Clean Coal Feasibility Study

Integrated Gasification Combined Cycle

Team 15: Don’t Get Coal’d Feet
Sam Cooper, Nathan Frazier, Paul Urane

Calvin College
Department of Engineering
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Executive Summary

Senior Design Team 15 “Don’t Get Coal’d Feet” is comprised of three chemical engineering students: Sam Cooper, Nathan Frazier, and Paul Urane. For our project, we have teamed up with two other design teams to investigate the feasibility of clean coal technology as a potential source of energy for the production of electricity in the United States.

Although alternative energies such as solar and wind power have great potential, they still only represent a very small portion of the United States’ overall energy portfolio. On the other hand, clean coal power production has an established foundation and the United States has an abundance of coal reserves at its disposal. There are several different clean coal options currently available, but based on plant capacity, thermal efficiency, and ability to sequester plant emissions, our team investigated the feasibility of implementing an integrated gasification combined cycle as a source of power production. Our team was responsible for designing the gasification unit, the water-gas shift reactors, and the oxygen generation unit for the plant. Team 14 was responsible for designing the separations equipment for H₂S and CO₂ removal and Team 16 was responsible for designing the turbines used in the combined cycle area of the plant. An additional team of geologists investigated potential plant locations based on coal availability and potential for CO₂ sequestration.

For gasifier design, we chose an entrained flow gasifier as the best type of gasifier to meet our plant’s needs and used Shrinking Core model kinetics to model the gasification reactions using Polymath. Using our kinetic model, we were able to size the gasifier and use the dimensions, along with materials of construction, to estimate unit cost. The water gas shift reactors were modeled in UNISIM using kinetic data from literature. Using this information, we specified dimensions and materials of construction for the reactors to estimate cost. For oxygen generation, we decided cryogenic distillation was the best method to meet the plant’s design requirements. The process was modeled and optimized in UNISIM. All the associated equipment was then specified and cost estimations were calculated.

Team 14 designed a H₂S and CO₂ removal system using a liquid adsorbent Selexol. They used UNISIM to model the removal process and provided specifications and cost estimates for all the needed equipment. Team 16 provided specification and cost estimations for the turbine units in the combined cycle. The geology team proposed a plant location in West Olive, MI because of its close by coal reserves and formations capable of storing CO₂ underground indefinitely.

In conjunction with Team 14, we built an integrated plant model incorporating all the plant unit operations in UNISIM and scaled the inputs so the plant provided a target 1000 MW of power. This integrated model was then used to determine plant production costs.

Using the capital cost and operating cost estimations, we were then able to perform a cash flow analysis and determined that that breakeven price of electricity for the plant was $0.14/kWh. In order for the plant to provide a 12% internal rate of return, a typical value for a utility company, the electricity would need to be sold for $0.18/kWh. Currently, West Michigan consumers pay an average of $0.10/kWh. This indicates that the construction of an integrated gasification combined cycle plant to provide energy to Michigan consumers would result in a significant increase in energy costs.
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1 Introduction

1.1 “Clean” Coal

According to the Environmental Protection Agency (EPA) the process of generating electricity by the combustion of fossil fuels is the single largest source of carbon dioxide emissions in the United States, representing 41 percent of all carbon dioxide emissions. Approximately one third of these emissions are due to the combustion of coal specifically, while the other two thirds are a result of the combustion of natural gas and petroleum. In 2006, 1.9 billion metric tons of CO₂ were emitted to the atmosphere by coal combustion.¹

In order to reduce the amount of carbon dioxide emissions to the environment, the Kyoto Protocol to the United Nations Framework Convention on Climate Change was adopted in the late 1990s and entered into force in 2005. Currently there are 192 counties and the European Union as a regional economic integration organization that have ratified the Kyoto Protocol. As of November 2011, the only remaining signatory yet to ratify the protocol is the United States.² Much controversy exists regarding the effectiveness of the Kyoto Protocol to reduce world carbon emissions. Though evidence has shown that the protocol has had little effect on the world’s carbon dioxide emissions³, it represents a first step by the governments of the world in the regulation of greenhouse gases.

Though the United States has not yet entered the Kyoto agreement, there have been attempts in Congress to establish limits of emissions linked to global climate change. The Waxman-Markey Bill was passed by the 111th United States House of Representatives (H.R. 2454) in 2009, but died in the Senate. This vote was the “first time either house of Congress had approved a bill meant to curb the heat trapping gases scientists have linked to climate change.”⁴

On March 27, 2012 the Obama administration’s EPA proposed the first rules to cut carbon dioxide emissions from new U.S. power plants. According to the new standard, no new power plants with capacities exceeding 25 MW can produce more than one pound of carbon dioxide per kilowatt-hour produced⁵. This new proposal eliminates coal as an electric generating source without carbon capture and storage (CCS) technology. To date, CCS has not been accomplished on a commercial scale.

We recognize the need to determine the cost implications of carbon capture technology on coal fired power plants. The goal of this project is to determine how much electricity production costs would increase with the implementation of clean coal technology in the United States.

¹ Environmental Protection Agency, 2011
² United Nations Framework Convention on Climate Change, 2011
³ The International Bank for Reconstruction and Development/The World Bank, 2010
⁴ Broder, 2009
⁵ Rust, 2012
1.2 Project Scope

This study can be summarized as a preliminary comparison between a base case conventional coal power plant and two clean coal power plants – one a conventional power plant with a pure oxygen feed and carbon sequestration technology, and the other an integrated gasification combined cycle plant (IGCC) with carbon sequestration technology. Both plants will be studied in order to determine which one could produce energy the most efficiently, cleanly, and at the lowest cost.

Each of the project teams has taken various parts of each of the three plants to study in greater detail. Specifically, Team 15: “Don’t Get Coal’d Feet” has focused on the Integrated Gasification Combined Cycle (IGCC) specifically, including gasifier design, the water gas shift reactions, and the oxygen generation unit design.

Team 16’s project goals included boiler design and the overall analysis of the conventional coal fired power plant and the conventional design with an oxygen feed process. They also were responsible for coal pre-processing for each of the proposed power plants. Team 14’s project goals included design of the separations equipment in all three proposed power plants. They have specifically focused on scrubber design, as well as technology for removing carbon dioxide from the process streams for sequestration. A team of geology students looked into what to do with the carbon dioxide once it has been sequestered. They are exploring the storage of CO₂ in oil reservoirs and other geological formations. The team of geologists is also concerned about the source of the coal and has determined heat values, compositions, and costs for several different types of coal.

The goal of this project was to determine the price of electricity produced by each plant. This provides a comparison that will show the difference in cost consumers would see if a clean coal power plant was built near their homes as opposed to a conventional coal power plant. Overall, we had to goal of determining whether or not it is feasible to implement clean coal in the United States.

1.2.1 Plant Overviews

1.2.1.1 Conventional

Many large scale conventional power plants employ the combustion of pulverized coal in a boiler to create steam. Steam created in the boiler is used to drive rotating turbines coupled to rotating electric generators which produce electricity. A conventional pulverized coal plant utilizes is a the Rankine cycle, where water is boiled in a steam generator, is expanded in a turbine to make work, and then is condensed before being pumped back to the steam generator. Figure 1 provides an overview of the equipment involved in a conventional coal power plant. Because this process is relatively simple and reliable, power plants employing this method have dominated the power industry in the United States for the greater part of the 20th century. Water is usually the working fluid of choice since it is nontoxic, abundant, and low cost.⁶

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⁶ Collins, 2000
For the scope of this project, we will analyze a 1000 MW pulverized coal power plant without carbon capture and sequestration technology for comparison purposes against our clean coal options. We realize that the conventional pulverized coal power plant is most likely less expensive to build than the clean coal options but with the new EPA limit of 1 lb CO$_2$/kWh, they can no longer be built because pulverized coal plants usually emit 1.6-1.9 lb CO$_2$/kWh. Team 16 is responsible for the design and economic analysis of the conventional coal option.

1.2.1.2 Conventional with Oxygen

The first clean coal option we are considering is a conventional coal power plant with two major pollution control equipment being added – an air separation unit which will feed pure oxygen to the boiler rather than air, and post combustion carbon dioxide capture technology. This plant will also be sized to produce 1000 MW of electricity in order to get the most accurate comparison to the conventional pulverized coal power plant, and is a level of power output that is achievable for both this clean coal option as well as the IGCC.
The basic process underlying this clean coal option is the same as a conventional pulverized coal plant. Steam created in the boiler is used to drive rotating turbines coupled to rotating electric generators which produce electricity by a Rankine cycle. Again, the only differences associated with this clean coal option compared to the conventional plant are the pure oxygen feed and the post combustion carbon capture technology. These differences are shown in the process flow diagram in Figure 2. Pure oxygen is of value in this plant because it increases the efficiency of the coal combustion and increases the CO₂ emission concentration leaving the plant because they are not diluted by significant amounts of N₂ that would be put through the system if ambient air were used as the feedstock. The more concentrated the CO₂ leaving the plant, the easier it is to capture it for sequestration. Team 16 is primarily responsible for determining costs for this clean coal option using information on oxygen generation and separations technology provided by Teams 15 and 14, respectively.

1.2.1.3 Integrated Gasification Combined Cycle

The second clean coal option being considered in this study is the Integrated Gasification Combined Cycle power plant (IGCC). For the IGCC, coal is usually processed and made into coal slurry depending on the type of gasifier used. This plant requires more coal pre-processing than the conventional plant and the conventional plant with oxygen because of the gasifier specifications for finer coal particle size. This can be done using a jaw crusher, where lumps of coarse coal are reduced from a diameter of four inches and greater to approximately one inch in diameter. The coal is then reduced further using a rod
mill to particles between 10-35 mesh, and then to greater than 140 mesh using a ball mill. From there the coal is mixed with water to create a slurry and is delivered to the gasifier via a slurry pump. For a conventional power plant, 4 mesh and larger are acceptable sizes for cyclone furnace firing.

In the gasifier coal particles or coal slurry is reacted with oxygen at high temperature and pressure to form raw syngas, which is then cleaned of sulfur and fed into water gas shift reactors combined with interstage cooling to achieve a higher conversion to hydrogen and carbon dioxide. The stream which is not CO\textsubscript{2} and H\textsubscript{2} is sent through a stripper that removes the CO\textsubscript{2} for sequestration. Once cleaned, the H\textsubscript{2} is sent to the hydrogen fuel gas turbine. Before being released to the atmosphere, the hot exhaust of the fuel gas turbine is utilized to heat water in the Heat Recovery Steam Generator (HRSG), which makes steam that can be used to drive a steam turbine. The combination of the fuel gas turbine and steam turbine is called a combined cycle, and is the reason that the overall plant efficiency of the IGCC is higher than conventional coal power plants. A flow diagram of this process is shown in Figure 3. A conventional coal plant generally operates at 32-38% efficiency and an IGCC plant operates at 45-50% efficiency.

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7 Seader, Henley, & Roper, 2010
8 World Coal Association, 2011
1.2.2 Environmental Considerations

Coal fired power plants are a major source of sulfur dioxide and nitrogen oxide emissions, which contribute to the formation of ground level ozone, as well as mercury emissions from the combustion of coal. Federal air regulations on coal fired power plants are likely to become more stringent in the future, and thus utility companies will be required to make decisions regarding the implementation of pollution control equipment and the retirement of existing coal technology. The new EPA regulation for carbon dioxide may just be the first of several increasingly restrictive regulations imposed on the coal power industry.

Currently the Environmental Protection Agency (EPA) has set the following National Ambient Air Quality Standards (NAAQS) for pollutants considered harmful to the public, health, and environment. The Clean Air Act identifies two types of national ambient air quality standards. Primary standards provide public health protection, including protecting the health of sensitive populations such as asthmatics, children, and the elderly. Secondary standards provide public welfare protection, including protection against decreased visibility and damage to animals, crops, vegetation, and buildings. A summary of some of these pollutants is provided in Table 1.

