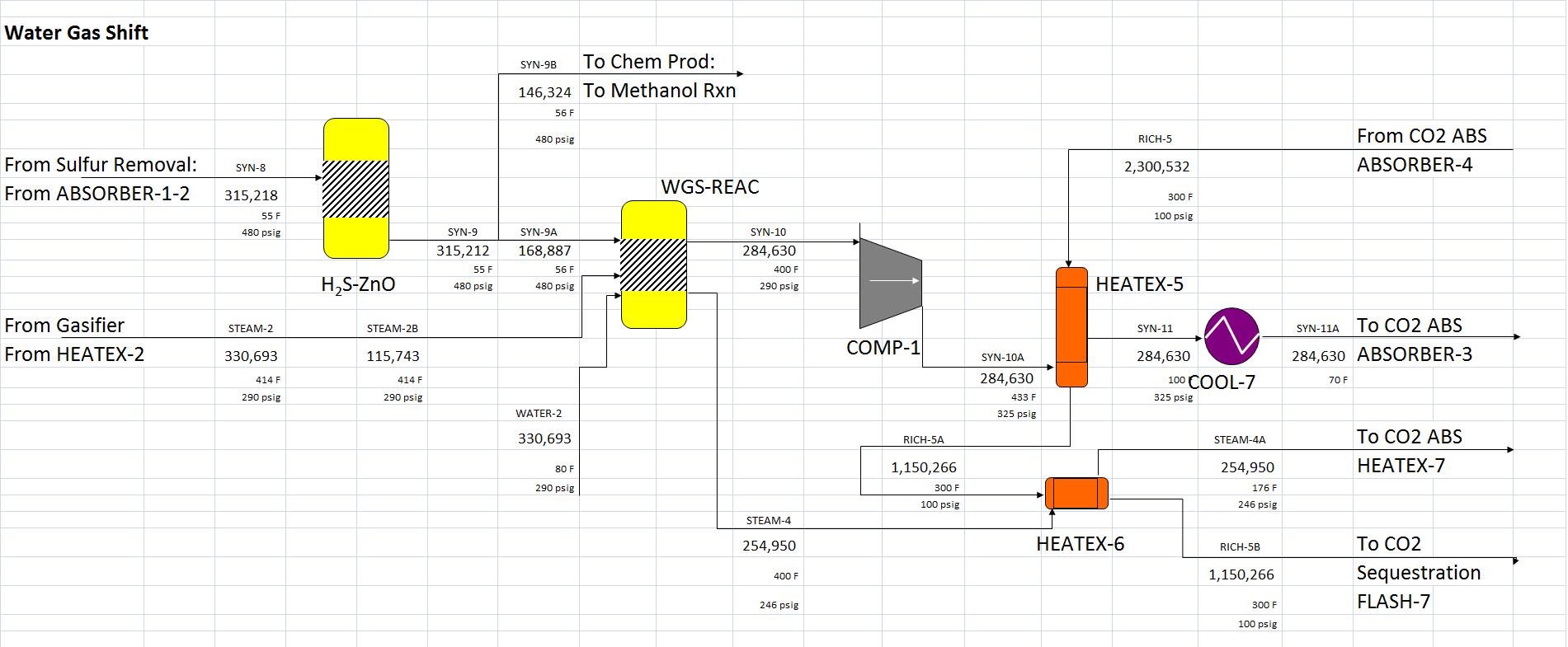
  


Figure A3.3: Water Gas Shift Section

Table A3.4: Water Gas Shift Simulation Data

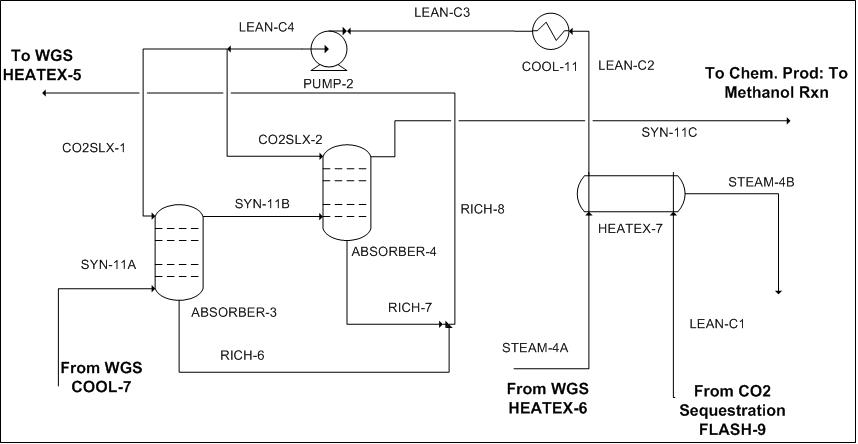


Figure A3.4: Carbon Dioxide Absorption Section



Table A3.5: Carbon Dioxide Absorption Simulation Data

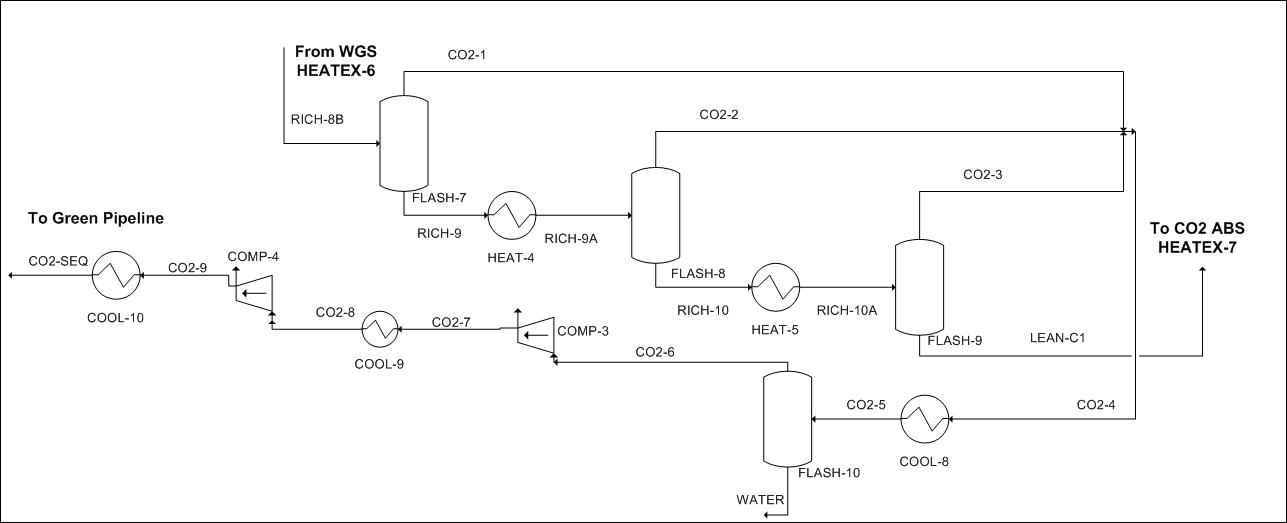


Figure A3.5: Carbon Dioxide Sequestration Section



Table A3.6: Carbon Dioxide Sequestration Simulation Data

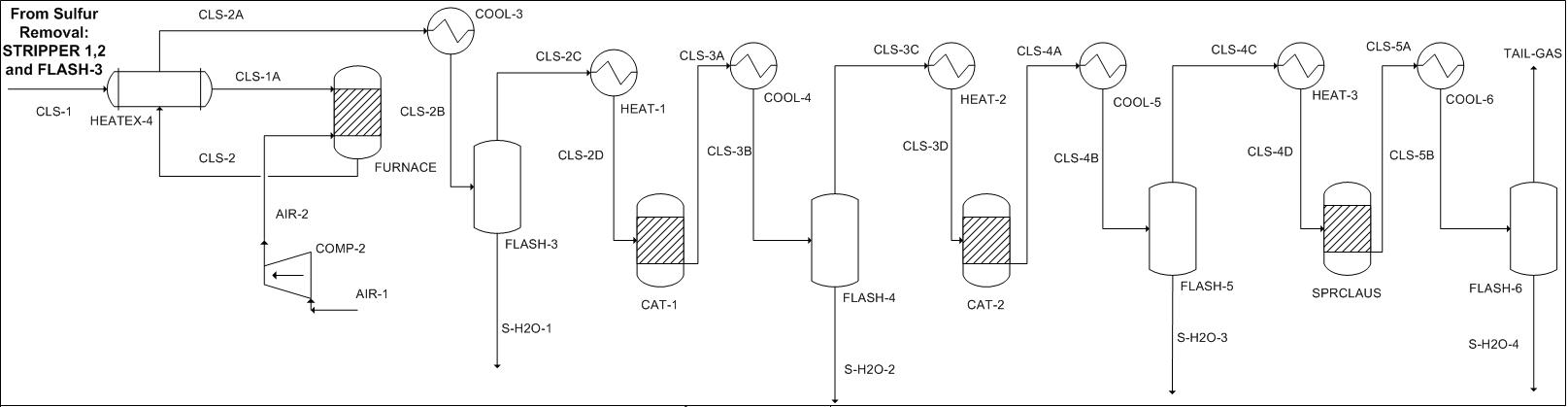


Figure A3.6: Claus Process



Table A3.7: Claus Process Simulation Data

# Appendix 4: Material and Energy Balance

## A4.1: Sample Energy Balance Around the Water Gas Shift

|  |  |
| --- | --- |
| **Energy absorbed by cooler ΔH = 7.17E07**  **HTS REACTOR**  **(lb-mols/hr)**  **T = 900 ˚F**  **T = 300 ˚F**  **T = 400 ˚F**  **Assuming overall process at constant pressure of 290 psi**  **Heat of Reaction ΔH = -8.53E07 (Btu/hr)**   |  | | --- | |  | |   **Other Assumptions:**  **Water is entering cooler at 85 F**  **Steam is created through the cooler is at 350 F**  **CO = 4946**  **H2O = 6425**  **N2 = 200**  **H2 = 2960**  **CO2 = 469**  **CH4 = 0.255**  **CO = 123**  **H2O = 1602**  **N2 = 200**  **H2 = 7783**  **CO2 = 5292**  **CH4 = 0.255**  **Amount of steam produced = 3491** |
|
|
|
|
|
|
|
|
|
|

(Figure A4.1: Schematic for the energy balance)

Calculations:

The energy balance was done on the water-gas shift reactor. In this reactor only one reaction takes place.

The table below summarizes the average Cp for the syngas in the reactor:

|  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- |
| Inlet Fraction Clean Syngas Component Flow for HTS | (lbmoles/hr) for HTS | Cp  (KJ/Kg K) | Molecular weight | CP  (btu/  lbmole\* F) | Outlet Syngas Comp (lbmole/ hr) form (HTS) | Average Cp (btu/lbmole\* F) for HTS |
| CO | 4946.233 | 1.0641 | 28.010 | 7.118 | 123.395 | 0.059 |
| H2O | 6424.688 | 3.5712 | 18.020 | 15.367 | 1601.850 | 1.641 |
| N2 | 199.954 | 1.0703 | 28.000 | 7.158 | 199.954 | 0.095 |
| H2\* | 2960.168 | 14.4004 | 2.020 | 6.948 | 7783.006 | 3.605 |
| CO2 | 469.211 | 1.0055 | 44.010 | 10.559 | 5292.049 | 3.725 |
| CH4\* | 0.255 | 2.8896 | 18.020 | 12.432 | 0.255 | 0.00021 |
| H2S | 0.000 | 0.000 | 0.000 | 0.000 | 0.000 | 0.000 |
| COS | 0.000 | 0.000 | 0.000 | 0.000 | 0.000 | 0.000 |
| Total | **15000.509** | **23.998** |  | **59.583** | **15000.509** | **9.125** |

The table below shows the assumption for the cooler:

|  |  |
| --- | --- |
|  | ΔH of Water (btu/lbm) |
| Assuming that cooling water is coming at 85 ⁰F | 53.08 |
| Saturated steam is being created 350 ⁰F | 1192.97 |

Based on this assumption;

# Appendix 5: Calculations

## A5.1: Determination of Zinc Oxide Catalyst Need

ZnO: 81.4 g/mol

H­2S: 34 g/mol

This is the overall reaction of zinc oxide to zinc sulfide which is a no reusable product.  
 From our Aspen model it was determined that there would be 6.835 lbm/hr of H2S.

This equation holds because the overall reaction is 1 to 1 for ZnO and H2S

Using 350 days for a year of plant running time.

Purchasing ZnO for $55 per lbm from UOP.

# Appendix 6: Annotated Equipment List

## A6.1: Gasification Process

****

|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Equipment Name** | **Description** | **Material** | **Unit** | **Size** | **Equipment Cost** | **Module Factor** | **Direct Cost** |
| Ring Gran Petcoke Crusher **Invalid source specified.** | Grinding Petroleum coke before pressurizing it with N2 gas | A285C | hp | 1050 | $428,860.00**Invalid source specified.** | 1.30**Invalid source specified.** | $557,518.00 |
| Gasifier(x2) | Petroleum Coke reacts with oxygen and produces raw syngas and slag | Membrane Wall |  |  | N/A |  | $132,000,000.00**Invalid source specified.** |
| Heat Exchanger 1 | Raw syn gas goes through 1st Heat exchanger which releases the high pressure steam |  | ft2 | 492 | $40,300.00 (Aspen) | 3.20 (Perry) | $128,960.00 |
| Heat Exchanger 2 | Raw syn gas goes to 2nd heat exchanger from where medium pressure steam is released |  | ft2 | 2857 | $117,400.00 (Aspen) | 3.20 (Perry) | $375,680.00 |
| Candle Filter | Raw syn gas goes through candle filter where solids are removed | Carbon Steel |  |  | $364,770.00  **Invalid source specified.** | 2.80 (Perry) | $1,021,356.00 |
| Cyclone 1 (x3) | Comes from candle filter and leaves fly ash behind while leaving from cyclone | Carbon Steel | in | d = 59 | $307,800.00 (Aspen) | 3.30 (Silla) | $1,015,740.00 |
| Total |  |  |  |  | **$ 1,259,130.00** |  | **$ 135,099,254.00** |

The above table A6.1 describes the equipment required for the gasification process only. During this process, raw synthesis gas is created which later proceeds to Sulfur Removal process.

The gasification process alone costs approximately $135 million including the installation costs. The required equipment cost approximately $1.2 million as shown in table A6.1. This estimated cost doesn’t include the cost of raw material such as, Petroleum coke, Oxygen, and Nitrogen.



## A6.2: Sulfur Removal

### A6.2.1: H2S (Acid Gas) Removal



|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Equipment Name** (Aspen) | **Description** | **Material** (Aspen) | **Unit** | **Size** | **Equipment Cost** | **Factor** | **Direct Cost** |
| COS Reactor | Converts all carbonyl Sulfide to Hydrogen Sulfide | CS | ft | d = 6.3 h = 19 | $188,000.00 (Aspen) | 1.9 (Perry) | $357,200.00 |
| Cooler 1 | Cools the syn gas coming from COS reactor |  | sq. Ft | 6570 | $162,300.00  (Aspen) | 2.7 (Perry) | $438,210.00 |
| Absorber (x2) | Selexol absorbs H2S and that H2S leaves from the bottom | Jacket (CS) tray(A285C) | ft | d = 10 h = 124 | $2,400,000.00 (Aspen) | 3.0 (Towler) | $7,200,000.00 |
| Stripper (x2) | Removes H2S from selxol ; H2S leaves column from top | Jacket (CS) tray(A285C) | ft | d = 17 h = 30 | $1,560,000.00  (Aspen) | 3.0 (Silla) | $4,680,000.00 |
| Flash Vessel 1 | Recovers H2S |  | cu. Ft | 383 | $106,000.00  (Aspen) | 2.8 (Silla) | $296,800.00 |
| Flash Vessel 2 | CO2 is removed from Selexol and H2S concentrated Selexol ; CO2 rich Selexol goes to cooler 2 and H2S rich Selexol goes to Scrubber |  | cu. Ft | 6980 | $601,230.00 (Aspen) | 2.8 (Silla) | $1,683,444.00 |
| Heat Exchanger 3 | Heating the H2S and Selexol then it goes to Flash |  | sq ft | 5367 | $121,980.00 (Aspen) | 3.2 (Perry) | $390,336.00 |
| Cooler 2 | CO2 rich Selexol goes back to H2S absorber |  | sq ft | 6739 | $168,800.00 (Aspen) | 2.7 (Perry) | $455,760.00 |
| **Total** |  |  |  |  | **$5,308,310.00** |  | **$15,501,750.00** |

This is the initial step in Sulfur recovery process. This process costs approximately $15.5 million excluding the cost of Claus, Super Claus Process, and the Tail Gas Treatment. The sizes have been determined through Aspen simulation. This cost doesn’t include the cost of Selexol that is used in this process to absorb H2S gas.



### A6.2.2: Claus Process



|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Equipment Name** (Aspen) | **Description** | **Material** | **Unit** | **Size** | **Equipment Cost** (Aspen) | **module factor** | **Direct Cost** |
| Heat Exchanger 4 | H2S rich Selexol |  | sq ft | 238 | $92,300.00 | 3.2 (Perry) | $295,360.00 |
| Claus Furnace | H2S is converted to SO2 | CS |  |  | $309,000.00 | 2.1 (Perry) | $648,900.00 |
| Cooler 3 |  |  | sq ft | 439 | $16,700.00 | 2.7 (Perry) | $45,090.00 |
| Flash Vessel 3 |  |  | cu ft | 680 | $86,758.00 | 2.8 (Towler) | $242,922.40 |
| Heater 1 |  |  | sq ft | 1574 | $38,400.00 | 2.0 (Silla) | $76,800.00 |
| Cat. Reactor 1 | H2S and SO2 react with each other and forms Elemental S8 |  |  |  | $223,300.00 | 1.9 (Perry) | $424,270.00 |
| Cooler 4 | Reaction is condensed and enters Flash 4 |  | sq ft | 257 | $13,600.00 | 2.7 (Perry) | $36,720.00 |
| Flash Vessel 4 | Condensed S8 as liquid |  | cu ft | 461 | $90,009.00 | 2.8 (Towler) | $252,025.20 |
| Heater 2 | Unreacted H2S and S8 is heated up and enters cat reactor 2 |  | sq ft | 996 | $25,100.00 | 2.0 (Silla) | $50,200.00 |
| Cat. Reactor 2 | Unreacted H2S and SO2 react with each other and forms Elemental S8 |  |  |  | 191,400.00 | 1.9 (Perry) | $363,660.00 |
| Cooler 5 | Reaction from cat reactor 2 is condensed and enters Flash 5 |  | sq ft | 162 | $11,800.00 | 2.7 (Perry) | $31,860.00 |
| Flash Vessel 5 | condensed S8 as liquid |  | cu ft | 461 | $86,266.00 | 2.8 (Towler) | $241,544.80 |
| **Total** |  |  |  |  | **$1,184,633.00** |  | **$2,709,352.40** |

Table A6.2.2 above for Claus Process describes the required equipment for the Claus Process which removes Sulfur from the H2S gas. This is the second step in Sulfur Recovery process which costs around $2.7 million including installation cost. These equipments have been determined based on Aspen Simulation.