<table>
<thead>
<tr>
<th>Pollutant [final rule cite]</th>
<th>Primary/Secondary</th>
<th>Averaging Time</th>
<th>Level</th>
<th>Form</th>
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<td>8-hour</td>
<td>9 ppm</td>
<td>Not to be exceeded more than once per year</td>
</tr>
<tr>
<td></td>
<td></td>
<td>1-hour</td>
<td>35 ppm</td>
<td></td>
</tr>
<tr>
<td>Lead [73 FR 66964, Nov 12, 2008]</td>
<td>primary and secondary</td>
<td>Rolling 3 month average</td>
<td>0.15 μg/m³ (a)</td>
<td>Not to be exceeded</td>
</tr>
<tr>
<td>Nitrogen Dioxide [75 FR 9474, Feb 9, 2010] [51 FR 52852, Oct 8, 1996]</td>
<td>primary</td>
<td>1-hour</td>
<td>100 ppb</td>
<td>98th percentile, averaged over 3 years</td>
</tr>
<tr>
<td></td>
<td>primary and secondary</td>
<td>Annual</td>
<td>53 ppb (2)</td>
<td>Annual Mean</td>
</tr>
<tr>
<td>Ozone [73 FR 16436, Mar 27, 2008]</td>
<td>primary and secondary</td>
<td>8-hour</td>
<td>0.075 ppm (2)</td>
<td>Annual fourth-highest daily maximum 8-hr concentration, averaged over 3 years</td>
</tr>
<tr>
<td>Particle Pollution [71 FR 61144, Oct 17, 2006]</td>
<td>PM_2.5</td>
<td>primary and secondary</td>
<td>Annual</td>
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<tr>
<td></td>
<td></td>
<td>24-hour</td>
<td>35 μg/m³</td>
<td>98th percentile, averaged over 3 years</td>
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<tr>
<td></td>
<td>PM_10</td>
<td>primary and secondary</td>
<td>24-hour</td>
<td>150 μg/m³</td>
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<td>Sulfur Dioxide [75 FR 35520, Jun 22, 2010] [38 FR 25678, Sept 14, 1973]</td>
<td>primary</td>
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<td>75 ppb (4)</td>
<td>99th percentile of 1-hour daily maximum concentrations, averaged over 3 years</td>
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<tr>
<td></td>
<td>secondary</td>
<td>3-hour</td>
<td>0.5 ppm</td>
<td>Not to be exceeded more than once per year</td>
</tr>
</tbody>
</table>

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9 Wisconsin Department of Natural Resources, 2007
10 Environmental Protection Agency, 2011
In March 2011, the EPA has proposed unprecedented national standards for mercury, arsenic, and other toxic air pollution specifically from power plants, and new standards are being written and reviewed. More extensive information regarding proposed emission standards (including those for other transition metals) can be found in the Federal Register, Vol. 76, No. 85 (Tuesday May 3, 2011).

The project team regards the health of the public and the wellbeing of the environment significantly important to the design of any future coal power plants. Design of the clean coal plant options will be conducted in such a way as to minimize any species emissions considered harmful to the environment, including the production of carbon dioxide, by using current EPA emissions standards as criteria. Both clean coal options (the IGCC and the conventional plant with oxygen) will include carbon dioxide capture and sequestration technology. Though it has been shown that carbon dioxide capture is generally expensive and detrimental to the overall efficiency of a power plant, increased possibility of future carbon tax and carbon curbing legislation may offset some of the costs associated with carbon capture and sequestration. Overall, we believe that the health of the public and wellbeing of the environment is of greater importance to the additional capital and operating costs associated with removal of carbon dioxide. The project team plans to meet or exceed the current emissions standards set by the EPA. The current carbon dioxide limit is 1 lb/kWh which would require sequestration of about 50% of the CO$_2$ emitted from a coal plant to meet this standard. However, we chose 90% as a minimum level of carbon dioxide sequestration so that the standard is exceeded by a large factor. This will provide the plant we design with a robustness that will enable it to withstand potential lowering of these emissions standards in the future. Although the standard today is 1 lb/kWh, we have no way to predict what it will be in 30 years.

According to the Global CCS Institute, three main processes are being developed to capture carbon dioxide from power plants that use coal or gas. These processes are post-combustion capture, pre-combustion capture, and oxy-fuel combustion capture. All involve separating the CO$_2$ from process or flue gases.

For the conventional coal plant with an oxygen feed, a post-combustion carbon dioxide capture process will be explored for the removal of 90% of carbon dioxide emissions. Upon the recommendation of Team 14, a CO$_2$ removal process using an adsorbent called Selexol was decided upon as the best method to economically meet out CO$_2$ removal goals. Selexol is also capable of removing sulfur from the stream leaving the gasifier so this process will be used for both separations. The decision process outlining different CO$_2$ and sulfur removal techniques and a decision matrix outlining Selexol as the best method can be found in Team 14’s report.

1.2.3 Facility Location

We have determined that West Olive, MI is a good location for the clean coal power plant based on several criteria researched by the team of geology students.

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11 Environmental Protection Agency, 2011
12 Global CCS Institute, 2011
One condition vital to any power plant is the availability of a heat sink. Often, bodies of water are used as heat sinks to dispense heat associated with the Rankine cycle, which is the underlying process in the conventional plant, the conventional plant with oxygen, and the steam turbine in the IGCC. If any of the clean coal designs require water to condense the steam used to drive the steam turbines, the facility should be sited near a body of water. West Olive, MI has Lake Michigan as an onsite water source, which eliminated the need for cooling towers. Overall, the plant efficiency depends heavily on the amount of cooling capacity that is available. Since Lake Michigan has plenty of water available, plant efficiency will be higher than if air cooled exchangers were used.

CO₂ sequestration options were also a main deciding factor in where the facility is located. If the plant is producing several thousand tons of carbon dioxide a day and sequestering it rather than pumping it into the atmosphere, a geological formation must be found where it can be safely stored underground. In West Olive, MI the geological formation called Mt. Simon Sandstone (Cambrian) has a storage capacity of greater than 29 billion metric tons with a cap rock type called Eau Claire Fm-Dolomite. This formation is in Ottawa County 5,050 ft deep and must be pumped to a depth beyond 5,900 ft. With minimum 29 billion metric ton capacity, this formation has a minimum storage lifetime of 2800 years.

Coal availability is another important criterion associated with the siting of a clean coal plant. Costs are minimized when the plant is fairly close to the coal it uses to generate power because of transportation costs. A map showing the coal regions of the United States is found in Figure 4. In West Olive, MI the clean coal facility will be sited near one of the coal regions associated with the type of coal required by the boiler or gasifier. Transportation to the site will be by train, where approximately 6 million short tons of coal are used per year. Most coal traveling to West Olive, MI comes from nearby Wyoming, MI which has better quality Illinois no. 6 coal.

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13 U.S. Department of the Interior, 2006
Figure 4: Coal regions of the continental United States
1.2.4 Power Capacity

To make sure all the different power generation options are compared on an even platform, we decided that we must design all the plants to provide the same amount of sellable energy. In order to determine how large to design our power plants, we first had gauge the size of coal plants already in existence in the United States. Table 2 shows a breakdown of existing coal fired power generation units by age and capacity.

Since the demand for electricity is high, we determined that a larger plant would give cost advantages due to the economies of scale principle, and may help lower the average cost of producing electricity. There are several larger plants operating in the United States, but currently the largest IGCC plants in operation are 500 MW plants. Therefore, we chose a plant size of 1000 MW because we decided that this was a realistic scale up factor from existing IGCC plants and, while it is not as large as the biggest pulverized coal plants, it is still large enough to be competitive with many coal plants online today.
Table 2: Size and capacity of coal fired power generation units in the United States\textsuperscript{14}

<table>
<thead>
<tr>
<th>Unit Size</th>
<th># of Units</th>
<th>Total Capacity</th>
</tr>
</thead>
<tbody>
<tr>
<td>0-10 MW</td>
<td>37</td>
<td>192 MW</td>
</tr>
<tr>
<td>10-20 MW</td>
<td>25</td>
<td>345 MW</td>
</tr>
<tr>
<td>20-50 MW</td>
<td>75</td>
<td>2,427 MW</td>
</tr>
<tr>
<td>50-100 MW</td>
<td>73</td>
<td>5,269 MW</td>
</tr>
<tr>
<td>100-250 MW</td>
<td>85</td>
<td>14,000 MW</td>
</tr>
<tr>
<td>250-500 MW</td>
<td>97</td>
<td>34,396 MW</td>
</tr>
<tr>
<td>500-750 MW</td>
<td>69</td>
<td>42,655 MW</td>
</tr>
<tr>
<td>750-1,000 MW</td>
<td>28</td>
<td>23,612 MW</td>
</tr>
<tr>
<td>1,000-1,500 MW</td>
<td>59</td>
<td>72,366 MW</td>
</tr>
<tr>
<td>1,500-2,000 MW</td>
<td>38</td>
<td>66,657 MW</td>
</tr>
<tr>
<td>Over 2,000 MW</td>
<td>29</td>
<td>73,920 MW</td>
</tr>
</tbody>
</table>

1.3 Project Team Management

1.3.1 Team Organization

The eleven person clean coal team was divided into three smaller sub-teams to make the work load more manageable. Rather than having each team do a full analysis of one plant, each team became “experts” in certain technologies that were utilized in all three plants. For example, our team (Team 15) conducted the research regarding the coal gasifier and water gas shift reactors for the IGCC and the air separation units (ASU) that would be required for the IGCC and the conventional plant with oxygen. We also were responsible of the overall analysis of the complete IGCC process. Team 16, a team with three mechanical engineers and one chemical engineer, focused on the boilers for the conventional plant and conventional with oxygen and the turbines used in all three plants and were responsible for the overall analysis of the two conventional processes. Team 14, a team with two each of chemical and mechanical engineers, focused on the gas scrubbers and carbon sequestration technologies used in all three plants. The team of geologists focused on the problem of emission sequestration as well as the properties of the coal we would be using and determining a location for the power plant to be built.

Coordinating three sub-teams to work on the same project added a level of complexity in comparison with normal senior design projects. To deal with overall team coordination, each sub-team designated a project manager to an executive team. The executive team, comprised of Sam Cooper, Ben Markosky, and Ike Kuiper was responsible for making the agendas for weekly full team meetings as well as inter-team communication. A visualization of this organizational model is seen in Figure 6.

\textsuperscript{14} U.S. Energy Information Administration, 2011
1.3.2 Schedule

A Gantt chart was developed at the beginning of each semester showing a breakdown of all the tasks completed to date and a plan for all the anticipated tasks to be completed by the end of the project. This chart shows all the major tasks that must be completed to include in the final report and has an estimate as to how long each task will take to complete. These tasks have been divided into four major lists: the gasifier design, the air separation unit design, the water gas shift design, and full design integration. These major tasks were then divided into smaller subtasks to create a timeline of development for all parts of the project.

For the gasifier design work, February and March consisted of work to create and optimize a Polymath model to determine all inputs and outputs and determine optimal vessel design. During this period we also moved into the process of modeling different catalysts in UNISIM to find the most effective one to use for the water gas shift and develop a rate law for the reaction. By mid-March a catalyst model was chosen and optimized. Work then began on the actual water shift reactor design and continued into mid-March. After the reactors were designed the entire modeled process of the gasifier and the water gas shift will be optimized. By April we were able to do a cost analysis and design of minor equipment to determine overall capital and operating costs for the gasifier and water gas shift reactors.

In February, the work on the air separation unit design consisted of the development of a UNISIM model for a cryogenic distillation process. This model is a rigorous distillation simulation and provided an estimate of the total energy cost needed to run a distillation plant for oxygen generation. In March, the UNISIM model was optimized and research was conducted to determine necessary oxygen requirements for the gasifier. By the end of March, the optimal oxygen output was determined, and research into the design of plate fin heat exchangers was conducted. In April, a full cost analysis of the unit design was performed to determine all capital and operating costs for the ASU.

In late April, all the cost analysis for all parts of the plant was completed and all the information from the different teams was combined to determining the total cost of the energy produced by the IGCC power plant. This is the comprehensive report delivered in paper form on May 9, 2012 which was the basis for our Senior Design Night Presentation on May 5, 2012. The Gantt chart for the first semester of work is shown in Appendix D. The Gantt chart for the second semester of work is shown in Appendix E.
2 Gasification Block

Gasification is a process where carbonaceous materials are reacted with oxygen at high temperatures to produce synthesis gas or “syngas,” a high energy mixture of carbon monoxide and hydrogen. Gasification technology is an effective way of extracting high energy fuel from a variety of organic materials including: wood, biomass, and coal. For the purposes of this study coal is the target reactant due to its availability in the United States, high energy content, and low cost.\(^{15}\)

The physical processes in any type of gasifier can be organized into four main categories: drying, pyrolysis, oxidation, and reduction, as illustrated in Figure 7. Although designs may vary, these four main processes constitute basic zones that can be found in any gasification reactor.