### A6.2.3: Super Claus Process



|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Equipment Name** | **Description** | **Material** | **Unit** | **Size** | **Equipment Cost** (Aspen) | **Factor** | **Direct Cost** |
| Heater 3 |  |  |  | d = 6.3 h = 19 | $27,500.00 | 2.0 (Silla) | $55,000.00 |
| Superclaus Reactor |  |  | ft2 | d = 6.3 h = 19.1 | $188,200.00 | 1.9 (Perry) | $357,580.00 |
| Cooler |  |  |  |  | $10,100.00 | 2.7 (Perry) | $27,270.00 |
| Flash Vessel 6 |  |  | ft2 | 6x12 | $27,700.00 | 2.8 (Silla) | $77,560.00 |
| **Total** |  |  |  |  | **$253,500.00** |  | **$517,410.00** |

In addition to Claus Process, Super Claus process is the third process required in Sulfur Recovery. The cost of this process has been estimated to $0.5 million including installation cost. This cost doesn’t include the price for the required catalyst. The total cost Claus Process would add up to $3.2 million without the cost of Selexol, required electricity, and tail gas treatment.



## A6.3: CO2 Capture

|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Equipment Name** (Aspen) | **Description** | **Material** | **Unit** | **Size** | **Equipment Cost** (Aspen) | **Factor** | **Direct Cost** |
| Compressor 1 |  |  | HP | 1596 | $1,060,600.00 | 2.6 (Perry) | $2,757,560.00 |
| Compressor 2 |  |  | HP | 2341 | $2,588,500.00 | 2.6 (Perry) | $6,730,100.00 |
| Compressor 3 |  |  | HP | 2218 | $1,180,100.00 | 2.6 (Perry) | $3,068,260.00 |
| Compressor 4 |  |  | HP | 1646 | $1,107,600.00 | 2.6 (Perry) | $2,879,760.00 |
| Cooler 7 |  |  |  |  | $16,300.00 | 2.7 (Perry) | $44,010.00 |
| Flash 7 |  |  | ft | D = 14.5 H = 12 | 438,000.00 | 2.8 (Silla) | $1,226,400.00 |
| Cooler 8 |  |  |  |  | $28,400.00 | 2.7 (Perry) | $76,680.00 |
| Flash 8 |  |  | ft | D = 15.5 H = 41 | 329,500.00 | 2.8 | $922,600.00 |
| Cooler 9 |  |  |  |  | $15,100.00 | 2.7  (Perry) | $40,770.00 |
| Flash 9 |  |  | ft | D = 15 H = 44.5 | 254,000.00 | 2.8 | $711,200.00 |
| Cooler 10 |  |  |  |  | $16,700.00 | 2.7 (Perry) | $45,090.00 |
| Flash 10 |  |  | ft | D = 5 H = 12 | 27,000.00 | 2.8 | $75,600.00 |
| Heat Exchanger 5 |  |  | ft2 | 3606.929 | 82,600.00 | 3.2 (Perry) | $264,320.00 |
| Heat Exchanger 6 |  |  | ft2 | 5287.342 | 108,900.00 | 3.2 (Perry) | $348,480.00 |
| Heat Exchanger 7 |  |  | ft2 | 1816.31 | 42,500.00 | 3.2 (Perry) | $136,000.00 |
| Heater 4 |  |  |  |  | 777,500.00 | 2.0 (Silla) | $1,555,000.00 |
| Heater 5 |  |  |  |  | 191,400.00 | 2.0 (Silla) | $382,800.00 |
| Heater 6 |  |  |  |  | 637,400.00 | 2.0 (Silla) | $1,274,800.00 |
| Absorber 3 |  |  | ft | D = 15.5 H = 98 trays =43 | 1,529,200.00 | 3.0 (Towler) | $4,587,600.00 |
| Absorber 4 |  |  | ft | D = 15.5 H = 98 trays=43 | 1,061,500.00 | 3.0 (Towler) | $3,184,500.00 |
|  |  |  |  |  |  |  |  |
| **Total** |  |  |  |  | **$11,492,800.00** |  | **$30,311,530.00** |

The required equipment for this process have been estimated based on aspen simulations. The total direct cost of this process is $ 30 million which doesn’t include the cost of catalyst and electricity.



## A6.4: Water-Gas Shift

# Appendix 7: Economic Evaluation

## A7.1: Individual



### A7.1.1: Annual Operating Costs

The annual raw material costs are described in the table below. The cost of the catalyst, solvent, and the petroleum coke were provided by the companies that were contacted. The cost of cooling water was obtained from the ASPEN simulation.

The annual operating costs were determined on the basis that plant would be operating 350 days/year.

Table A7.1.1: Raw material Costs

|  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- |
| **Raw Materials** | **Quantity** | | **Price ($)** | | **Total Cost (day)** | **Cost ($)/(year)** |
| **Petcoke (T/D)** | 2000 | (ton/day) | 75.00 | ($/mt) | $ 150,000 | $ 52,500,000 |
| **Zinc oxide (lbm/day)** | 392.73 | (lbm/day) | 55.00 | ($/lbm) | $ 21,600 | $ 7,560,053 |
| **selexol** | 4000000 | (lbm/day) | 3.20 | ($/day) | $ 12,800,000 | $ 12,800,000 |
| **Ferro-chrome** | 53001.535 | (lbm/day) | 1.33 | ($/lbm) | $ 70,492 | $ 24,672,215 |
| **Aluminum oxide** | 449.52171 | (lbm/day) | 2548.75 | ($/110 lbm) | $ 10,416 | $ 3,645,468 |
| **oxygen** | 1655 | (ton/day) | 61.00 | ($/ton) | $ 100,955 | $ 35,334,250 |
| **Total** |  |  |  |  | $ 13,052,508 | $ 136,511,985 |

Table A7.1.2: Operating Labor Cost

|  |  |
| --- | --- |
| Shifts | 3 |
| operator per shift | 3 |
| additional operators | 2 |
| total operators | 11 |
| Hours per operator | 2920 |
| salary ($20/hr) per operator | $ 58,400.00 |
| total salary | $ 642,400.00 |
| Fringes (40% of the salary) | $ 256,960.00 |
| total salaries (per year) | $ 899,360.00 |

Table A7.1.3: Total Annual operating costs:

|  |  |
| --- | --- |
| Type of the cost | Cost($/year) |
| Raw materials | $136,511,985 |
| Cooling Water | $ 1903448.74 |
| Maintenance cost (3% of the capital cost) | $6,569,118 |
| Electricity | $ 17,765,927 |
| Salaries and fringes | $899,360 |
| Total | $163,649,839 |

### A7.1.2: Expected Revenues

Table A5.1.4: Expected Revenues

|  |  |  |  |
| --- | --- | --- | --- |
| **Products** | **Quantity Produced(ton/day)** | **Price ($/ton)** | **Annual revenue** |
| **Syn Gas** | 1652 | $ 300 | $173,460,000 |
| **Sulfur** | 100 | $ 170 | $5,950,000.00 |
| **Carbon dioxide** | 2000 | $ 40 | $28,000,000.00 |

### A7.1.3: Economic Summary

|  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- |
| **Year** |  | 0 | 1 | 2 | 3 |
| **Capital Cost** |  |  |  |  |  |
|  | $ (**218,970,596)** |  |  |  |  |
| **Revenues** |  |  |  |  |  |
| **1652 ton /day Syn gas at $300/ton** |  |  |  |  | $ **104,076,000** |
| **sulfur** |  | $ **-** | $ **-** | $ **-** | $ **3,570,000** |
| **2000 ton of CO2 per day at $ 40/ton** |  | $ **-** | $ **-** | $ **-** | $ **16,800,000** |
|  |  |  |  |  |  |
| **Expenses** |  |  |  |  |  |
| **Equipment cost** |  | $18,247,550 | $91,237,748 | $,742,649 | $,247,550 |
| **Engineering cost** |  | $**3,649,510** | $1,824,755 | $10,948,530 | $1,824,755 |
| **non-engineering cost** |  | $ 3,649,510 | $**10,948,530** | $ 7,299,019.86 |  |
|  |  |  |  |  |  |
| **Loan Expense** | $ (**218,970,596)** | $ - | $ - | $ - | $ 17,517,648 |
| **Utilities:** |  |  |  |  |  |
| **Cooling water** |  | $ | $ | $ | $ 1,158,679 |
| **Electrical** |  | $ | $ | $ | $**10,659,556** |
| **Natural Gas** |  |  |  |  | $ **890,790** |
|  |  |  |  |  |  |
| **Depreciation** |  | $ | $ | $ | $ 12,880,623 |
| **Salaries and Fringes** |  | $ | $ | $ | $ 899,360 |
| **Maintenance** |  |  |  |  |  |
| **3% of cap cost** |  | $ | $ | $ | $6,569,118 |
| **Raw Materials** |  | $ | $ | $ | $87,027,191 |
|  |  |  |  |  |  |
| **Total Expenses** |  | $ 25,546,569 | $ 104,011,033 | $ 72,990,199 | $ 137,602,965 |
| **Income before Taxes** |  | $ | $ | $ | $ (**13,156,965)** |
| **Taxes, 40%** |  | $ | $ | $ | $ (5,262,786) |
| **Income After Taxes** |  | $ - | $ - | $ - | $ (7,894,179) |
| **Add Back Depreciation** |  | $ - | $ - | $ - | $ 12,880,623 |
| **Cash Flow From Operations** |  | $ (**25,546,569)** | $ (**104,011,033)** | $ (72,990,199) | $ 4,986,444 |
| **Cumulative Cash Flow** |  | $ (**25,546,569)** | $ (**129,557,602)** | $ (**202,547,801)** | $ (**197,561,357)** |

|  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- |
| **Year** | 4 | 5 | 6 | 7 | 8 |
| **Revenues** |  |  |  |  |  |
| **1652 ton /day Syn gas at $300/ton** | $ 178,663,800 | $ 184,023,714 | $ 189,544,425 | $ 195,230,758 | $ 201,087,681 |
| **sulfur** | $ 5,950,000 | $ 6,128,500 | $ 6,312,355 | $ 6,501,726 | $ 6,696,777 |
| **2000 ton of CO2 per day at $ 40/ton** | $ 28,000,000 | $ 28,840,000 | $ 29,705,200 | $ 30,596,356 | $ 31,514,247 |
|  |  |  |  |  |  |
| **Loan Expense** | $ 17,094,988 | $ 16,638,516 | $ 16,145,527 | $ 15,613,098 | $ 15,038,074 |
| **Utilities:** |  |  |  |  |  |
| **Cooling water** | $ 1,931,132 | $ 1,989,066 | $ 2,048,738 | $ 2,110,200 | $ 2,173,506 |
| **Electrical** | $ 18,298,904 | $ 18,847,872 | $ 19,413,308 | $ 19,995,707 | $ 20,595,578 |
| **Natural Gas** | $ 1,529,190 | $ 1,575,065 | $ 1,622,317 | $ 1,670,987 | $ 1,721,116 |
|  |  |  |  |  |  |
| **Depreciation** | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 |
| **Salaries and Fringes** | $ 926,341 | $ 954,131 | $ 982,755 | $ 1,012,238 | $ 1,042,605 |
| **Maintenance** |  |  |  |  |  |
| **3% of cap cost** | $ 11,276,986 | $ 11,615,295 | $ 11,963,754 | $ 12,322,667 | $ 12,692,347 |
| **Raw Materials** | $ 127,423,345 | $ 131,246,045 | $ 135,183,426 | $ 139,238,929 | $ 143,416,097 |
|  |  |  |  |  |  |
| **Total Expenses** | $ 191,361,508 | $ 195,746,613 | $ 200,240,447 | $ 204,844,447 | $ 209,559,946 |
|  |  |  |  |  |  |
| **Income before Taxes** | $ 21,252,292 | $ 23,245,601 | $ 25,321,533 | $ 27,484,392 | $ 29,738,759 |
| **Taxes, 40%** | $ 8,500,917 | $ 9,298,240 | $ 10,128,613 | $ 10,993,757 | $ 11,895,504 |
| **Income After Taxes** | $ 12,751,375 | $ 13,947,361 | $ 15,192,920 | $ 16,490,635 | $ 17,843,255 |
| **Add Back Depreciation** | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 |
|  |  |  |  |  |  |
| **Cash Flow From Operations** | $ 25,631,998 | $ 26,827,984 | $ 28,073,543 | $ 29,371,259 | $ 30,723,879 |
| **Cumulative Cash Flow** | $ (171,929,358) | $ (145,101,374) | $ (117,027,831) | $ (87,656,573) | $ (56,932,694) |

Table A7.1.5: Economic Analysis

|  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- |
| **Year** | 9 | 10 | 11 | 12 | 13 |
| **Capital Cost** |  |  |  |  |  |
|  |  |  |  |  |  |
| **Revenues** |  |  |  |  |  |
| **1652 ton /day Syn gas at $300/ton** | $ 207,120,311 | $ 213,333,921 | $ 219,733,938 | $ 226,325,956 | $ 233,115,735 |
| **sulfur** | $ 6,897,681 | $ 7,104,611 | $ 7,317,749 | $ 7,537,282 | $ 7,763,400 |
| **2000 ton of CO2 per day at $ 40/ton** | $ 32,459,674 | $ 33,433,464 | $ 34,436,468 | $ 35,469,562 | $ 36,533,649 |
| **Loan Expense** | $ 14,417,049 | $ 13,746,342 | $ 13,021,978 | $ 12,239,665 | $ 11,394,768 |
| **Utilities:** |  |  |  |  |  |
| **Cooling water** | $ 2,238,711 | $ 2,305,872 | $ 2,375,048 | $ 2,446,300 | $ 2,519,689 |
| **Electrical** | $ 21,213,445 | $ 21,849,849 | $ 22,505,344 | $ 23,180,505 | $ 23,875,920 |
| **Natural Gas** | $ 1,772,750 | $ 1,825,932 | $ 1,880,710 | $ 1,937,132 | $ 1,995,245 |
|  |  |  |  |  |  |
| **Depreciation** | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 |
| **Salaries and Fringes** | $ 1,073,883 | $ 1,106,099 | $ 1,139,282 | $ 1,173,461 | $ 1,208,665 |
| **Maintenance** |  |  |  |  |  |
| **3% of cap cost** | $ 13,073,117 | $ 13,465,311 | $ 13,869,270 | $ 14,285,348 | $ 14,713,909 |
| **Raw Materials** | $ 147,718,580 | $ 152,150,137 | $ 156,714,641 | $ 161,416,081 | $ 166,258,563 |
|  |  |  |  |  |  |
| **Total Expenses** | $ 214,388,158 | $ 219,330,166 | $ 224,386,898 | $ 229,559,114 | $ 234,847,381 |
|  |  |  |  |  |  |
| **Income before Taxes** | $ 32,089,508 | $ 34,541,830 | $ 37,101,258 | $ 39,773,687 | $ 42,565,404 |
| **Taxes, 40%** | $ 12,835,803 | $ 13,816,732 | $ 14,840,503 | $ 15,909,475 | $ 17,026,162 |
| **Income After Taxes** | $ 19,253,705 | $ 20,725,098 | $ 22,260,755 | $ 23,864,212 | $ 25,539,242 |
| **Add Back Depreciation** | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 |
|  |  |  |  |  |  |
| **Cash Flow From Operations** | $ 32,134,328 | $ 33,605,722 | $ 35,141,378 | $ 36,744,835 | $ 38,419,866 |
| **Cumulative Cash Flow** | $ (24,798,366) | $ 8,807,356 | $ 43,948,734 | $ 80,693,569 | $ 119,113,435 |

|  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- |
| **Year** | 14 | 15 | 16 | 17 | 18 | 19 |
| **Capital Cost** |  |  |  |  |  |  |
|  |  |  |  |  |  |  |
| **Revenues** |  |  |  |  |  |  |
| **1652 ton /day Syn gas at $300/ton** | $ 240,109,207 | $ 247,312,483 | $ 254,731,858 | $ 262,373,814 | $ 270,245,028 | $ 278,352,379 |
| **sulfur** | $ 7,996,302 | $ 8,236,192 | $ 8,483,277 | $ 8,737,776 | $ 8,999,909 | $ 9,269,906 |
| **2000 ton of CO2 per day at $ 40/ton** | $ 37,629,659 | $ 38,758,548 | $ 39,921,305 | $ 41,118,944 | $ 42,352,512 | $ 43,623,088 |
|  |  |  |  |  |  |  |
| **Expenses** |  |  |  |  |  |  |
| **Loan Expense** | $ 10,482,278 | $ 9,496,789 | $ 8,432,461 | $ 7,282,987 | $ 6,041,555 | $ 4,700,808 |
| **Utilities:** |  |  |  |  |  |  |
| **Cooling water** | $ 2,595,279 | $ 2,673,138 | $ 2,753,332 | $ 2,835,932 | $ 2,921,010 | $ 3,008,640 |
| **Electrical** | $ 24,592,197 | $ 25,329,963 | $ 26,089,862 | $ 26,872,558 | $ 27,678,735 | $ 28,509,097 |
| **Natural Gas** | $ 2,055,103 | $ 2,116,756 | $ 2,180,259 | $ 2,245,666 | $ 2,313,036 | $ 2,382,427 |
|  |  |  |  |  |  |  |
| **Depreciation** | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 |
| **Salaries and Fringes** | $ 1,244,925 | $ 1,282,272 | $ 1,320,740 | $ 1,360,363 | $ 1,401,174 | $ 1,443,209 |
| **Maintenance** |  |  |  |  |  |  |
| **3% of cap cost** | $ 15,155,326 | $ 15,609,986 | $ 16,078,285 | $ 16,560,634 | $ 17,057,453 | $ 17,569,176 |
| **Raw Materials** | $ 171,246,320 | $ 176,383,709 | $ 181,675,221 | $ 187,125,477 | $ 192,739,242 | $ 198,521,419 |
|  |  |  |  |  |  |  |
| **Total Expenses** | $ 240,252,051 | $ 245,773,237 | $ 251,410,783 | $ 257,164,240 | $ 263,032,827 | $ 269,015,400 |
|  |  |  |  |  |  |  |
| **Income before Taxes** | $ 45,483,117 | $ 48,533,987 | $ 51,725,657 | $ 55,066,293 | $ 58,564,622 | $ 62,229,973 |
| **Taxes, 40%** | $ 18,193,247 | $ 19,413,595 | $ 20,690,263 | $ 22,026,517 | $ 23,425,849 | $ 24,891,989 |
| **Income After Taxes** | $ 27,289,870 | $ 29,120,392 | $ 31,035,394 | $ 33,039,776 | $ 35,138,773 | $ 37,337,984 |
| **Add Back Depreciation** | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 | $ 12,880,623 |
|  |  |  |  |  |  |  |
| **Cash Flow From Operations** | $ 40,170,494 | $ 42,001,015 | $ 43,916,017 | $ 45,920,399 | $ 48,019,397 | $ 50,218,607 |
| **Cumulative Cash Flow** | $ 159,283,928 | $ 201,284,944 | $ 245,200,961 | $ 291,121,360 | $ 339,140,757 | $ 389,359,364 |

|  |  |
| --- | --- |
| **NPV** | $284,270,857 |
| **IRR** | 3.96% |
| **interest** | 8% |

### A7.1.4: Sensitivity Analysis

## A7.2: Combined Economics

### A7.2.1: Annual Operating Costs

### A7.2.2: Expected Revenues

|  |  |  |  |
| --- | --- | --- | --- |
| **Products** | **Quantity Produced (ton/yr)** | **Price ($/ton)** | **Annual revenue** |
| **Sulfur** | 35,000 | $ 170 | $5,950,000.00 |
| **Carbon dioxide** | 700,000 | $ 40 | $28,000,000.00 |
| **Propionic Acid** | 831 | $1800 | $ 1,495,800.00 |
| **Industrial AcA** | 7560 | $330 | $ 2,494,800.00 |
| **Glacial AcA a** | 495,600 | $ 1050 | $ 520,380,000.00\* |

(note:\* This revenues will be split in half and will be shared between both team)

### A7.2.3: Combined Economic Summary

|  |  |
| --- | --- |
| NPV | $2,418,301,994 |
| IRR | 22.61% |
| Interest | 8.00% |
| Inflation | 3.00% |