A schematic of the process is as follows:

1. The coal enters the gasifier and is heated. This heating removes the moisture content in the coal and prepares it for pyrolysis.\(^{16}\)
2. The coal undergoes pyrolysis, or the thermal decomposition of organic matter to yield char ash and volatile gasses.
3. The volatile compounds contact the oxygen feed in the gasifier and undergo multiple, competing, oxidative reactions.\(^{17}\)
4. The oxidized gases, and volatile gases from pyrolysis, react further in an oxygen deprived environment to undergo reduction. It is in this phase of gasification that the majority of carbon monoxide and hydrogen are produced and leave as syngas.\(^{18}\)

![Figure 7: Schematic of overall gasification process](image)

After the volatile components in the coal have been released through pyrolysis, they are contacted with oxygen and undergo a series of competing reactions. Research on gasification suggested that thirteen unique reactions would be significant during the oxidation and reduction phases, shown in Table 3.\(^{19}\)

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\(^{15}\) Raymond Kirk, 1985  
\(^{16}\) Visagie, 2008  
\(^{17}\) Visagie, 2008  
\(^{18}\) Visagie, 2008  
\(^{19}\) Visagie, 2008
Table 3: Proposed reactions occurring in the gasification stage

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>( C + H_2O ) \rightarrow CO + H_2 \</td>
<td></td>
</tr>
<tr>
<td>( C + CO_2 ) \rightarrow 2CO</td>
<td></td>
</tr>
<tr>
<td>( C + 2H_2 ) \rightarrow CH_4</td>
<td></td>
</tr>
<tr>
<td>( CO + H_2O ) \rightarrow CO_2 + H_2</td>
<td></td>
</tr>
<tr>
<td>( C + \frac{1}{2}O_2 ) \rightarrow CO</td>
<td></td>
</tr>
<tr>
<td>( C + O_2 ) \rightarrow CO_2</td>
<td></td>
</tr>
<tr>
<td>( 2H_2 + O_2 ) \rightarrow 2H_2O</td>
<td></td>
</tr>
<tr>
<td>( CO_2 + 4H_2 ) \rightarrow 2H_2O + CH_4</td>
<td></td>
</tr>
<tr>
<td>( CO + \frac{1}{2}O_2 ) \rightarrow CO_2</td>
<td></td>
</tr>
<tr>
<td>( CH_4 + \frac{1}{2}H_2O ) \rightarrow CO + 3H_2</td>
<td></td>
</tr>
<tr>
<td>( CH_4 + 2O_2 ) \rightarrow CO_2 + 2H_2O</td>
<td></td>
</tr>
<tr>
<td>( CO + 3H_2 ) \rightarrow H_2O + CH_4</td>
<td></td>
</tr>
<tr>
<td>( CO + H_2 ) \rightarrow H_2O + C</td>
<td></td>
</tr>
</tbody>
</table>

For the initial simulation performed in this study, only four out of the thirteen proposed reactions were used. These four reactions were determined to have the largest effect on the overall gasification process and for initial estimates we decided that it would be acceptable use them exclusively in the model. The simplification to only four reactions, shown in Table 4, made it easier to model the gasifier for quick estimations but it will be necessary to explore the effect of all thirteen reactions in the coming semester.

Table 4: The four reactions used to model the gasifier.

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Product</th>
</tr>
</thead>
<tbody>
<tr>
<td>( C + \frac{1}{2}O_2 ) \rightarrow CO</td>
<td></td>
</tr>
<tr>
<td>( C + O_2 ) \rightarrow CO_2</td>
<td></td>
</tr>
<tr>
<td>( C + H_2O ) \rightarrow CO + H_2</td>
<td></td>
</tr>
<tr>
<td>( C + CO_2 ) \rightarrow 2CO</td>
<td></td>
</tr>
</tbody>
</table>

Depending on the inlet conditions and compositions, it is possible to favor certain reactions to achieve a desired syngas composition. Higher amounts of oxygen fed to the reactor display greater combustion characteristics while limiting the amount of oxygen results in greater concentrations of the desired product. This effect is shown in Figure 8 which plots product composition from the oxidation phase of gasification as a function of feed composition.

Varying the ratio of oxygen to coal in the feed ratio greatly affects the composition of gas leaving the gasifier. When this ratio is below a value of roughly 0.7, there will be hydrogen, carbon monoxide, trace amounts of methane, and non-combusted coal leaving the gasifier. On the other hand, when the feed ratio is above roughly 2.5 the coal is combusted to carbon monoxide, and water vapor. If the ratio of

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20 Visagie, 2008  
21 Visagie, 2008  
22 Visagie, 2008
oxygen to coal is too high, the oxygen will pass through the gasifier without reacting and will be found in the exiting gas composition.

When the ratio of oxygen to coal in the feed is between the values of 0.7 and 2.5, gasification will occur to yield varying amounts of: CH₄, H₂S, SO₂, H₂O, H₂, CO₂, and CO. Of these molecules, carbon monoxide and hydrogen are the most desirable in gasification. Consequently, it is advantageous to operate at a feed ratio of around 0.7 to 1 in order to maximize the quality of the syngas leaving the gasifier.

Since syngas contains a large amount of carbon monoxide, chemical processes like the water gas shift reaction can be utilized to produce additional hydrogen gas and carbon dioxide by reacting water and carbon monoxide.

Furthermore, carbon dioxide and hydrogen are two molecules with a large difference in size, making efficient separation possible. Consequently, carbon can be effectively sequestered while hydrogen is fed to a gas turbine for combustion. Using this method, the energy content stored in coal is efficiently utilized while greenhouse gas emissions are sequestered.

While effective conventional combustion always requires a specific variety of coal, gasification processes do not require this specification. This characteristic is especially relevant in technology that operates at

---

Figure 8: Syngas composition as a function of oxygen and coal feed ratios.

Since syngas contains a large amount of carbon monoxide, chemical processes like the water gas shift reaction can be utilized to produce additional hydrogen gas and carbon dioxide by reacting water and carbon monoxide.

Furthermore, carbon dioxide and hydrogen are two molecules with a large difference in size, making efficient separation possible. Consequently, carbon can be effectively sequestered while hydrogen is fed to a gas turbine for combustion. Using this method, the energy content stored in coal is efficiently utilized while greenhouse gas emissions are sequestered.

While effective conventional combustion always requires a specific variety of coal, gasification processes do not require this specification. This characteristic is especially relevant in technology that operates at

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23 Visagie, 2008
higher temperatures and pressures (approximately 1000 °C and 40 MPa)\textsuperscript{24} where low grade lignite undergoes pyrolysis to release volatile components just as easily as higher grade anthracite. For this study medium grade bituminous coal was considered as the primary fuel source for the plant due to its low cost and high availability in the United States. While not every gasifier design has the capability of using any type of coal feedstock, the gasifier design we decided to use includes this flexibility.

\subsection*{2.1 Types of Gasifiers}

There are many different types of gasifiers that operate at a variety of temperatures, pressures, and capacities. Despite the wide variety of technologies and operating conditions, gasifiers can be divided into four categories based on the type of bed utilized: fixed, molten bath, fluidized, and entrained.

\subsubsection*{2.1.1 Fixed Bed Gasifiers}

In a fixed bed gasifier, oxygen is passed through a fixed grate holding coal. Fixed bed gasifiers can be separated into two different categories: counter-current, or “up draft”, and co-current, or “down-draft” gasifiers. Although very similar in design, counter-current and co-current configurations differ mainly on location of oxidant input and syngas output.

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{fixed-bed-gasifier.png}
\caption{Design and reactive zones for a fixed bed co-current gasifier.\textsuperscript{25}}
\end{figure}

\textsuperscript{24} Visagie, 2008
\textsuperscript{25} Graz
In co-current gasification (Figure 9), coal is fed at the top of the reactor while the oxidant enters near the middle section. The heavy fly ash accumulates at the bottom of the gasifier while syngas flows out of the bottom directly before the fly ash catch.

In counter-current gasification (Figure 10), the oxygen, and possibly steam, is fed at the bottom of the reactor while the syngas product exits out of the top. The main difference between co-current and counter-current gasification is how the gasification zones are organized within the reactor. In co-current designs, oxidation occurs in the middle of the reactor and reduction reactions transpire at the bottom. In counter-current, air enters at the bottom prompting oxidation. Reduction for the countercurrent system occurs at the middle of the gasifier.

As far as gasification technology is concerned, fixed bed gasification is the most developed. Consequently, these systems usually have the lowest capital cost and a solid industry knowledge base. This is an important factor to consider because it means that fixed bed systems will not only be easier to maintain and troubleshoot, but the design and integration into production will be much more reliable.

Another advantage of fixed bed systems is that the coal does not have to be processed with pulverization as most other systems require. Instead, large pieces of coal about the size of a typical brick are acceptable. A disadvantage, however, is that fixed bed gasifiers cannot easily use coal that has a tendency to cake. If caking coal is used, the gasifier design becomes more complex because a stirrer or a

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26 Graz
rotating grate must be added to the design. Although fixed bed systems are generally simple, this complexity could increase the opportunity of malfunction and increased regular maintenance.

The Roga test, ISO 335:1974, is a standardized test to measure the caking power of coal. The method is defined by carbonization of a mixture of 1 g of coal and 5 g of standard anthracite. The crucible coke from this carbonization is submitted to a drum test and given a rating on the Roga index based upon the mechanical strength of the crucible coke.

Fixed bed gasifiers usually run at lower temperatures and pressures, an attribute which is beneficial to the thermal efficiency of the system. As a semi-batch system, however, they have a lower capacity and cannot produce as much syngas as continuous gasifiers of the same size.

2.1.2 Fluidized Bed Gasifiers

In fluidized bed gasifiers, the oxidation gases, generally air or oxygen, are fed through the bottom of the reactor and up through fine particles of pulverized coal creating a boiling effect. This effect greatly increases the surface area between the coal and oxygen creating an ideal environment for gasification reactions. Typically coal is fed into the gasifier near the middle of the vessel and the rising gas elevates it to the top of the fluidized bed where it undergoes drying and pyrolysis. As it rises in the vessel the mixture passes through several constriction points. The flow passes through these sections where the cross section of the vessel is restricted and the solids flow back adjacent to the wall of the vessel due to an eddy effect. The particles then flow along the wall which directs them back into the fast gas phase passing through the center of the gasifier. There is high gas-solid interaction in the area due to the turbulent nature of the flow. A diagram of the fluidized bed flow through a constriction point is illustrated in Figure 12. Once the coal is allowed to cycle in the fluidized bed it undergoes the oxidative and reductive reactions to product syngas. This syngas then rises though the fluidized bed and exits out of the top of the gasifier. Any coal carried with this exiting syngas is caught by the cyclones which cycle the solids back to the fluidized bed. Ash that was left after pyrolysis exits out of the bottom of the reactor.

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27 International Organization for Standardization, 2011
28 International Organization for Standardization, 2011
29 Visagie, 2008
30 Schmid, Proll, Pfeifer, & Hofbauer
Figure 11: General design of a fluidized bed gasifier.\textsuperscript{31}

Figure 12: Particle movement above constriction in a fluidized bed gasifier.\textsuperscript{32}

\textsuperscript{31} U-GAS ® Gasification Process, 1995
\textsuperscript{32} Schmid, Proll, Pfeifer, & Hofbauer
Fluidized bed systems usually run at higher temperatures (800-900°C), but lower pressures than fixed bed gasifiers (800-1000 kPa). A disadvantage of using a fluidized bed gasifier is it is unable to process caking coals which are bituminous coals that swell and agglomerate when heated, producing a consolidated cake. This cake disrupts the flow patterns in the fluidized bed and lowers the thermal efficiency of the gasifier. Therefore, a pretreatment process is used where the coal is heated by passing steam and air through it rather than heating it through the wall of gasifier. This process prevents agglomeration of the coal to the gasifier walls. An advantage of fluidized gasifiers is that, in the category of continuous flow systems, they have been the most developed. We know of at least five plants currently operating using this type of gasifier. These plants are listed in Table 5. Similarly to fixed bed gasification, these systems will be easier to model accurately for integration with production. A considerable improvement that fluidized gasification bed has over fixed bed technology is the great increase in capacity for reactor of the same volume.

2.1.3 Molten Bath Gasification

Molten bath configurations have been among the most recent developments in gasification technology. In molten bath systems coal is crushed and fed into a hot molten salt bath, which is most frequently molten sodium carbonate salt. The coal dissolves into the salt bath and reacts with the oxidant (generally air or oxygen) to produce carbon monoxide while the other constituents in the coal, such as sulfur, float on top of the molten bath until they are removed when the salt is recycled.

![Figure 13: A molten bath gasifier design from Rockwell International](image)

33 Visagie, 2008
34 Shadle, Berry, & Syamlal, 2002
35 Forney, Kenny, Gasior, & Field, 1964
36 Visagie, 2008
<table>
<thead>
<tr>
<th>Industrial Gasifiers</th>
<th>Gasifier Bed Type</th>
<th>Water Jacket</th>
<th>Pressure (MPa)</th>
<th>Temperature Range (°C)</th>
<th>Reagents</th>
<th>Flow</th>
<th>Coal</th>
<th>Situated</th>
<th>Coal Residence Time</th>
<th>Gas Velocity (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>British Gas/Lurgi Slagging</td>
<td>Fixed</td>
<td>yes</td>
<td>0.5-2.6</td>
<td>400-1300</td>
<td>c, s,o</td>
<td>semi-batch</td>
<td>bituminous, caking</td>
<td>Westfield, Scotland</td>
<td>10-15 min</td>
<td>0.75</td>
</tr>
<tr>
<td>Lurgi, Dry Ash</td>
<td>Fixed</td>
<td>yes</td>
<td>2.7</td>
<td>720-1150</td>
<td>c</td>
<td>s,o/a</td>
<td>semi-batch</td>
<td>almost any</td>
<td>Secunda, South Africa</td>
<td>1 hour</td>
</tr>
<tr>
<td>GFERC Slagging</td>
<td>Fixed</td>
<td>no</td>
<td>n/s</td>
<td>n/s</td>
<td>c</td>
<td>s,o</td>
<td>semi-batch</td>
<td>noncaking, high moisture (35%)</td>
<td>Grand Forks, North Dakota</td>
<td>-</td>
</tr>
<tr>
<td>Wellman-Galusha</td>
<td>Fixed</td>
<td>yes</td>
<td>0.1</td>
<td>620-1315</td>
<td>c</td>
<td>s,a</td>
<td>continuous</td>
<td>anthracite, bituminous, coke, sub-bituminous</td>
<td>Reading, Pennsylvania</td>
<td>4 hours</td>
</tr>
<tr>
<td>Woodall-Duckham</td>
<td>Fixed</td>
<td>yes</td>
<td>0.1</td>
<td>120-1200</td>
<td>c</td>
<td>s,a</td>
<td>continuous</td>
<td>non or weakly caking</td>
<td>Milan, Italy</td>
<td>several hours</td>
</tr>
<tr>
<td>Battelle Ash Agglomerating</td>
<td>Fluidized</td>
<td>no</td>
<td>0.7-0.9</td>
<td>870-890</td>
<td>c,a</td>
<td>s</td>
<td>s</td>
<td>semi-batch</td>
<td>non-caking</td>
<td>West Jefferson, Ohio</td>
</tr>
<tr>
<td>Carbon Dioxide Acceptor</td>
<td>Fluidized</td>
<td>yes</td>
<td>1</td>
<td>n/s</td>
<td>a</td>
<td>c,s</td>
<td>semi-batch</td>
<td>sub-bituminous, lignite</td>
<td>Rapid City, South Dakota</td>
<td>n/a</td>
</tr>
<tr>
<td>Hygas</td>
<td>Fluidized</td>
<td>no</td>
<td>7.9-8.1</td>
<td>315-1000</td>
<td>c</td>
<td>s,o</td>
<td>continuous</td>
<td>any</td>
<td>Chicago, Illinois</td>
<td>n/a</td>
</tr>
<tr>
<td>Synthane</td>
<td>Fluidized</td>
<td>n/s</td>
<td>4.1-6.9</td>
<td>815-980</td>
<td>c</td>
<td>s,o</td>
<td>continuous</td>
<td>any</td>
<td>Bruceton, Pennsylvania</td>
<td>n/a</td>
</tr>
<tr>
<td>Winkler</td>
<td>Fluidized</td>
<td>no</td>
<td>0.1-0.4</td>
<td>980-1150</td>
<td>s,o/a</td>
<td>c,s,o</td>
<td>continuous</td>
<td>lignite, sub-bituminous</td>
<td>Leuna, Germany</td>
<td>n/a</td>
</tr>
<tr>
<td>Babcock and Wilcox</td>
<td>Entrained</td>
<td>no</td>
<td>0.1-2</td>
<td>980-1870</td>
<td>c,s,o</td>
<td>continuous</td>
<td>most types, preferably dried</td>
<td>Morgantown, West Virginia</td>
<td>1 second</td>
<td>4.5-9</td>
</tr>
<tr>
<td>Bi-Gas</td>
<td>Entrained</td>
<td>yes</td>
<td>3.4-10.4</td>
<td>925-1650</td>
<td>c,s,o</td>
<td>continuous</td>
<td>any</td>
<td>S.O.Homer City, Pennsylvania</td>
<td>2-10 seconds</td>
<td>-</td>
</tr>
<tr>
<td>Combustion Engineering</td>
<td>Entrained</td>
<td>yes</td>
<td>0.1</td>
<td>1650-1750</td>
<td>c,s,a</td>
<td>continuous</td>
<td>any</td>
<td>Windsor, Connecticut</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Koppers-Totzek</td>
<td>Entrained</td>
<td>yes</td>
<td>0.04-0.05</td>
<td>1500-1925</td>
<td>c,s,o</td>
<td>continuous</td>
<td>any</td>
<td>New Delhi, India</td>
<td>1 second</td>
<td>-</td>
</tr>
<tr>
<td>Rockwell International</td>
<td>Molten Bath</td>
<td>no</td>
<td>0.1-1.9</td>
<td>980</td>
<td>c,s,o/a</td>
<td>semi-batch</td>
<td>any</td>
<td>8Santa Susana, California</td>
<td>60.5-0.7 seconds</td>
<td>n/a</td>
</tr>
<tr>
<td>Saarberg/Otto</td>
<td>Molten Bath</td>
<td>n/s</td>
<td>1.5</td>
<td>1500-1700</td>
<td>c,s,o</td>
<td>semi-batch</td>
<td>any</td>
<td>Wesseling, Germany</td>
<td>1.8 seconds</td>
<td>n/a</td>
</tr>
</tbody>
</table>
Molten gasifiers are advantageous because they allow for a less complex system with no moving parts to be used, though the bath circulation does increase the complexity. Also, molten gasifiers give the operator much more flexibility in choosing a coal to be gasified, since a molten bath system is able to process any type of coal.

As a relatively new technology, however, molten bath configurations are not well developed and consequently accurate design and integration is more difficult. The low level of development means that many attributes of molten bath gasifiers that have not been fine tuned. For example, the high temperatures of operation in conjunction with a molten salt may be corrosive to the refractory lining in the vessel. Also, particles from the coal may sometimes become entrained in the flow of syngas and kill the catalyst in the water gas shift reactors.

2.1.4 Entrained Flow Gasifiers

In entrained flow systems, the coal can enter the reactor as either a finely pulverized solid or it can be mixed with steam and enter as a slurry. Once coal enters, it is carried through the gasifier by the oxidant gases where it reacts as it travels through the system. The syngas produced from entrained gasification leaves from the top of the reactor along with a portion of the ash produced during coal pyrolysis. The rest of the inorganics in the coal melt to become slag and coat the refractory lining of the gasifier. The slag runs down the walls of the gasifier and collects in the water quench at the bottom where they are then removed, ground up, and sold to concrete and roofing companies.

The greatest advantage of entrained gasification is its extremely high capacity per volume. Operation at very high temperatures and pressures (1500°C and 3 MPa) contributes to an exceptionally low residence time of coal. These extreme conditions also enable any type of coal to be gasified. The high temperatures and pressures, however, make it necessary to carefully consider the lining on the inside of the gasifier and corrosion that could ensue. Also, the syngas leaves the gasifier at temperatures too high for current separation technology to handle. It is necessary to cool the syngas before treatment, reducing the overall thermal efficiency of the gasification system.
2.1.5 Choice of Gasifier

The different types of gasifiers were analyzed and evaluated on multiple levels in order to ensure that the best design would be used for the IGCC plant. Many factors were considered in this decision: extent of development, conditions of operation, complexity of the technology, capacity, efficiency, and capital cost.

In order to evaluate the different gasifiers designs fairly and effectively we developed a decision matrix that gave varying weights to each category. For each type of gasifier a score from 1 to 10 was given depending on the degree to which that system excelled or was found wanting. A score of 10 meant that the particular system was very good in the particular category. For example, a score of 10 for temperature and pressures meant that the gasifier functioned at low temperatures and pressures.

\[37\text{ D.O.E}\]
making it operationally easier to handle. Furthermore, a score of 10 in the capital cost section means that the system is a more cost effective alternative compared to other gasifiers.

The variables considered for the decision of the gasifier are as follows:

**Cost (15%)**:  
The capital cost of equipment is naturally extremely important in any profit seeking adventure. Consequently, cost was a variable that was given one of the most significant weights in the decision matrix. Fixed bed gasifiers are the most technologically developed and operate at the lowest temperatures and pressures. The materials required for fixed bed gasifiers construction are usually less expensive than other systems like the entrained and fluidized bed gasifiers. Accordingly, fixed bed gasifiers received the best rating: 10. Molten bath gasifiers, on the other hand, are associated with high operating costs and low technological development. This makes them more expensive and they received a rating of 4 in the decision matrix.

**Extent Developed (5%)**:  
Although this variable was partially accounted for in the cost section, it was important to evaluate the system based upon the amount known about it. More developed systems are easier to troubleshoot and maintain while less developed systems will be less proven in industry. Less weight was given to this category because, as already mention, it was accounted for in the cost section. The fixed bed gasifier received the best score in this category due to its well established knowledge base. Molten bath gasifiers, once again, as the newest gasification technology received the lowest score of 1. Entrained and fluidized bed gasifiers are about equally developed with a firm knowledge base. Those systems consequently earned a score of 7.

**Capacity (20%)**:  
The capacity that the gasifier system could handle was very important to consider in the design of the IGCC. Our simulations indicated that we would have to process around 4000 tons of coal per day. In this category the semi-batch gasifiers were unable to effectively compete with continuous flow systems. Entrained bed gasifiers, known to have the highest capacity and fastest throughput, earned a score of 10 while fluidized bed gasifiers were close behind with a score of 9. The fixed bed and molten bath systems, when compared to continuous configurations, were found lacking and consequently scored a 3.

**Temperature (5%)**:  
Temperatures of operation would affect the energy requirements, safety, and operability of each system. High temperature systems would be more dangerous and require more exotic metals for the construction of the vessel. Entrained and molten bath gasifiers operated at the highest temperatures, 1500°C for the entrained systems and 1700°C for molten systems, and consequently received scores of 4 and 1 respectively. Fluidized Bed systems, on the other hand, operated at temperatures near 1000°C and earned a score of 8. Fixed bed gasifiers are typically run at approximately 800°C and received the best score of 10.
Pressure (5%):
Similarly to temperatures, the pressures of operation would play a role in the energy requirements, safety, and operability of each system. Molten bath gasifiers have the ability to operate at atmospheric pressures and consequently scored 10. Fixed bed systems, however, tend to require high pressures ranging from 0.1 to 4 MPa and scored lowest out of all the configurations. Lastly, entrained and fluidized gasifiers tend not to exceed pressures of 3 MPa and both scored values of 6.

Waste (15%):
The amount of byproduct waste, or tar and fly ash, associated with each configuration was assessed and given considerable weight in the decision of the gasifier. As a clean coal technology it was important to consider alternatives that would reduce the environmentally harmful byproducts associated with coal. Systems that reduced the amount of tar that would be produced through gasification, such as the molten bath, entrained, and fluidized configurations received scores of 10, 8, and 7 respectively. The fixed bed gasifiers are notorious for significant amounts of tar production and consequently received a 1 in the matrix.

Coal Flexibility (10%):
The ability to use different grades of coal is an advantage that gasification holds over boiler combustion. Not every gasifier design, however, holds this flexibility. Fixed Bed gasifiers, which scored a 5, are more constrained by the caking tendency and quality of coal. Molten bath systems operate at such high temperatures that the grade and caking tendency of coal can be disregarded. Entrained and fluidized systems also share the advantage of significant coal flexibility due to their higher operation temperatures but are not as effective as molten bath gasifiers.

Complexity (5%):
The degree of complexity for each gasifier was assessed to give a determination of how difficult each system would be to operate and maintain. Fixed bed gasifiers are typically the least complex of all systems and consequently scored highest. Molten Bath systems are mechanically very simple but have the added complexity of molten salt technology giving them the lowest score. For the continuous gasifiers, entrained technology was deemed to be less complex than the fluidized bed systems. Fluidized bed gasifiers carried the extra mechanical complexity of cyclones to catch and recycle the fluidized particles.

Efficiency (20%):
Other than capacity, efficiency was the most important design variable considered for the gasifier. The cold gas efficiency is the ratio between the chemical energy content in the product gas compared to the chemical energy in the coal.

\[
\text{Cold Gas Efficiency} = \frac{\text{Product gas mass flow} \times \text{heating value} - \text{additional fuel mass flow} \times \text{heating value}}{\text{coal mass flow} \times \text{heating value}}
\]  

Eq 1
This category represented how effectively the gasifier was able to produce syngas from a given amount of coal. Entrained gasifiers, with high temperatures and pressures of operation, are able to attain high conversions of coal to syngas. Their cold gas efficiency is the highest at values ranging near 80%. Molten Bath gasifiers operate at similarly high temperatures and received the same score as the entrained systems. Fluidized bed gasifiers, having old gas efficiencies in the range of 70%, are also effective gasifiers due to its design which allows large surface area contact between the oxidant and coal. Although they are not as effective as the entrained and molten configurations, they received a score of 7 due to their proficiency. Fixed bed gasifiers, however, notorious for lower cold gas efficiencies (ranging around 60%) and higher amounts of tar, received a low score of 2.