The tables below describes the expense and revenue summary for the plant. This economic analysis is based on the 17 years of the plant life.

|  |  |  |  |  |
| --- | --- | --- | --- | --- |
| Year | 0 | 1 | 2 | 3 |
| **Capital Cost** | 738,306,084.40 |  |  |  |
|  |  |  |  |  |
| **Revenues** |  | 558,320,600.00 | 585,477,818.00 | 613,970,132.54 |
| 831 tons Propionic Acid at $1800/ton |  | 1,495,800.00 | 1,540,674.00 | 1,586,894.22 |
| 7560 tons Industrial AcA at $330/ton |  | 2,494,800.00 | 2,569,644.00 | 2,646,733.32 |
| 495,600 tons Glacial AcA at $1050/ton |  | 520,380,000.00 | 546,399,000.00 | 573,718,950.00 |
| Sulfur | 100 ton/day @ $170 | 5,950,000.00 | 6,128,500.00 | 6,312,355.00 |
| CO2 | 2000/day @ $40 | 28,000,000.00 | 28,840,000.00 | 29,705,200.00 |
| **Expenses** |  |  |  |  |
| Loan Expense |  | 164,313,040.86 | 164,313,040.86 | 164,313,040.86 |
| Utilities |  |  |  |  |
| Steam Generation |  | 16,000,000.00 | 16,480,000.00 | 16,974,400.00 |
| Cooling water |  | 37,000,000.00 | 38,110,000.00 | 39,253,300.00 |
| Electrical |  | 38,000,000.00 | 39,140,000.00 | 40,314,200.00 |
|  |  |  |  |  |
| Sum of Years Depreciation |  | 82,034,009.38 | 77,208,479.41 | 72,382,949.45 |
| Salaries and Fringes |  | 2,842,200.00 | 2,927,466.00 | 3,015,289.98 |
| Maintenance : 3% of cap cost |  | 22,149,182.53 | 22,813,658.01 | 23,498,067.75 |
|  |  |  |  |  |
| Raw Materials |  | 54,000,000.00 | 55,620,000.00 | 57,288,600.00 |
| Total Expenses |  | 416,338,432.77 | 416,612,644.28 | 417,039,848.04 |
| Income before Taxes |  | 141,982,167.23 | 168,865,173.72 | 196,930,284.50 |
| Taxes, 40% |  | 56,792,866.89 | 67,546,069.49 | 78,772,113.80 |
|  |  |  |  |  |
| Income After Taxes |  | 85,189,300.34 | 101,319,104.23 | 118,158,170.70 |
|  |  |  |  |  |
| Add Back Depreciation |  | 82,034,009.38 | 77,208,479.41 | 72,382,949.45 |
|  |  |  |  |  |
| Cash Flow From Operations |  | 167,223,309.72 | 178,527,583.65 | 190,541,120.15 |
|  |  |  |  |  |
| Cumulative Cash Flow | -738,306,084.40 | -571,082,774.68 | -392,555,191.04 | -202,014,070.89 |

|  |  |  |  |  |
| --- | --- | --- | --- | --- |
| Year | 4 | 5 | 6 | 7 |
| **Capital Cost** |  |  |  |  |
| **Revenues** | 643,863,615.52 | 675,227,621.93 | 708,134,953.44 | 742,662,030.03 |
| 831 tons Propionic Acid at $1800/ton | 1,634,501.05 | 1,683,536.08 | 1,734,042.16 | 1,786,063.43 |
| 7560 tons Industrial AcA at $330/ton | 2,726,135.32 | 2,807,919.38 | 2,892,156.96 | 2,978,921.67 |
| 495,600 tons Glacial AcA at $1050/ton | 602,404,897.50 | 632,525,142.38 | 664,151,399.49 | 697,358,969.47 |
| Sulfur | 6,501,725.65 | 6,696,777.42 | 6,897,680.74 | 7,104,611.16 |
| CO2 | 30,596,356.00 | 31,514,246.68 | 32,459,674.08 | 33,433,464.30 |
|  |  |  |  |  |
| **Expenses** |  |  |  |  |
| Loan Expense | 164,313,040.86 | 164,313,040.86 | 164,313,040.86 | 164,313,040.86 |
| Utilities |  |  |  |  |
| Steam Generation | 17,483,632.00 | 18,008,140.96 | 18,548,385.19 | 19,104,836.74 |
| Cooling water | 40,430,899.00 | 41,643,825.97 | 42,893,140.75 | 44,179,934.97 |
| Electrical | 41,523,626.00 | 42,769,334.78 | 44,052,414.82 | 45,373,987.27 |
| Sum of Years Depreciation | 67,557,419.49 | 62,731,889.52 | 57,906,359.56 | 53,080,829.60 |
| Salaries and Fringes | 3,105,748.68 | 3,198,921.14 | 3,294,888.77 | 3,393,735.44 |
| Maintenance : 3% of cap cost | 24,203,009.78 | 24,929,100.07 | 25,676,973.08 | 26,447,282.27 |
| Raw Materials | 59,007,258.00 | 60,777,475.74 | 62,600,800.01 | 64,478,824.01 |
| Total Expenses | 417,624,633.81 | 418,371,729.05 | 419,286,003.04 | 420,372,471.16 |
|  |  |  |  |  |
| Income before Taxes | 226,238,981.71 | 256,855,892.88 | 288,848,950.39 | 322,289,558.87 |
| Taxes, 40% | 90,495,592.68 | 102,742,357.15 | 115,539,580.16 | 128,915,823.55 |
| Income After Taxes | 135,743,389.03 | 154,113,535.73 | 173,309,370.24 | 193,373,735.32 |
| Add Back Depreciation | 67,557,419.49 | 62,731,889.52 | 57,906,359.56 | 53,080,829.60 |
| Cash Flow From Operations | 203,300,808.51 | 216,845,425.26 | 231,215,729.80 | 246,454,564.92 |
| Cumulative Cash Flow | 1,286,737.63 | 218,132,162.88 | 449,347,892.68 | 695,802,457.60 |

|  |  |  |  |  |
| --- | --- | --- | --- | --- |
| Year | 8 | 9 | 10 | 11 |
| **Capital Cost** |  |  |  |  |
| **Revenues** | 778,889,070.32 | 816,900,280.79 | 856,784,054.49 | 898,633,179.66 |
| 831 tons Propionic Acid at $1800/ton | 1,839,645.33 | 1,894,834.69 | 1,951,679.73 | 2,010,230.12 |
| 7560 tons Industrial AcA at $330/ton | 3,068,289.32 | 3,160,338.00 | 3,255,148.14 | 3,352,802.58 |
| 495,600 tons Glacial AcA at $1050/ton | 732,226,917.94 | 768,838,263.84 | 807,280,177.03 | 847,644,185.88 |
| Sulfur | 7,317,749.50 | 7,537,281.98 | 7,763,400.44 | 7,996,302.46 |
| CO2 | 34,436,468.23 | 35,469,562.28 | 36,533,649.15 | 37,629,658.62 |
|  |  |  |  |  |
| **Expenses** |  |  |  |  |
| Loan Expense | 164,313,040.86 | 164,313,040.86 | 164,313,040.86 | 0.00 |
| Utilities |  |  |  |  |
| Steam Generation | 19,677,981.85 | 20,268,321.30 | 20,876,370.94 | 21,502,662.07 |
| Cooling water | 45,505,333.02 | 46,870,493.01 | 48,276,607.80 | 49,724,906.04 |
| Electrical | 46,735,206.89 | 48,137,263.09 | 49,581,380.99 | 51,068,822.42 |
| Sum of Years Depreciation | 48,255,299.63 | 43,429,769.67 | 38,604,239.71 | 33,778,709.74 |
| Salaries and Fringes | 3,495,547.50 | 3,600,413.93 | 3,708,426.34 | 3,819,679.13 |
| Maintenance : 3% of cap cost | 27,240,700.74 | 28,057,921.76 | 28,899,659.41 | 29,766,649.19 |
| Raw Materials | 66,413,188.73 | 68,405,584.39 | 70,457,751.93 | 72,571,484.48 |
| Total Expenses | 421,636,299.22 | 423,082,808.01 | 424,717,477.98 | 262,232,913.08 |
|  |  |  |  |  |
| Income before Taxes | 357,252,771.10 | 393,817,472.77 | 432,066,576.51 | 636,400,266.59 |
| Taxes, 40% | 142,901,108.44 | 157,526,989.11 | 172,826,630.61 | 254,560,106.64 |
| Income After Taxes | 214,351,662.66 | 236,290,483.66 | 259,239,945.91 | 381,840,159.95 |
| Add Back Depreciation | 48,255,299.63 | 43,429,769.67 | 38,604,239.71 | 33,778,709.74 |
|  |  |  |  |  |
| Cash Flow From Operations | 262,606,962.30 | 279,720,253.34 | 297,844,185.62 | 415,618,869.70 |
|  |  |  |  |  |
| Cumulative Cash Flow | 958,409,419.90 | 1,238,129,673.23 | 1,535,973,858.85 | 1,951,592,728.54 |

|  |  |  |  |  |
| --- | --- | --- | --- | --- |
| Year | 12 | 13 | 14 | 15 |
| **Capital Cost** |  |  |  |  |
|  |  |  |  |  |
| **Revenues** | 942,545,058.77 | 988,621,938.44 | 1,036,971,150.89 | 1,087,705,367.43 |
| 831 tons Propionic Acid at $1800/ton | 2,070,537.02 | 2,132,653.13 | 2,196,632.73 | 2,262,531.71 |
| 7560 tons Industrial AcA at $330/ton | 3,453,386.66 | 3,556,988.26 | 3,663,697.91 | 3,773,608.85 |
| 495,600 tons Glacial AcA at $1050/ton | 890,026,395.18 | 934,527,714.94 | 981,254,100.68 | 1,030,316,805.72 |
| Sulfur | 8,236,191.53 | 8,483,277.28 | 8,737,775.60 | 8,999,908.86 |
| CO2 | 38,758,548.38 | 39,921,304.83 | 41,118,943.98 | 42,352,512.30 |
|  |  |  |  |  |
| **Expenses** |  |  |  |  |
| Loan Expense | 0.00 | 0.00 | 0.00 | 0.00 |
| Utilities |  |  |  |  |
| Steam Generation | 22,147,741.93 | 22,812,174.19 | 23,496,539.42 | 24,201,435.60 |
| Cooling water | 51,216,653.22 | 52,753,152.81 | 54,335,747.40 | 55,965,819.82 |
| Electrical | 52,600,887.09 | 54,178,913.70 | 55,804,281.11 | 57,478,409.54 |
| Sum of Years Depreciation | 28,953,179.78 | 24,127,649.82 | 19,302,119.85 | 14,476,589.89 |
| Salaries and Fringes | 3,934,269.51 | 4,052,297.59 | 4,173,866.52 | 4,299,082.52 |
| Maintenance : 3% of cap cost | 30,659,648.67 | 31,579,438.13 | 32,526,821.27 | 33,502,625.91 |
| Raw Materials | 74,748,629.02 | 76,991,087.89 | 79,300,820.53 | 81,679,845.14 |
| Total Expenses | 264,261,009.21 | 266,494,714.13 | 268,940,196.10 | 271,603,808.42 |
|  |  |  |  |  |
| Income before Taxes | 678,284,049.56 | 722,127,224.31 | 768,030,954.79 | 816,101,559.01 |
| Taxes, 40% | 271,313,619.82 | 288,850,889.72 | 307,212,381.92 | 326,440,623.60 |
| Income After Taxes | 406,970,429.74 | 433,276,334.58 | 460,818,572.88 | 489,660,935.41 |
| Add Back Depreciation | 28,953,179.78 | 24,127,649.82 | 19,302,119.85 | 14,476,589.89 |
| Cash Flow From Operations | 435,923,609.52 | 457,403,984.40 | 480,120,692.73 | 504,137,525.30 |
| Cumulative Cash Flow | 2,387,516,338.06 | 2,844,920,322.46 | 3,325,041,015.19 | 3,829,178,540.49 |

|  |  |  |
| --- | --- | --- |
| Year | 16 | 17 |
| **Capital Cost** |  |  |
| **Revenues** | 1,140,942,864.57 | 1,196,807,803.43 |
| 831 tons Propionic Acid at $1800/ton | 2,330,407.66 | 2,400,319.89 |
| 7560 tons Industrial AcA at $330/ton | 3,886,817.11 | 4,003,421.62 |
| 495,600 tons Glacial AcA at $1050/ton | 1,081,832,646.00 | 1,135,924,278.30 |
| Sulfur | 9,269,906.13 | 9,548,003.31 |
| CO2 | 43,623,087.66 | 44,931,780.29 |
|  |  |  |
| **Expenses** |  |  |
| Loan Expense | 0.00 | 0.00 |
| Utilities |  |  |
| Steam Generation | 24,927,478.67 | 25,675,303.03 |
| Cooling water | 57,644,794.41 | 59,374,138.25 |
| Electrical | 59,202,761.83 | 60,978,844.69 |
| Sum of Years Depreciation | 9,651,059.93 | 4,825,529.96 |
| Salaries and Fringes | 4,428,054.99 | 4,560,896.64 |
| Maintenance : 3% of cap cost | 34,507,704.69 | 35,542,935.83 |
| Raw Materials | 84,130,240.50 | 86,654,147.71 |
| Total Expenses | 274,492,095.01 | 277,611,796.10 |
|  |  |  |
| Income before Taxes | 866,450,769.55 | 919,196,007.32 |
|  |  |  |
| Taxes, 40% | 346,580,307.82 | 367,678,402.93 |
|  |  |  |
| Income After Taxes | 519,870,461.73 | 551,517,604.39 |
|  |  |  |
| Add Back Depreciation | 9,651,059.93 | 4,825,529.96 |
|  |  |  |
| Cash Flow From Operations | 529,521,521.66 | 556,343,134.36 |
|  |  |  |
| Cumulative Cash Flow | 4,358,700,062.15 | 4,915,043,196.50 |

### A7.2.4: Sensitivity Analysis

# Appendix 8: Utilities

# Appendix 9: Conceptual Control Scheme

# Appendix 10: General Arrangement – Major Equipment Layout

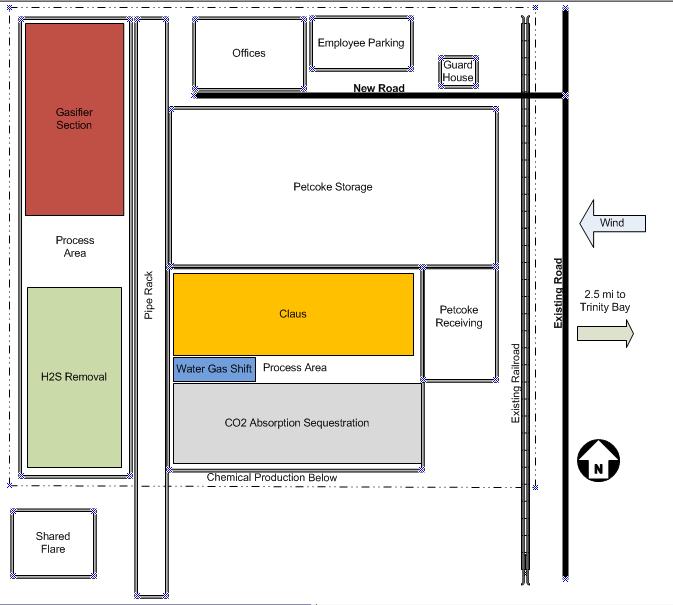


Figure A10.1: Overall Plant Layout

# Appendix 11: Distribution and End-Use Issue Review

# Appendix 12: Constraints Review

# Appendix 13: Applicable Standards

# Appendix 14: Project Communications File

## A14.1: Timeline of Events

01/14/11: Groups finalized and project start.  
  
01/18/11: First Group Meeting.  
  
01/20/11: First Mentor Meeting - Chose Petroleum Coke as feedstock.  
  
01/21/11: Group Meeting - Assigned and worked on PowerPoint #1 sections.  
  
01/24/11: Mentor Meeting - Discussed and ran through PowerPoint #1.  
  
01/25/11: First Presentation Given - Briefly spoke with chem. production group on details and scale up.  
  
01/26/11: Group Meeting - Assigned parts for next presentation. Decided on Shell membrane wall gasifier.  
  
01/27/11: Brief Leader Meeting with Group Golf. Talked about specifics for what each other needs.  
  
01/28/11: Group Meeting - Discussed gasifier more. Went over preliminary mass balance.  
  
02/01/11: Group Meeting - Discussed gasifier and looked at process flow.  
Mentor Meeting - Teleconference - Discussed assigning everyone a specific section of the overall process to work on.  
  
02/04/11: Group Meeting - Checked in to see how the research and calculations were going.   
  
02/08/11: Group Meeting - Discussed individual developments.   
Mentor Meeting - Discussed scope of the project, where to place focus, determined more action items, ans what is expected for the next presentation.  
  
02/09/11: Group Meeting - Worked on flow sheets, econ, and discussed some of the overall process concerns.  
  
02/11/11: Group Meeting - Individual work for PowerPoint #2  
  
02/14/11: Group Meeting - Worked on cleaning up PowerPoint #2  
Mentor Meeting - Went over PowerPoint #2, determined areas in need of repair  
  
02/15/11: Presentation #2  
  
02/17/11: Group Meeting - Discussed what needs to be done for the next presentation  
  
02/18/11: Group Meeting - Worked on individual parts, and Aspen sim  
  
02/21/11: Group Meeting - Worked on correcting the abstract  
  
02/22/11: Group Meeting - Discussed catalysts in the system  
Mentor Meeting - Looked over abstract and determined sections in need of improvement  
  
02/23/11: Group Meeting - Individual work  
  
02/25/11: Group Meeting - Process clarification, Aspen modeling  
  
02/28/11: Group Meeting - Individual work  
  
03/01/11: Group Meeting - Worked on PowerPoint and Aspen  
  
03/02/11: Mentor Meeting - Presented work  
  
03/03/11: Group Meeting - Worked on slides  
  
03/04/11: Group Meeting - Aspen work and slides  
  
03/07/11: Mentor Meeting - Went over PowerPoint #3  
  
03/08/11: Presentation #3  
  
03/11/11: Group Meeting - Discussed group merging possibility and land costs  
  
03/15/11: Mentor Meeting - Discussed what needs to be done to finalize the project  
  
03/16/11: Group Meeting - Talked about plant layout and poster  
  
03/17/11: Joint Team Meeting - Merged economy with Team Golf  
  
03/19/11: Group Meeting - Worked on organization of individual parts  
  
03/28/11: Group Meeting - Worked on economics, PFD, control scheme, and report  
  
03/29/11: Group Meeting - Individual work  
Meeting with Team Golf - Discussed combined economics  
Mentor Meeting - Discussed what needs to be done for the next presentation and the poster

03/31/11: Group Meeting - Worked on catalysts in the process and the report

## A14.2: Company Contacts

UIG Response to Team Hotel:

|  |  |  |  |
| --- | --- | --- | --- |
| |  | | --- | | https://mail.google.com/mail/images/cleardot.gifhttps://mail.google.com/mail/images/cleardot.gif  Walt Dwarnick   to me | |  |  |

Lipi,  
  
This size plant would be in the US $ 100 million range.  It would also  
require about 40 megawatts of power to operate.  
  
Good luck with your project  
  
Best Regards,  Walt  
  
Walter Dwarnick  
Universal Industrial Gases  
Sales and Marketing Manager  
Office: 610-515-8585  
Mobile: 484-894-4262  
Website: [www.uigi.com](http://www.uigi.com/)  
  
-----Original Message-----  
From: Penny Kornet [mailto:[olil@uigi.com](mailto:olil@uigi.com)] On Behalf Of [pkornet@uigi.com](mailto:pkornet@uigi.com)  
Sent: Monday, February 28, 2011 9:08 AM  
To: Walt Dwarnick  
Subject: FW: Yahoo! WebHosting Email

I am a senior student in Chemical Engineering field at UIC, IL

We are working on a project for Senior Design Expo.