Table 6: Feature comparison of the main gasifier types

<table>
<thead>
<tr>
<th>Attribute</th>
<th>Fixed Bed</th>
<th>Fluidized Bed</th>
<th>Entrained Bed</th>
<th>Molten Bed</th>
</tr>
</thead>
<tbody>
<tr>
<td>Maturity</td>
<td>Well established with many commercial designs</td>
<td>Less Developed than fixed bed</td>
<td>Less Developed than fixed bed</td>
<td>Least developed of all types</td>
</tr>
<tr>
<td>Complexity</td>
<td>High level of mechanical complexity (internal moving parts)</td>
<td>No internal moving parts. Design of distributor plate is critical</td>
<td>No internal moving parts. Geometry less complicated than fluidized bed</td>
<td>Normally less complex, although bath circulation increases complexity</td>
</tr>
<tr>
<td>Treatment</td>
<td>Requires sized coal, fines handled separately</td>
<td>Fines should be removed to prevent elutriation</td>
<td>No fines declined</td>
<td>No pre-treatment required</td>
</tr>
<tr>
<td>Product gas</td>
<td>Multi-stage cleanup required</td>
<td>Superb gas-solid contact</td>
<td>Sensible heat in has gas to be recovered - decreases efficiency</td>
<td>Entrainment of alkali particles can prevent usage of synthesis reactors</td>
</tr>
<tr>
<td>Ash disposal</td>
<td>Ash contains small amounts of carbon</td>
<td>Dry ash gravitates to bottom - contains more carbon</td>
<td>Increased thermal loss in ash</td>
<td>Ash contains fairly high carbon quantity and should be separated from catalyst which is recycled</td>
</tr>
<tr>
<td>Temperature</td>
<td>Counter current flow causes temperature gradient. Low thermal losses.</td>
<td>Can operate at lower temperatures</td>
<td>Highest temperatures of all categories</td>
<td>Temperatures achieved can be corrosive to refractory</td>
</tr>
</tbody>
</table>

We decided that an entrained flow gasifier would be the best choice to be used to the IGCC power plant using the decision matrix shown in Table 7. There are two main types of entrained flow gasification technology, dry processes with the licensors Shell, Prenflow, and GSP, and coal-slurry processes with licensors Texaco (now GE Energy) and E-gas.

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38 Wen & Chaung, 1979
39 Roberts, 2000
40 Roberts, 2000
Table 7: Decision matrix for gasifier designs

<table>
<thead>
<tr>
<th>Gasifier Bed Type</th>
<th>Cost</th>
<th>Extent Developed</th>
<th>Capacity</th>
<th>Temperatures</th>
<th>Pressures</th>
<th>Waste</th>
<th>Coal Flexibility</th>
<th>Complexity</th>
<th>Efficiency</th>
<th>Total Score</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fluidized</td>
<td>7</td>
<td>7</td>
<td>9</td>
<td>8</td>
<td>5</td>
<td>7</td>
<td>8</td>
<td>6</td>
<td>7</td>
<td>7.4</td>
</tr>
<tr>
<td>Entrained</td>
<td>6</td>
<td>7</td>
<td>10</td>
<td>4</td>
<td>6</td>
<td>8</td>
<td>9</td>
<td>8</td>
<td>8</td>
<td>8.0</td>
</tr>
<tr>
<td>Fixed</td>
<td>10</td>
<td>10</td>
<td>3</td>
<td>10</td>
<td>6</td>
<td>1</td>
<td>5</td>
<td>10</td>
<td>10</td>
<td>5.0</td>
</tr>
<tr>
<td>Molten Bath</td>
<td>4</td>
<td>1</td>
<td>3</td>
<td>1</td>
<td>10</td>
<td>10</td>
<td>10</td>
<td>2</td>
<td>8</td>
<td>6.0</td>
</tr>
</tbody>
</table>

Since the plant would be operating on a 1000 MW scale, we were aware that a large amount of coal would have to be put through our gasifier. Thus, our choice of gasifier was limited by the requirement for high capacity. Since fixed bed and molten bath gasifiers are considered semi-batch processes and characteristically low capacity, we decided that they, and similar systems, would be inadequate for our needs. A Texaco entrained flow design was consequently chosen due to its prevalence in industry, allowing for many operation references, and its ability to meet our fore-mentioned design requirements.

A single Texaco gasifier, though characterized by having the highest coal throughput at 2600 tons/day\(^{41}\), is unable to meet the 1000 MW requirement which would require a coal throughput of around 13,000 tons/day. Consequently, we decided that two 500 MW Texaco style gasifiers would be used in parallel. Texaco gasifiers typically run at only 250 MW, however, 500 MW gasifiers are an existing proprietary technology. Our design of the 500 MW gasifiers uses kinetics from the Texaco pilot plant research gasifier in order to scale up to a 500 MW system.\(^{42}\) Although the performance of the larger gasifier would be slightly different than the smaller gasifier, with lower carbon conversions typical with larger gasifiers, we anticipated these differences and are confident that our modeling in the scale up is accurate.

The Texaco gasifier utilizes a pressurized coal-water slurry and oxygen at least 95% purity as feed stocks for the reaction.\(^{43}\) It uses a single nozzle structure and simplified control systems compared to counterpart entrained flow gasifiers such as Shell, Prenflo, and E-gas.\(^{44}\) Also, the furnace temperature, which operates between 1400°C and 1600°C\(^{45}\), is lower than counterpart gasifiers leading to lower operating costs. Furthermore, the temperature of the furnace can be easily controlled by varying the flow rate of coal slurry, and oxygen. The Texaco gasifier design typically operates between 10 bar and

\(^{41}\) Controls
\(^{42}\) Wen & Chaung, 1979
\(^{43}\) Controls
\(^{44}\) Philips
\(^{45}\) Philips
70 bar. These high pressures of operation are conducive to both downstream separation and water gas shift processes. Typically, Texaco gasification systems lead to less fly ash in the crude gas, an advantageous aspect that leads to lower capital costs for separation equipment.

Coal treatment costs are lower in Texaco designs. The slurry of approximately 65 wt% solids is made via wet rod mills. The coal does not need to be dried, as in other gasification systems, and can be sent the mills directly from the delivery train. Also, the coal particle size requirement of <0.5 mm is bigger than other gasifiers. Since coal processing requirements on Texaco gasifiers are less stringent compared to other systems, it is possible to have larger production scales than other systems. A summary of all these gasifier types and characteristics are shown in Table 8.

Table 8: Comparison of entrained flow gasifiers.

<table>
<thead>
<tr>
<th>Type</th>
<th>Gasification Technology</th>
<th>Gasification Agent</th>
<th>Slag Discharge Style</th>
<th>Production Scale [tons/day]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Coal feed</td>
<td>Dry Coal</td>
<td>Liquid</td>
<td>2600</td>
<td>2500 2000</td>
</tr>
<tr>
<td>Coal Particle Size [mm]</td>
<td>&lt;0.1</td>
<td></td>
<td></td>
<td>2600 95% O₂ 2000 720 85% O₂ 720 95% O₂</td>
</tr>
<tr>
<td>Gasification Temperature [℃]</td>
<td>1400-1600</td>
<td></td>
<td></td>
<td>1500-1900</td>
</tr>
<tr>
<td>Gasification Agent</td>
<td>95% O₂</td>
<td></td>
<td></td>
<td>2600</td>
</tr>
<tr>
<td>Production Scale [tons/day]</td>
<td>2500</td>
<td></td>
<td></td>
<td>2000</td>
</tr>
</tbody>
</table>

2.2 Air or Oxygen for IGCC Feed

Gasification can operate with both air and oxygen as the reagent for oxidation. Considering, however, that air is composed of approximately 21% oxygen and 79% nitrogen, it would take much more air to achieve the same syngas product as it would if oxygen was used. Also, the extra component of nitrogen provides an additional heat sink that could make it more difficult to get the gasifier up to higher temperatures. This would be an especially relevant issue considering that fluidized and entrained gasifiers need to operate at higher temperatures in order to get the desired syngas product.

The primary goal of the IGCC system is to provide an effective way of sequestering carbon emissions. If air is used as the oxidant in the gasifier, the nitrogen composition of the syngas product would be very high, approximately 50 to 60 mol%. Consequently, separation of carbon dioxide would be increasingly difficult with the addition of nitrogen to the syngas product stream.

Using air would require a significantly larger gasifier to account for the higher oxidant flow rates. Also, the extra energy required to make up for the additional inert would be grossly parasitic upon the plants power production. Most importantly, the high content of nitrogen in the syngas product would overly

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46 Controls
47 Roberts, 2000
48 Controls
49 Controls
50 Visagie, 2008
complicate the separation processes required for the sequestration of carbon dioxide. Consequently, despite the added cost of an air separation unit, we decided that it would be more profitable to use a pure oxygen feed rather than an air as the oxidant. Even from a qualitative point of view, the capital costs that would be required for higher capacity gasification, heating, and separations when using air would far outweigh those needed for an oxygen feed.

2.3 Gasifier Modeling

2.3.1 Langmuir-Hinshelwood Kinetics

Two types of kinetics models are generally accepted throughout the literature to explain gasification reactions: the shrinking core model, and Langmuir-Hinshelwood kinetics. The strength of the shrinking core model lies in the low pressure atmospheric gasification systems while Langmuir-Hinshelwood kinetics, which account for pressure in the rate law, are effective in high pressure systems. Since the Texaco gasifier commonly operates at pressures of 40 bar, we decided to use Langmuir-Hinshelwood kinetics for our rigorous design model. Kinetic rate laws, based on Langmuir-Hinshelwood kinetics, were thus developed to explain the activity of the $\text{O}_2$-$\text{CO}_2$-$\text{H}_2\text{O}$ system. Although there are thirteen reactions occurring in the gasification system, shown in Table 3, four reactions are dominant and can be used for the design of the reaction unit. These reactions are shown in Table 9.

| Table 9: The reactions used in the development of the Langmuir-Hinshelwood model |
|---------------------------------|---------|
| $\text{C} + \frac{1}{2}\text{O}_2 \rightleftharpoons \text{CO}$ | (1) |
| $\text{C} + \text{O}_2 \rightleftharpoons \text{CO}_2$ | (2) |
| $\text{C} + \text{H}_2\text{O} \rightleftharpoons \text{CO} + \text{H}_2$ | (3) |
| $\text{C} + \text{CO}_2 \rightleftharpoons 2\text{CO}$ | (4) |

Of the four reactions describing the gasification system, two involve oxygen reacting with char to make carbon monoxide and carbon dioxide. These oxidation reactions can be described by the following rate model developed by Essenhigh.\(^{51}\)

$$r_{\text{O}_2} = \frac{6k_{a,y}\text{O}_2}{D_{\text{particle}}(1 + \frac{k_a}{k_d}y_{\text{O}_2})} \frac{\text{gmoles}}{\text{cm}^3\text{s}}$$

The oxygen-coal reactions typically react very quickly in the gasifier and can be explained by Langmuir Hinshelwood kinetic rate laws.\(^{52}\) Adsorption of the oxygen on the surface of the coal particle and desorption of the product CO and CO\(_2\) are the limiting step of the reaction. The rate constants of adsorption and desorption, for the kinetic rate law of oxygen consumption, can be modeled as a function of temperature by the Arrhenius equation. Our design utilized kinetic parameters developed by Essenhigh in 1991.

\(^{51}\) Essenhigh, 1991
\(^{52}\) Essenhigh, 1991
Table 10: Kinetic parameters for the carbon-oxygen reactions\textsuperscript{53}

<table>
<thead>
<tr>
<th>(k_0) [g/cm\textsuperscript{2}s]</th>
<th>(E_a) [kcal/mol]</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.14</td>
<td>10.04</td>
</tr>
<tr>
<td>2617</td>
<td>32.95</td>
</tr>
</tbody>
</table>

The reaction of oxygen with coal leads to a production of CO and CO\textsubscript{2}. The fraction of CO to CO\textsubscript{2} produced from the oxygen-coal reactions varies depending on the temperature of the reaction system and follows an Arrhenius type model.

\[
\frac{CO}{CO_2} = A e^{\left(-\frac{E}{RT}\right)}
\]

According to a study performed by Roberts in 2000 the parameters for equation 5 are:

\(A = 103.5\) and \(E = 50\) kJ/mol at pressures greater than 1 atm. \textsuperscript{54}

This model can, consequently, be used to predict the production of CO and CO\textsubscript{2} as a fraction of the total oxygen rate. As temperature increases, the rate of the CO reaction becomes faster compared to the rate of production of CO\textsubscript{2}. In Polymath these reactions were modeled by multiplying the rate of oxygen disappearance by a linear function of temperature predicting the fraction of each species in the product gas.

To develop expressions for the fraction of CO\textsubscript{2} and CO produced from the disappearance of oxygen, we plotted equation 5 which predicts the fraction of CO to CO\textsubscript{2} as a function of temperature. The combination of CO\textsubscript{2} and CO produced account for all of the oxygen consumed in the gasifier, and consequently their fractions must add to a total of 1 at every temperature. With this knowledge it is possible to develop linear expressions for the fraction of and CO and CO\textsubscript{2} produced from oxygen (Figure 15).

The following rate equations were developed by combining Equation 3 with the CO and CO\textsubscript{2} mole fraction relationships from Figure 15. \textsuperscript{55}

\[
r_{(CO)_1} = f(T)_{CO} \ast (-r_{O_2}) = (-0.0006T + 1.2165)(-r_{O_2}) \frac{gmol}{cm^3 s}
\]

\[
r_{(CO_2)_1} = f(T)_{CO_2} \ast (-r_{O_2}) = (0.0006 - 0.2165)(-r_{O_2}) \frac{gmol}{cm^3 s}
\]

\textsuperscript{53} Essenhigh, 1991
\textsuperscript{54} Roberts, 2000
\textsuperscript{55} Roberts, 2000
As temperature increases, the fraction of CO produced from oxygen-coal reactions also increases.

The rate of water and carbon dioxide reaction with coal can be modeled by the following Langmuir-Hinshelwood kinetics. Carbon dioxide is produced by the formation oxygen reactions which react rapidly to 100% conversion. The carbon dioxide then proceeds to react with coal to produce more carbon dioxide. The water, on the other hand, reacts with carbon to produce hydrogen and carbon dioxide.

\[ r_{H_2O} = \frac{-K_2P_{H_2O}}{1 + K_4P_{CO_2} + K_5P_{H_2O}} \]  \hspace{1cm} \text{Eq 7}

\[ r_{(CO_2)_2} = \frac{-K_1P_{CO_2}}{1 + K_4P_{CO_2} + K_5P_{H_2O}} \]  \hspace{1cm} \text{Eq 8}

\[ r_{(CO)} = \frac{2K_1P_{CO_2} + K_2P_{H_2O}}{1 + K_4P_{CO_2} + K_5P_{H_2O}} \]  \hspace{1cm} \text{Eq 9}
The kinetic parameter of the water and carbon dioxide reaction rates follow an Arrhenius type expression using the constants shown in Table 11.

### Table 11: Kinetic parameters for the steam, carbon dioxide, and carbon monoxide gasification reactions

<table>
<thead>
<tr>
<th>$A_i$ [Pa$^{-1}$ s$^{-1}$]</th>
<th>$E_{Ai}$ [J/mol]</th>
</tr>
</thead>
<tbody>
<tr>
<td>$K_1$</td>
<td>1.45E+01</td>
</tr>
<tr>
<td>$K_2$</td>
<td>6.47E+02</td>
</tr>
<tr>
<td>$K_4$</td>
<td>1.16E-05</td>
</tr>
<tr>
<td>$K_5$</td>
<td>5.59E-04</td>
</tr>
</tbody>
</table>

The final Polymath model fits industry data on Texaco gasifiers well. The deviation of our Polymath model from experimental results is no larger than 3% for molar composition of both carbon monoxide and hydrogen. The deviation carbon dioxide composition is slightly higher at 11.3% but this was deemed acceptable considering that the fraction of CO/CO$_2$ is an estimation and could be tuned to better fit experimental data. The Langmuir Hinshelwood kinetic model proved to be consistent with experimental data from a Texaco pilot gasifier.\(^{56}\)

### Table 12: Composition of syngas for Texaco Gasifier determined by Polymath simulation

<table>
<thead>
<tr>
<th>Molar Composition of Syngas [%]</th>
<th>CO</th>
<th>H2</th>
<th>CO2</th>
<th>CH4</th>
<th>H2S</th>
<th>Ar</th>
<th>H2O</th>
</tr>
</thead>
<tbody>
<tr>
<td>Wen Experimental(^{57})</td>
<td>50.71</td>
<td>35.79</td>
<td>13.14</td>
<td>0.09</td>
<td>0.03</td>
<td>0.24</td>
<td>0</td>
</tr>
<tr>
<td>Polymath Model (LH Kinetics)</td>
<td>52.2</td>
<td>35.1</td>
<td>11.8</td>
<td>XX</td>
<td>XX</td>
<td>XX</td>
<td>0.9</td>
</tr>
</tbody>
</table>

As show in Figure 16, the oxygen reactions proceed very quickly and reach 100% conversion within the first 5 cubic meters of the reaction vessel. The carbon dioxide and steam reactions then proceed at slower rates until complete conversion of coal occurs at approximately 30 cubic meters. This is consistent with theory which suggests that the oxygen reactions occur very quickly compared to the H2O-char and CO2-char reactions which proceed slower due to their lower diffusion rates.

Although Langmuir-Hinshelwood kinetics were effective at predicting the composition of syngas leaving the reactor, they had a drawback in their inability to account for the size of the reaction. It was necessary, to scale the gasifier volume in order to output a number consistent with literature.\(^{58}\) This shortcoming in the kinetic model of gasification prompted our team to consider other models, namely the shrinking core model.

---

\(^{56}\) Wen & Chaung, 1979  
\(^{57}\) Wen & Chaung, 1979  
\(^{58}\) Wen & Chaung, 1979
Figure 16: Polymath simulation of syngas composition traveling down the gasifier
2.3.2 Shrinking Core Model

Langmuir Hinshelwood kinetics were effective at predicting both final syngas composition and the kinetic rates of each species relative to each other. The Langmuir-Hinshelwood model, however, was inadequate in regard to entrained flow reactor sizing. In order to get a reliable estimate of the gasifier volume, it was necessary to model reaction kinetics in the context of particle fluid dynamics. The shrinking core model proved to be optimal for this scenario, where both the velocity of the coal particle down the gasifier, and the surface area of each coal particle for time in the reactor, were taken into account. Although Langmuir Hinshelwood kinetics showed better responses to different gasification pressures, the shrinking core model was superior in predicking carbon conversion, reactor volume, and particle velocity. The syngas compositions predicted by the shrinking core model were consistent with both experimental results and the compositions predicted by our Langmuir-Hinshelwood kinetic model. Consequently, the shrinking core model was chosen for the design of the Texaco entrained flow gasifier.

In shrinking core kinetics, the coal particle is modeled to disappear as it travels down the gasifier reacting with \( \text{O}_2 \), \( \text{H}_2\text{O} \), and \( \text{CO}_2 \). As the coal particle reacts away, the surface area of contact between char and reactant gases decreases until it becomes negligible at which point there is no longer any coal particle left to react. The shrinking core rates, which are dependent on available surface area, thus approaches zero when the entire coal particle is reacted away.

Regardless of kinetics, coal composition and quality is an important variable that determines the composition of syngas. Due to the considerable variety in coal feedstock, the composition of syngas varies widely. For this study, bituminous Illinois No. 6 Coal was used in the modeling of the gasification kinetics because its dry fuel analysis, Table 13, would be close to the coal that would be used in the scaled up one GW capacity IGCC.

<table>
<thead>
<tr>
<th>Coal Sample</th>
<th>Coal Quality</th>
<th>Dry Fuel Analysis, wt%</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>C</td>
</tr>
<tr>
<td>1</td>
<td>Illinois No. 6 Coal</td>
<td>74.05</td>
</tr>
<tr>
<td>2</td>
<td>Illinois No. 6 Coal</td>
<td>73.04</td>
</tr>
</tbody>
</table>

The Gasifier can be conceptually divided into three zones:

1. Pyrolysis and Volatile combustion
2. Combustion and Gasification
3. Gasification

---

Table 13: Typical Ultimate Analysis for the feedstock used in Texaco’s pilot plant tests

59 Wen & Chaung, 1979
Figure 17: Process schematic for Texaco gasifier, displaying stages of reaction in gasifier.

In Texaco style gasification, the 65 wt% coal slurry enters the gasifier with a 98 mol% oxygen stream at 90 °C. A high pressure nozzle pressurizes the feed stream to approximately 40 bar and expands the flow of the feed stream at a specific angle, ranging from 3-10 degrees, into the gasifier. This angle can be varied to optimize for mixing in the gasifier and increase carbon conversion. A study by Yu in 2009 provides models for multiple cases of expansion angle. The gasifier is maintained at a temperature of 1400 °C primarily by varying the oxygen/coal feed ratio.

Upon entrance to the gasifier, the coal undergoes pyrolysis to yield volatile compounds and char. The pyrolysis reactions can be represented as:

\[ C_{\alpha i}H_{\beta i}O_{\gamma i}N_{\delta i}S_{\varepsilon i}A \rightarrow C_{\alpha}H_{\beta}O_{\gamma}N_{\delta}S_{\varepsilon}A + \text{volatiles} \]

\[ \text{volatiles} = CO + H_2 + CO_2 + CH_4 + H_2S + Ar + tar \]

Highly dependent on the quality and type of coal, the values for \( \alpha, \beta, \gamma, \delta, \) and \( \varepsilon \) are variable. For Illinois No. 6 coal, however, typical values could be characterized by \( \alpha \approx 6.2, \beta \approx 6.3, \gamma \approx 0.1, \) and \( \varepsilon \approx 0.05 \) based upon the dry fuel analysis of the coal shown in table 13.

---

60 Yu, Jianjun, Liang, Guo, & Zhou, 2009
61 Yu, Jianjun, Liang, Guo, & Zhou, 2009
62 Wen & Chaung, 1979
Texaco style gasifiers, the oxygen and coal are both fed at the top of the reactor. Consequently, the pyrolysis zone is rich in oxygen and the volatile compounds released from coal are immediately oxidized. These oxidation reactions are highly exothermic and thus a large amount of heat is produced in the gas phase feeding the endothermic gasification reactions that occur farther down the gasifier.\textsuperscript{63}

Although the reaction kinetics in the gasifier are modeled using plug flow reactor expressions, the nature of mixing in the entrained flow reactor causes the temperature profile to follow CSTR (Constant stirred tank reactor) characteristics. The coal slurry and oxygen enter the gasifier near 300 degrees Celsius and spike to the temperature of the exiting syngas, 1400 °C. This spike in temperature is attributed to the exothermic oxidation reactions, pyrolysis, and good mixing.

\begin{figure}
\centering
\includegraphics[width=\textwidth]{temperature_profile.png}
\caption{Temperature profile down the entrained flow gasifier.}
\end{figure}

\textsuperscript{63} Roberts, 2000
The rate of pyrolysis and volatile release occurs so quickly, within 0.25 seconds, that the accuracy of the rate equation does not affect the outcome of the complete model significantly. The space time of the coal particles in the gasifier is 5 seconds and the gasification reactions are approximately 20 times slower than pyrolysis, so deviation in the pyrolysis rate equation is acceptable considering the overall system. The rate of char production, and volatile yield, can be estimated by the following expression developed by Wen and Chaung in 1979:

\[
\frac{dF_{\text{Volatiles}}}{dt} = 1.14 \times 10^5 \exp\left(-\frac{8900}{T}\right) (V^* - V) \text{ g/cm}^3\text{s}
\]

\[
V^* = V_{(1 \text{ atm})}^* (1 - c * \ln(P_{\text{total}}))
\]

Where \(V^*\) is the potential ultimate yield of volatiles at pressures ranging from 1 atm to 100 atm, and \(V\) is the apparent yield of volatiles at the pressure of interest, namely 40 bar. Although the rate of pyrolysis is a straightforward process to model, product compositions from pyrolysis are complex and extremely difficult to model theoretically. An experiment conducted by Loison and Chauvin (1964) provided product distribution data for char pyrolysis. This data indicated that by using material balances based on the dry fuel analysis of coal (C, H, O, N, and S) and knowledge of the C/CO\(_2\) and H\(_2\)O/CO\(_2\) ratios, satisfactory estimations could be made on the product distribution following coal pyrolysis. The yield of hydrogen is independent of coal rank while CO and CO\(_2\) increase with increasing volatile matter.

For this study, it was assumed that in pyrolysis CO\(_2\) accounted for 15% of the total volatile release rate while CO accounted for 9% of the total volatile release rate. The hydrogen in the coal was assumed to have been devolatilized fully, according to the dry proximate analysis of coal, and oxidized to water. Our design builds upon the work performed by Wen and Chaung (1979) and Loison and Chauvin (1964) predicting volatile composition after pyrolysis.

In the combustion and gasification stage the following reactions compete to form complexes with the char produced from pyrolysis:

\[
C + O_2 \rightarrow CO_2
\]

\[
C + \frac{1}{2}O_2 \rightarrow CO
\]

\[
C + H_2O \rightarrow H_2 + CO
\]

\[
C + CO_2 \rightarrow 2CO
\]

---

64 Wen & Chaung, 1979  
65 Wen & Chaung, 1979  
66 Wen & Chaung, 1979
The oxygen-char reactions occur fastest and to completion. In this char-O2 system, oxygen diffuses onto the surface of the char and reactions with carbon to form either carbon dioxide, or carbon monoxide. The relative rates of these oxygen reactions are highly dependent upon temperature and follow an Arrhenius type expression.\[^{67}\]

\[
\frac{CO}{CO_2} = A \exp \left( \frac{-E_a}{RT} \right)
\]

Eq 17

Where \( E_a = 50 \text{ kJ/mol}, A = 103.5, \) and \( R = 8.314 \text{ J/mol*K}. \[^{68}\]

As temperature increases, the rate of production of carbon monoxide, due to the oxygen-char reactions, increases while the rate of production of carbon dioxide decreases. At the operating temperature of 1400 °C, the rate of production of CO can be characterized as 58% of the oxygen disappearance rate while the rate of production of CO2 can be characterized as 42% of the oxygen disappearance rate. These fractions of CO and CO2 were determined from Figure 15, where the fractions of CO and CO2 from the char-oxygen reaction are shown as a function of temperature. Production of CO2 and CO was consequently modeled in Polymath by their relationship to O2 disappearance.