We are using Shell Coal gasifier as our role model to gasify petroleum coke in order to achieve pure synthesis gas.

As shell Gasifier is an Entrained bed, it requires 95% pure Oxygen for the reaction.

Part of our project is to find out the cost of this AIR SEPARATION UNIT PLANT.

I was wondering if you can help me with my project by giving me an estimate of how much it would cost us to install an CRYOGENIC AIR SEPARATION OXYGEN PLANT that generates 3000 metric tons/day 95% pure Oxygen continuously every day.

We value your time and privacy.

We only need an estimate as a part of our project.

We appreciate your response and help.

* Lipi Vahanwala  
  Senior Design Group, Hotel  
  Undergraduate Chemical Engineering Student at UIC, IL

|  |  |  |  |  |
| --- | --- | --- | --- | --- |
| |  | | --- | | https://mail.google.com/mail/images/cleardot.gif  **I made contact with KFO to find out the cost for Zinc Oxide catalyst. Export Manager got back to me with the response below:**  [**sale@global-global.com**](mailto:sale@global-global.com) to me | | |  |  |
|  |  | | |  |

Dear Lipi Vahanwala,

Thank you for your email!

We would quote our best favourable price for Zinc oxide desulfurizar TS-308:

90% 4mm--------------FOB China USD18/kg

Package: 35kg/drum

Our product can meet your need.

Additionally, please let me spend a little time to introduce our company. KFO Japan Co. Ltd. is biotech laboratories founded inJapan in 2001 and China in 2003, KFO focuses on research and development of chemical & nutritional technologies. Our manufacturing and quality in China are supervised by Japan.

We hope you will find us as your reliable partner with consistent quality + competitive price.

I 'm looking forward to your reply!

Thank you in advance and best wishes,

Mr. Daniel Lee    (Export Manager)

Ms. Molly         (Assistant)

KFO Japan Co. Ltd.

Tel: +86 152 8026 8510

Fax: +86 592 376 1310

[sale@global-global.com](mailto:sale@global-global.com)

http://www.[catalyst-catalyst.net](http://catalyst-catalyst.net/" \t "_blank)

NOTE: We will reply your email within 24 hours, if you can't receive our emails in time, it would be the problem of the email system, please resend email to bk.technology@hotmail.com

----- Original Message -----

**From:** [Lipi](mailto:lipiv98@gmail.com)

**To:** [sales@catalyst-catalyst.net](mailto:sales@catalyst-catalyst.net)

**CC:**

**Sent:** 2011-02-26 07:14

**Subject:** Re:price for Zinc oxide desulfurizar TS-308

|  |
| --- |
| * I am currently a Chemical Engineering student at University of Illinois at Chicago. * We are currently using Shell Coal Gasifier (SCGP) as a role model to produce syn gas as a part of my project. * Since we are using petroleum coke as our feed-stock which contains sulfur, the part of our project is to get rid of that sulfur using Zinc-Oxide. * I have read the information provided on your company’s website and it seems to me that Zinc Oxide desulfurizer TS-308 is a better fit for our need. * However, we operating sulfur removal operation at 750 F and 437 psi which is relatively higher in pressure and temperature.      * I was wondering if you can help me out with my project by giving me an estimate of how much would it cost to purchase 10 lbs/day of Zinc Oxide desulfurizer that is capable of removing sulfur from the raw syn-gas at approximately 750 F and 437 psi. |

**To Exxon, Asbury, and Oxbow**

Hello,

My name is Russell and I am a chemical engineering student at University

of Illinois at Chicago. Some classmates and I are working on a senior

design project and would like to look into your Petroleum Coke product.

Listed below are several questions that I have regarding your product.

What type/grade of petroleum coke do you offer?

What size supply can you satisfy?

What is the composition?

Pricing?

(No Response from Exxon or Oxbow)

**2/8/11: To Asbury**

Hello Albert,

Thank you for your response. As part of my senior design class my group

and I are trying to design a a gassification process that will yield

syngas from petroleum coke. We need a calcinated coke that is low in

sulfur and as fine as possible. We will need about 2000 tons per day. If

you can get an approximate price that would be great.

Also, if you could please provide a general overview of the coke. Any sort

of knowledge on the coke would be fine. For instance: what is the end

sulfur conent (i.e. sulfur oxide output), hardness of the coke,

approximate and ultimate composition.

I don't know if you would be able to answer this question but would you

also happen to know what the general cost of a gasifier (or even a

gasification plant) would cost?

Any other sort of information that you feel would be useful would also be

much appreciated.

Thanks

**2/8/11: From Asbury**

Russel,

Please define "low" sulfur. In other words, are what is the maximum

amount of sulfur your syngas unit can handle, or what is the maximum

allowable S content allowed in the coke?

AVT

Albert V. Tamashausky

**2/8/11: To Asbury**

Albert,

At the end of the process we will only be allowed to have 5 ppm sulfur

in our syngas. This is probably unattainable without some sort of extra

sulfur removal step. If possible anything lower than 6 percent will

suffice.

Russell Cabral

**2/11/11: From Asbury**

Hello Russell,

6% max is good because it opens the door to some of the lower cost

petroleum coke materials.

2000 tons/day of any single coke is quite a bit. Do you mean 200 tons

a day? A standard coke barge only holds 1500 tons.

Please double check the amount. I'm not sure anyone could supply that

amount of the same coke.

AVT

Albert V. Tamashausky

**2/14/11: To Asbury**

Albert,

It is supposed to be 2000 tons, but 1500 tons will also suffice. Is it

possible to get any rate?

-Russell

**2/14/11: From Asbury**

Russell,

Assume about $75-$95/ton. This is for green (uncalcined ) coke. This is

a very volatile material, price wise, at this time. Last year this same

product was about 1/2 the cost. Higher demand, as in your application,

will probably up the price.

Keep in mind that 1500 tons/day is all or most of the total output from

a single, larger refinery. I assume you would be installing your unit

in the refinery to save shipping cost and to take advantage of their

handling and distribution system.

AVT

Albert V. Tamashausky

**To Conocophillips**

I am a chemical engineering student at the University of Illinois at Chicago and I have a few questions regarding your petroleum coke refinery in Sweany, Texas.

1) How much petroleum coke do use annually?

2) Who supplies the petroleum coke?

3) What is your energy output?

4) What was the start up cost of the refinery?

Any input would be much appreciated

- Russell

(No Response)

**To ZEEP (Building Gasification Unit in Beaumont, TX)**

Hello,

I'm a student in the chemical engineering department at the University of Illinois at Chicago and for my senior design project my classmates and I will be designing a gasification plant using petcoke. I read that you will be building a gasification unit using petcoke in Beaumont, Texas. I was wondering if you could answer a few questions of mine.

What suppliers do you get your petroleum coke from?

What is the composition of the petcoke you will be using?

Were there any sites that you were looking into before you settled on Beaumont? and what were your qualifications?

How do you plan on transporting your final product?

Are you shipping in or making the gases needed for your operation (i.e. oxygen needed for gasification)?

Did you require any steps for sulfur removal? If so what were they?

What type of gasifier will you be using?

I realize that you will not be able to answer some of the questions but any input would be much appreciated.

(No Response)

**To Port of Victoria**

Tony Rigdon   
Executive Director  
tonyrigdon@portofvictoria.com

Hello,

I am a chemical engineering student at the University of Illinois at Chicago and for my senior design project my classmates and I will be designing a gasification plant. For our theoretical plant we will need about 200 acres of land and I was wondering if you would be able to give a ballpark estimate to what that would cost. We would need access to the railways and waterways.

Any input on the matter would be much appreciated.

(No Response)

Hi Mr. Kanyuh,

My name is Ryan Kosak and I am with the senior design group Hotel. I was wondering if I could get any information on Selexol, mainly just the price. Mr. Keesom also mentioned UOP was a good source for the Zinc Oxide catalyst as well. Again I am just really looking for the price so if you or someone else with UOP could help it would be greatly appreciated. Thank you.

-Ryan Kosak

Ryan,

       A good budgetary price for Selexol is $3.20 / lb.  For the Zinc Oxide catalyst, what service are you looking to use that for since we have a number of catalysts in our portfolio?  If you let me know what the service is I can check and see if I can get a budgetary price for you.

Regards,

**Adam Kanyuh**

Hi Mr. Kanyuh,

Thank you for the Selexol pricing. As for the zinc oxide its purpose in our process is to be a safety net and catch any of the remaining sulfur in our syngas stream. It takes place after our H2S absorption system and before the syngas goes through our water gas shift and CO2 absorption. At that point there should be barely any sulfur components in the stream. Thanks again for your help.

-Ryan Kosak

Ryan,

      Attached is a UOP adsorbent which is used for H2S removal from hydrocarbon streams.     A good budgetary price to use is $55 / lb.  I hope that helps.

Regards,

**Adam Kanyuh**

Dear Mr. Kanyuh,

Thank you very much for this.

-Ryan Kosak

Lipi

SELEXOL is a physical solvent.

   IT only works well at relatively high pressures.  
   and, it will need to be refrigerated to work well.

    I do modelling for SELEXOL plants, and I’m just going to give you some general information.

1)      I’m assuming you are operating at a pressure of 500 psig at least

2)      The CO2 amount in the gas to be treated is likely 30 vol%. So, the partial pressure is about 145 psia.

3)      For 200 Tonnes per day, that is 440K lbs or 10,000 lbmol/D or 7 lbmol/min.  
   If we assume a rich loading of 1:1, then you’d need 7lbmol/min of SELEXOL to be circulated.  
    at 280 molecular wt, this is 1960 lbs a minute or 228 usg/min.