The rate expressions for oxygen, steam, and carbon dioxide, could be expressed by the following general form: \[^{69}\]

\[
\text{rate} = \frac{1}{k_{\text{diff}}} + \frac{1}{k_s y^2} + \frac{1}{k_{\text{dash}} (1 - 1)} (P_i - P_i^*) a \frac{g}{cm^3 s}
\]

Eq 18

\[
Y = \left[ \frac{1 - x}{1 - f} \right]^{1/3}
\]

Eq 19

Where \( k_{\text{diff}} \) is the gas film diffusion constant [g/cm²*atm*s], \( k_s \) is the char surface reaction rate constant [g/cm²*atm*s], \( k_{\text{dash}} \) is the ash film diffusion constant [g/cm²*atm*s], \( Y \) is the rate of carbon disappearance with respect to radius of char particle, \( x \) is the conversion any time after pyrolysis is complete, \( f \) is the conversion once pyrolysis is finished, and \( a \) is the contact area between solid and gas per unit volume of reactor [cm²/cm³]. \[^{70}\]

Due to the extremely high operating temperatures (1400 °C) the surface reaction for each system occurs rapidly. The limiting reaction, in the kinetic model, is absorption of each reactant molecule onto the surface of the char and desorption from the complex. This concept is illustrated by the diffusion coefficients that determine each rate equation.

\[^{67}\] Roberts, 2000
\[^{68}\] Roberts, 2000
\[^{69}\] Wen & Chaung, 1979
\[^{70}\] Wen & Chaung, 1979
The contact area between the solid char particles and reactant gases were modeled by the following expression. This equation is highly dependent on the solid particle velocity entering the reactor, $v_{si}$. The nozzle design determines entering solid velocity, and consequently has a great effect on the contact area between char particles and reactant gases. For the Polymath design, the optimal entrance velocity to the reactor, $v_{si}$, was determined to be 200 cm/s based upon a reactor cross sectional area, $A_t$, of 9.62 m$^2$ (reactor diameter of $D=3.5$ m) and a coal mass flow rate, $W_s$, of 74.59 kg/s. The optimal entrance velocity was determined from data provided by Wen and Chaung (1979) who simulated a Texaco pilot gasifier for entrance velocities ranging from 200 cm/s to 800 cm/s. Our Polymath analysis and the Wen study indicated that 200 cm/s would provide the best carbon conversion at our specified particle diameter of 0.05 cm. Average char density, $\rho_s$, was estimated to be approximately 1.81 g/cm$^3$ while the particle diameter, $d_p$, was a design specification for Texaco gasifiers at 0.05 cm. The following expressions that were used to design our Texaco gasifier were developed by Wen and Chaung (1979):

$$a = \left(\frac{W_s}{A_t v_{si}}\right) \left(\frac{6}{\rho_s d_p}\right)$$

Eq 20

The velocity of char particle traveling through the downflow entrained gasifier was modeled via rate relationships developed from Stokes law.

$$\frac{dz}{dt} = v_s$$

Eq 21

The solid entraining velocity was thus calculated as a function of length down the reactor with the following relationship.

$$v_s = v_{si}e^{-b\Delta t} + \left(v_g + v_f\right)(1 - e^{-b\Delta t})$$

Eq 22

For simplification, the velocity of the gas entering the gasifier, $v_g$, was assumed to be equal to the entrance velocity of the char particle, $v_{si}$, at 200 cm/s. $\Delta t$ represents the solid residence time, an important design feature calculated from the Polymath model. For the ultimate design of the gasifier, this residence time was calculated to be 5.1 s. The constant, $b$, has units s$^{-1}$ and consequently becomes dimensionless when multiplied by the solid residence time in Equation 22. This parameter compares the solid particle size and density to the viscosity of the gases in the gasifier.

$$b = \frac{18\mu}{\rho_s d_p^2}$$

Eq 23

The density and viscosity of the gases in the gasifier were estimated to be $\rho_g = 7.796E-3$ g/cm$^3$, and $\mu = 6.3E-4$ P, while acceleration due to gravity was approximated to be $g=9810$ cm$^2$/s. The gas properties were estimated in UNISIM at 1400 °C and 40 bar.

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71 Wen & Chaung, 1979

72 Controls
The velocity term, $v_t$, accounts for the velocity of the solid particle compared to the density of the gases in the gasifier and the direction of flow in the gasifier. In our design, downflow gasification, the $v_t$ parameter is positive and adds to the gas velocity term, $v_g$ (Equation 22). Alternatively, in upflow gasification (See section 2.1 – Types of Gasifiers) the $v_t$ term is negative because the coal particles must move against gravity.

$$v_t = \frac{(\rho_s - \rho_g)d_p^2g}{18\mu} \quad \text{Eq 24}$$

The rate of each reaction system: Char-O$_2$, Char-H$_2$O, and Char-CO$_2$, was represented by Equation 25. Each system used the same form for the rate equation, but had different parameters based upon the characteristics of surface reaction and diffusion for that particular system.

$$rate = \frac{1}{k_{diff}} + \frac{1}{k_sY^2} + \frac{1}{k_{dash}Y^{\frac{1}{2}}\left(\frac{c}{c^*}\right)}(P_i - P_i^*)a \left(\frac{g}{cm^3s}\right) \quad \text{Eq 25}$$

$$Y = \left[\frac{1-x}{1-f}\right]^{1/3} \quad \text{Eq 26}$$

The following parameters were used in our Polymath model of the gasification kinetics. Each of the reaction systems refers back to equation 19.

1. **Char-O$_2$ Reaction:**

   $$k_s = 8710\exp\left(-\frac{17967}{T}\right) \quad \text{Eq 27}$$

   $$k_{diff} = \frac{\left(0.292\varphi\left(\frac{4.26}{T}\right)^{1.75}\right)}{\left(\frac{T}{1800}\right)} \quad \text{Eq 28}$$

   Where $\varphi$ is the mechanism factor based on the stoichiometric relation between Oxygen and char. Due to the complexity of char structures, the stoichiometric relations for carbon reacting with oxygen are not a simple 1:1 ratio.

   $$\varphi = \left[\frac{(2Z + 2) - Zd_p - 0.005}{0.095}\right] \quad \text{Eq 29}$$

   $$\text{for } 0.005cm \leq d_p \leq 0.1cm$$

---

73 Wen & Chaung, 1979
\[ Z = \frac{[CO]}{[CO_2]} = 2500 \exp \left( \frac{-6249}{T} \right) \]  

Eq 30

Where \( T \) is in units of Kelvin.

\[ P_i - P_i^* = P_{O_2} \]  

Eq 31

Where pressure is in units of atm.

2. **Char-Steam Reaction:**

Much like the oxygen-char reactions, H2O-char reactions do not follow a simple 1:1 stoichiometric ratio. The stoichiometric relationship between char and steam was consequently determined by trial-and-error. The Polymath model was run multiple times at different stoichiometric ratios for carbon and oxygen until hydrogen production and H2O consumption were consistent with literature. This relationship was determined to be 6:1, H2O:char.

\[ k_s = 247 \exp \left( \frac{-21060}{T} \right) \]  

Eq 32

\[ k_{diff} = \frac{10 \times 10^{-4} \left( \frac{T}{2000} \right)^{0.75}}{P_{total}d_p} \]  

Eq 33

\[ k_{eq} = \exp \left( 17.644 - \frac{30260}{1.8T} \right) \]  

Eq 34

\[ P_i - P_i^* = P_{H_2O} - \frac{P_{H_2} - P_{CO}}{k_{eq}} \]  

Eq 35

3. **Char-CO2 reaction:**

\[ k_s = 247 \exp \left( \frac{-21060}{T} \right) \]  

Eq 36

\[ k_{diff} = \frac{7.45 \times 10^{-4} \left( \frac{T}{2000} \right)^{0.75}}{P_{total}d_p} \]  

Eq 37

\[ P_i - P_i^* = P_{CO_2} \]  

Eq 38

The Polymath gasifier design model was consistent with literature. Pyrolysis, and oxidative reactions, occur rapidly and reach completion within roughly 10% of the final reactor volume. The slower

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\(^{74}\) Wen & Chaung, 1979

\(^{75}\) Wen & Chaung, 1979
gasification reactions, primarily the H2O-char and CO2-char system, proceed gradually throughout the entire gasifier volume. Figure 19 illustrates these concepts.

Table 14: Comparison of syngas product composition and carbon conversion with literature and industry Texaco Gasifiers.76

<table>
<thead>
<tr>
<th>Molar Composition of Syngas</th>
<th>CO</th>
<th>H2</th>
<th>CO2</th>
<th>CH4</th>
<th>H2S</th>
<th>Ar</th>
<th>H2O</th>
<th>NH3</th>
<th>Xc [%]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Wen Experimental</td>
<td>50.71</td>
<td>35.79</td>
<td>13.14</td>
<td>0.09</td>
<td>0.03</td>
<td>0.24</td>
<td>0</td>
<td>XX</td>
<td>92.7</td>
</tr>
<tr>
<td>Wen Literature Model</td>
<td>47.86</td>
<td>37.3</td>
<td>14.45</td>
<td>0.055</td>
<td>0.074</td>
<td>0.26</td>
<td>0</td>
<td>XX</td>
<td>94.87</td>
</tr>
<tr>
<td>Star Controls</td>
<td>54</td>
<td>34</td>
<td>11</td>
<td>0.01</td>
<td>0.3</td>
<td>XX</td>
<td>0.69</td>
<td>XX</td>
<td>96</td>
</tr>
<tr>
<td>Polymath Model (LH Kinetics)</td>
<td>52.2</td>
<td>35.1</td>
<td>11.8</td>
<td>XX</td>
<td>XX</td>
<td>XX</td>
<td>0.9</td>
<td>XX</td>
<td>100</td>
</tr>
<tr>
<td>Polymath Model (Shrinking Core)</td>
<td>50.2</td>
<td>34.03</td>
<td>12.3</td>
<td>XX</td>
<td>0.33</td>
<td>0.67</td>
<td>2</td>
<td>0.48</td>
<td>95.03</td>
</tr>
</tbody>
</table>

The char reactions with H2O, O2, and CO2 have the largest influence on gasifier design and consequently were focused upon the most. Other species constituting the syngas, such as H2S, CH4, and NH3, are important for separation processes downstream from the gasifier. The species, however, occur in such small concentrations that assumptions could be made to simplify the design process.

Assumptions:

1. Hydrogen Sulfide Production
   The production of Hydrogen Sulfide was modeled by completing reacting away all sulfur species in the coal (according to the dry fuel analysis) with hydrogen during pyrolysis. This assumption, although not theoretically perfect, is close enough to reality to be considered acceptable for simplified design.

2. Ammonia Production
   Nitrogen species in the coal were modeled in a similar way by reacting away all nitrogen species in the coal with hydrogen. This assumption is accurate because most of the nitrogen in coal is oxidized to NOx and then promptly reduced to NH4 during the gasification stage due to the high concentration of hydrogen. The nitrogen species consequently leaves the gasifier as ammonia.

3. Methane Production
   Methane production from char-Hydrogen reactions was ignored in the Polymath model because they do not occur at a high enough rate to significantly affect hydrogen concentration. Also, methane occurring in such small concentrations, typically 0.01 mol%, does not significantly affect downstream processes. For a perfect model, CH4 production could be modeled, but we

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76 Wen & Chaung, 1979
determined that methane could be removed from the gasification model without significantly affecting the overall design process in order to streamline Polymath calculations.

Overall, the Polymath simulation was an accurate predictor of CO, CO2, and H2 production. Also, carbon conversion for the Texaco gasifier was modeled accurately at 95%. This is consistent with experimental results in literature\textsuperscript{77}, and experiential results from industry.\textsuperscript{78}

\textsuperscript{77} Wen & Chaung, 1979
\textsuperscript{78} Controls
Figure 19: Composition of syngas as a function of reactor volume. Pyrolysis and oxidation occurs rapidly with the first 5 m$^3$ of the reactor.
Figure 20: Carbon conversion as a function of reactor volume is consistent with literature
Figure 21: Process Flow Diagram for Coal Treatment and Gasification Block
2.4 Detailed Process Description

Figure 21 is a comprehensive flow sheet of the coal processing and gasification process that gives a detailed view of the equipment involved.

2.4.1 Coal Handling

In the coal handling section, coal is unloaded from the train into the dump hopper where it is loaded on a conveyor belt. The conveyor belt transfers the coal to an elevator which transports to coal into the silo. The coal then undergoes magnetic separation to take out large metal impurities, and is crushed by a hammer mill. A dust control system insures that excessive particles do not cause explosion during the hammer mill crushing. After the initial crushing, a conveyor belt transfers the coal to a second elevator which lifts the crushed coal into the mill feed hopper. The coal is fed into the wet rod mill along with demineralized water and 65 wt% coal-water slurry is produced. The coal slurry is then pressurized by pump P-100G for subsequent introduction into the gasifier. The coal handling process, although covered in the process description, was considering outside the bounds of the IGCC design. Coal handling costs were estimated and detailed in Section 2.7.1 – Coal Handling Costing, but a full design was determined to be out of the scope of our project.

2.4.2 Pump P-100G

P-100G is a stainless steel horizontal split case pump that pressurized the coal slurry from the wet rod mill from 14.7 psi to 518 psi to be introduced into the gasifier R-100G. Due to the highly corrosive nature of the coal slurry, the pump motor is a totally enclosed, fan cooled enclosure.