4)      I’ll assume that you only need to get to about 3% CO2 in the sweetened gas.  
In that case you will need in the flow scheme.  
- a contactor with about 10 theorectical stages. This will contain about 80 ft of IMTP 40 SS packing.  
   My guess is, (and its only a guess), the tower will be about 4 ft diameter.  
- 1st flash tank:  you have to design it to run 50% full, residence time of say 5 minutes.  
       pressure in the flash tank will be about 250 psig.  
       the flash gas will be recycled back to the contactor to save much of the Co and H2 coabsorbed with teh CO2.

-        2nd flash tank:  will run close to atmospheric pressure, again 5 minutes residence time, run half full.  
Flash gas will likely have to go to incineration to ensure burning the remaining little bit of CO (nasty stuff

-        3rd flash tank: Vacuum flash, say about 9 psia. (so you need a vac pump, not big, but you need one)

-        Chiller: You will need a propane chiller in the loop to provide enough chilling to run the SELEXOL into the contactor  
at about 32 F. This won’t be large, but you still need this chiller.

-        Pumps: One small pump to suck off the vac flash and pump to say 50 psig to get through the chiller, then a high  
pressure pump to get SEELXOL to top of the contactor

-        A small “dehydrator” to remove excess water from the feedgas. (5% slipstream, tiny tower, heater at bottom to heat   
to 300 F, little condenser at top to cool to about 220 F so water escapes, but not SELEXOL)

5)      SELEXOL initial fill (you’ll use it at close to 100% (no water). I’d estimate you’ll need minimum 30 minutes of circulating material  
    so, 6840 gallons of solvent. ($150K)

I really can’t help too much with capital estimates. Its not my thing. However, if I was guessing, you’re likely looking at $3 million at least.

I’d suggest that you must have access to something to simulate this. Eg, PROMAX, ASPEN, HYSIS, etc. While those packages aren’t perfect,

     At least they would confirm some of my estimates.

I’m amazed that whatever you are gasifying has no sulphur present. Usually, H2S and COS are issues and more equipment is needed.  
   If you have to get down to < 3% CO2 in the finished gas, you have to go to using a thermal regenerator. (heating it up). This adds lots more expense and equipment.

**Jack McJannett**   
**Dow Oil & Gas**

I am a senior student in Chemical Engineering field at UIC, IL

We are working on a project for Senior Design Expo.

We are using Shell Coal gasifier as our role model to gasify petroleum coke in order to achieve pure synthesis gas.

Our goal is remove 200 t/d Carbon Dioxide from the syn gas using Selexol.

Part of our project is to find out all the equipment required to remove 200 t/d CO2 and the cost of the overall CO2 removal process.

I was wondering if you can help me with my project by giving me an estimate of the equipment required in order to remove CO2 from the syngas and their quantity. Also, I would like to know how much the overall process will cost us. (With or without Selexol)

We value your time and privacy.

We only need an estimate as a part of our project.

We appreciate your response and help.

* Lipi Vahanwala  
  Senior Design Group, Hotel  
  Undergraduate Chemical Engineering Student at UIC, IL

## A14.3: Mentor Notes

### A14.3.1: Individual Mentor Meetings

**Notes from teleconference (2/1/11)**

Attendance: Ryan, Russell, Lipi, Vijeta

Assign 1 person to each major part of the process – **Action Item**

* Gasifier – Lipi
* Gas Clean Up – Ryan
* Water Gas Shift – Tom
* ASU, Oxygen – Russel
* Materials and Energy Balance – Vijeta

Look into downstream aspects; removals, temperatures, and pressures

Regenerate the system

Understand the lock hoppers in the gasifier

Get O2 from a third party – **Action Item**

N2 can be used further, sent to a compressor or turbine for power generation

H2S in syngas, look up how to remove it

Material balance, add moles/hour to the data section in process flow data

Follow each component (C,H,etc) through all the steps

Look into the rates of the reaction in the gasifier (calculate)

Determine the engineering focus aka how deep do we need to look into each process – **Action Item**

Ask Dr. Perl about how far we need to go

**Mentor Notes 2/8/11**

* Companies to look into for O2 – Air Products, Lindi, Boc
* We need to talk to the Acetic Acid group more often, we need to look into their methanol production
* We need to know their pressures and temperatures required for the methanol synthesis
* Set up meeting with AA group (Planning on meeting 2/10/11)
* Determine price of CO and H2
* Factor equipment costs
* The gasifier is built in a discrete size (find the sizing)
* Look into selling off extra syngas
* Heat recovery and steam production will need to be looked into
* Why did we choose this gasifier (1 slide in the upcoming presentation)
* Flow sheets
* Mass balance
* Mole balance
* Volumetric balance (maybe)
* Use the approximate analysis in calculations

**Mentor Notes 2/14/11**

* Fix slide 5 (capitization)
* Slide 11 CO2 in WGS
* Add Superclaus
* Make the chem. production group take our specifications
* Add ZnO/CuO cleaning step to remove trace sulfur
* Move slide 14 and 15
* Fix slide with mat. balance, C balance incorrect
* Explain why mat balance (N, S) did not balance, what will be done
* Change slide 16 to pound moles
* \*How much steam can be generated from the WGS energy?
* What do you use the steam for?
* Look into WGS reactor designs (sizing)
* Set operating temperatures and pressures
* Remove slides 19 and 21
* Add schematic of gasifier (Lipi)
* Use more of a block flow
* Place a block flow in the beginning
* Add tons petcoke in, tons syngas out, etc
* Restate who we are and what we are doing
* Mention objective to supply syngas to chem. production group at T,P, comp
* Move flow sheets before gasifier
* \*\*Ask yourself what is important about this slide
* Rethink objectives
* Move quench row in table to the top to show more importance
* Possibly highlight section on block flow then talk about it to show the listener what exactly being talked about
* Resize flow sheet
* Mention that CO2 is a work in progress
* Add temperatures to the cleaning section

**Mentor Meeting 03/15/11**

* Add discussion on where the indices come from for the econ (something that should be put the report
* Find the engineering cost
* Look into labor costs for 3 year plant start (might be included in direct cost)
* Plant construction refers to sewers, roads, land, etc
* Land cost for Texas
* Combining labor, building, and other costs that can be merged (cost effective)
* DECIDE on Location: Victoria or Pasadena
* Initial control scheme
* Excel
* Pressure and Temperature monitors
* Process streams on one sheet
* Mass and Moles
* Selectivity work
* Pick example for cost estimating for the next presentation
* Tail gas treatment from Claus
* POSTER
* CO2 recovery
* Clean products from dirty feed
* Theme (Catchy)
* Story
* Good Layout
* BFD with numbers
* Abstract
* Challenges
* BFD/PFD
* Petcoke to chemical
* Petcoke (normally for power)
* Low value feedstock
* High value product
* Mention ASPEN simulation for material and energy balance
* Equipment sizing
* Econ
* Assumptions
* Table or graph of income
* Sensitivity of pricing
* Cash flow vs. time
* NPV and IRR

**Mentor Notes: 3/2/11**

* Send coke through all the paces of the process as a check on Aspen
* ASU – verify cost and figure out if the plant is being purchased or is just the O2
* Is it more cost effective to set up ASU or by over the fence
* Add description of where the prices came from in the economy excel sheet
* Learn the Look Up Index function in excel
* Look into Alberta, Canada for cheaper petcoke
* Adjust cost of syngas with the heating value of natural gas
* Determine a fair syngas price
* Assume 3 years to build the plant
* Look into better sources for ZnO
* For presentation #3 talk about…
* Econ issues
* CO2 footprint
* Price of petcoke being too high
* Eastman Chemical Company (Tennessee)
* Chem Production from petcoke
* Cover some of the holes from last presentation
* Tell a story on heating and cooling
* Use blocks to show
* Heat out of gasifier
* Start with flow diagram then go more in-depth
* Look in Higman for carrier gas solution

**Mentor Meeting 03/29/11**

* Sharing cooling water tower
* Make sure total econ is up to date
* Air price, settle on a company
* Sort out price of ZnO, go with UOP or the other
* Present econ separate from the overall presentation
* Make a note of the CO2 footprint with and without CO2 capture (for presentation)
* Share buildings and services with Golf
* Focus on Water Gas Shift for control scheme
* Online GC
* Control philosophy
* flow
* level
* pressure
* temperature
* composition
* Probably will be asked about gasifier controls
* batch control
* pressure
* Come up with a catchier sub title
* “Weaving gold out of straw”
* Move bullet points of overview to abstract section and reduce abstract
* Discuss the problem of valuing a intermediate product
* Annualized cost with and without CO2 capture
* Reformat printouts for the stream flows
* Print out sheets for the mentors
* Slag 🡪 Clinker 🡪 Cement
* Find clinker price
* Find factor for electricity cost
* Look into metal reclamation from the slag
* High Ni and V
* Add elemental composition to petcoke flow sheets

### A14.3.2: Presentation Comments

**Mentors’ Comments on Presentation 1:**

* Sulfur recovery
  + Lower temperature required (Need to be done at < 100 ⁰C)
* Location
  + Need to work out the final location for the plant with chem production team
  + Also need to take on account the production of acetic acid
* Possible CO2 Recovery
  + Needs to set up the CO2 footprint for our process
  + Figure out how large it is going to be
  + Can it be sold as a product?
* Amount of Syngas produced
  + As of right not chem production team is only using 1000 ton/day syn gas
  + Can we do something else with the rest of the syngas?

**Mentor’s Comments on Presentation 2:**

1. What is the heat rate?

* It is used as the measure of the electricity produced
* It was a substitute for the efficiency

1. Why do we use N2 as the transport medium for the feed transport? Why not CO2
2. How are we going to absorb the CO2 from the gas?
3. What is quench?

* Needs better explanation for quench.

1. Needs Better block diagram for water-gas shift reaction

Other questions asked to other groups?

* Use heat recovery generator to generate steam for the recovery of the heat
* What to do with the slag and the ash? (might be useful for cement?)

**Mentors’ Comments on Presentation 3:**

None

# Appendix 15: Information Sources and References

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