2.4.3 Gasifier R-100G

R-100G is a vertical reaction vessel that receives a 65 wt% coal slurry and 98 mol% Oxygen feed at 212°F and 1000 psi. The gasifier operates at a pressure of 580 psi and 2550 F. The coal undergoes pyrolysis and multiple gasification reactions to produce syngas. The syngas from gasifier R-100G leaves at 2550 F and enters the radiant syngas cooler (E-100G) to be cooled to 1380 F.

2.4.4 Radiant Syngas Cooler E-100G

E-100G is a radiant syngas cooler that exchanges heat between water and syngas. The vertical tube lined pressure vessel is positioned directly below the gasifier and cools syngas from 2550 F to 1380 F. The hot syngas enters at the top of the vessel and travels downward where it is contacted with the slag water quench. The syngas flow is then redirected along the outside of the vertical tube lining where it is further cooled until it exits out of the top of the reaction vessel. The cooled syngas then progresses to convective heat exchanger E-101G where it is cooled to 86 F.
2.5 Radiant Syngas Cooler Design

The radiant syngas cooler (RSC) is positioned directly below the gasification vessel and functions to cool the hot syngas so it may be sent through sulfur removal systems. Syngas exits the gasifier and enters the radiant syngas cooler at approximately 1400 °C. The gas is cooled to 750°C in the RSC where it then enters a convective syngas cooler (CSC), which is a standard shell and tube heat exchanger.

The hot syngas enters at the top of the RSC where flows through the center of the vessel and is cooled by a bundle of tubes arranged cylindrically to make a wall (Figure 23). The syngas flows lengthwise along the tubes until reaches the slag pool located at the bottom of the RSC (Figure 22). The syngas flow is then redirected upward along the outside of the cylindrical tube wall. The syngas finally exits out of the top of the RSC. Liquid slag from the gasifier coats the refractory lining at the top of the RSC, and the
tube lining. The slag flows down the RSC wall where it is caught in the slag quench pool. The slag is collected, crushed, and sold to concrete and roofing companies.\textsuperscript{79}

\begin{figure}
\centering
\includegraphics[width=\textwidth]{figure23}\caption*{Figure 23: Design and flow description of the radiant syngas cooler.\textsuperscript{80}}
\end{figure}

\textsuperscript{79} Roberts, 2000
\textsuperscript{80} Yu, Jianjun, Liang, Guo, & Zhou, 2009
The Radiant Syngas Cooler design was largely based on the existing radiant syngas cooler from the Polk 250MW IGCC Gasification Plant. Height, weight, and diameter were all taken from that design. Heat exchange area required for our design was in turn estimated by modeling the RSC as a double pass shell and tube heat exchanger with syngas on the shell side and water on the tube side.

The specified design is for syngas entering at 1400°C to be cooled to 700°C at 580 psi. Water enters at 30°C and leaves as steam at 500°C. RSC technology is highly proprietary and it is difficult to acquire specific design for the vessel. The general heat exchange design specifications are as follows:

RSC Height = 139 ft

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81 Kraft, Alexander, Fry, Thomas, & Albrecht, 2009
82 Jenkins, 1994
RSC Vessel Diameter = 26 ft
RSC Weight = 1.63 x 10^6 lb
RSC Heat Exchange Area = 1.71 x 10^4 ft^2

The radiant syngas cooler is a vertically positioned, tube lined pressure vessel. Due to the corrosive nature of the syngas traveling through the cooler, materials of construction will be ZIRCHROM 90 refractory brick positioned near the top of the cooler, Haynes HR-160 Alloy coating for the tubes and walls of the vessel, Croloy underneath the alloy coating, all encased by a Carbon Steel shell.

ZIRCHROM 90 is a refractory brick with a zirconia, alumina, and chromium oxide basis. It is sold by Saint-Gobain Ceramics (SGC) and is the common industrial material used for the high temperature and corrosive environments in gasification.

<table>
<thead>
<tr>
<th>CHARACTERISTICS</th>
<th>Standards or methods</th>
<th>Units</th>
<th>Typical values</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>CHEMICAL ANALYSIS</strong></td>
<td>FLUO X or ICP</td>
<td>%</td>
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</tr>
<tr>
<td>Cr₂O₃</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>ZrO₂</td>
<td></td>
<td></td>
<td>7</td>
</tr>
<tr>
<td><strong>PHYSICAL CHARACTERISTICS</strong></td>
<td>ISO 5017</td>
<td>g·cm⁻³</td>
<td>4.2</td>
</tr>
<tr>
<td>- Bulk density</td>
<td>ISO 5017</td>
<td>%</td>
<td>16</td>
</tr>
<tr>
<td>- Apparent porosity</td>
<td>EN 993-5</td>
<td>MPA</td>
<td>120</td>
</tr>
<tr>
<td>- Crushing strength at ambient temperature</td>
<td>ASTM C 583</td>
<td>MPA</td>
<td>10</td>
</tr>
<tr>
<td>- Modulus of rupture at 1500°C</td>
<td>ISO 3187</td>
<td>%</td>
<td>&lt; 0.4</td>
</tr>
<tr>
<td>- Creep under load of 0.2 MPa at 1500°C between 5 h and 25 hours</td>
<td>pr. EN 993-19</td>
<td>10⁻⁶K⁻¹</td>
<td>6.8</td>
</tr>
<tr>
<td>- Coefficient of thermal expansion</td>
<td>ISO 8884-2</td>
<td>W·m⁻¹·K⁻¹</td>
<td>2.4</td>
</tr>
<tr>
<td>- Thermal conductivity at 1000°C</td>
<td>By attribute</td>
<td>%</td>
<td>± 1</td>
</tr>
<tr>
<td>- Tolerances of dimensions</td>
<td>AQL 6.5 %</td>
<td>mm</td>
<td>± 1.5</td>
</tr>
</tbody>
</table>

Haynes Hr-160 alloy is a Ni-CO-Cr-Si alloy that is easily fabricated using both gas-tungsten-arc welding (TIG) and gas-metal-arc welding (MIG) processes. It performs 17x better than stainless steels at

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83 Saint-Gobain
84 Saint-Gobain
temperatures ranging from 980 °C to 1100 °C in resisting the attack of corrosive sulfur compounds. This alloy is ideal for ease of fabricating the RSC tubes while maintaining performance at high temperatures.

Croloy is a corrosion resistant Chromium-molybdenum alloy steel. It is highly resistance to corrosion and is suitable for the high temperature applications of gasification. This material is used to resist corrosion on the water side of the RSC tubes. These materials of construction were chosen, as opposed to stainless steel or other alloys, because of their resilience to high temperature, high corrosion environments, and their ease of fabrication.

2.6 Convective Syngas Cooler Design

The convective syngas cooler directly follows the radiant syngas cooler and functions to further cool the syngas from approximately 700°C to a manageable temperature of 35 °C for sulfur and CO2 separations. Water enters at 30°C and leaves at 45°C.

The convective syngas cooler was designed as a one pass triangular pitch shell and tube heat exchanger. The shell diameter was specified to be 37 inches, and the tube diameter to be ¾ inches. The heat exchanger uses a total of 1074 tubes. See Appendix examples for further detail on the heat exchanger design. The final heat exchange area is 834 m².

2.7 Economics: Gasifier and Coal Handling

The majority of costing was conducting using the Guthrie method outlined in Seider. However, this resource did not include methods specifically for estimating costs of gasification vessels and coal treatment equipment. The costs of these units were consequently calculated by plotting historical costs and developing regressions. These regressions provided the capital cost of the unit as a function of coal feed rate.

2.7.1 Coal Handling Costing

The cost for the wet rod slurry mills to make coal slurry were calculated by investigating the cost for past coal plants. Coal handling includes movement from train to mill, coal grinding, and slurry preparation. The plot included seven data points collected in 2000 for coal treatment units processing from 2000 to 5000 tons coal/day.

85 Haynes International
86 Kraft, Alexander, Fry, Thomas, & Albrecht, 2009
87 Haynes International
Since the regression was developed for units in the year 2000, it was necessary to adjust the capital cost for the year 2015 when our IGCC would be built. The chemical engineering cost index was consequently estimated to be 574 for the year 2015. The capital cost estimated from the regression was thus multiplied by the factor $\frac{574}{394}$ which adjusts from year 2000 dollars to year 2015 dollars.

\[
C_{CH} = 9.92 \times m_{coal}
\]

Rubin 2007 Eq 21
Combined Purchase Cost for Coal Handling/Treatment

\[
C_{CH} = \left(\frac{574}{394}\right) \times 9.314 \times 10^7
\]

Coal Handling Costs were estimated to be $93,140,000 for one 500MW train.

2.7.2 Gasifier and Radiant Syngas Cooler Costing

The cost for the gasifier and radiant syngas cooler were calculated by investigating the cost for past coal plants. These historical costs were plotted and a regression was developed to estimate capital cost of the gasification system as a function of gasifier coal feed rate.

---

Rubin, Berkenpas, Frey, Chen, McCoy, & Zaremsky, 2007
Since the regression was developed in terms of year 2000 dollars, it was necessary to adjust to year 2015 dollars. The projected chemical engineering cost index for the year 2015 was estimated to be 574 based on historical trends of the cost index.

\[
C_G = 15.88 \cdot m_{\text{coal}}^{0.943} \quad \text{Rubin 2007 Eq 22}
\]

\[
C_{R100G} = \left( \frac{574}{394} \cdot C_G^{1000} \right) = 9.044 \times 10^7
\]

The capital cost of a 500MW gasifier and radiant syngas cooler was determined to be $90,440,000.

---

Rubin, Berkenpas, Frey, Chen, McCoy, & Zaremsky, 2007

D.O.E
3 Water Gas Shift Block

The water gas shift (WGS) reaction is an industrial process where steam is mixed with carbon monoxide to produce hydrogen and carbon dioxide. It is both reversible and exothermic with a $\Delta H_{\text{rxn}} = -41$ KJ/mol. From Le Chatlier’s principle, and considering the exothermic properties of the water gas shift, the forward reaction is preferred at lower temperatures while higher temperatures favor the reverse.

$$\text{CO} (g) + \text{H}_2\text{O} (v) \leftrightarrow \text{CO}_2 (g) + \text{H}_2 (g)$$

There are multiple side reactions associated with the water gas shift reaction. These compete with the desired reaction to yield carbon and methane. The undesired carbon that could result from by-product reactions has a tendency to block active catalyst sites resulting in fouling and an increased pressure drop across the reactor. These undesirable reactions are only an issue at low H2O/CO feed ratios, particularly when they are below values of 2.

Table 16: Proposed Side Reactions competing with the water gas shift

<table>
<thead>
<tr>
<th>Possible Side Reactions of the WGS Reaction</th>
</tr>
</thead>
<tbody>
<tr>
<td>2CO $\equiv$ C + CO$_2$</td>
</tr>
<tr>
<td>CO + H$_2$ $\equiv$ C + H$_2$O</td>
</tr>
<tr>
<td>CO$_2$ + 2H$_2$ $\equiv$ C + 2H$_2$O</td>
</tr>
<tr>
<td>2CO + 2H$_2$ $\equiv$ CO$_2$ + CH$_4$</td>
</tr>
<tr>
<td>CO + 3H$_2$ $\equiv$ CH$_4$ + H$_2$O</td>
</tr>
<tr>
<td>CO$_2$ + 4H$_2$ $\equiv$ CH$_4$ + 2H$_2$O</td>
</tr>
<tr>
<td>C + 2H$_2$ $\equiv$ CH$_4$</td>
</tr>
<tr>
<td>4CO + 2H$_2$O $\equiv$ CH$_4$ + 3CO$_2$</td>
</tr>
<tr>
<td>CH$_4$ + 2CO $\equiv$ 3C + 2H$_2$O</td>
</tr>
<tr>
<td>CH$_4$ + CO$_2$ $\equiv$ 2C + H$_2$O</td>
</tr>
</tbody>
</table>

The effect of temperature on the water gas shift is better understood by observing the trend of the Gibbs free energy and the equilibrium constant of the reaction. At approximately 1100 K the Gibbs free energy shifts from a negative to a positive value. The equilibrium constant, on the other hand, decays to an asymptote at higher temperatures.

91 Bustamante, 2004  
92 Carbo, Jansen, Dijkstra, van den Brink, & Verkooijen, 2007  
93 Callaghan, 2006  
94 Callaghan, 2006
Figure 27: Gibbs Energy of the Water Gas Shift as a function of temperature \(^{95}\)

Figure 28: Equilibrium constant of the water gas shift as a function of temperature \(^{96}\)

The tendency for the water gas shift to favor the reverse reaction at higher temperatures is important because it limits the conversion achievable in the water gas shift reactors at high temperatures.

In order to surpass the kinetic limitation at high reaction temperatures, it is necessary to use multiple adiabatic reactors with inter-stage cooling. Typically, the process is separated into two stages: the high temperature shift and the low temperature shift. The high temperature shift (HTS) reactors normally operate at temperatures ranging from 350-600°C, while the low temperature shift (LTS) reactors usually operate at temperatures ranging from 150-400°C. Before each reactor, steam is mixed with the product gases to cool the system and provide additional water for the reaction. \(^{97}\)

\(^{95}\) Callaghan, 2006  
\(^{96}\) Callaghan, 2006  
\(^{97}\) Callaghan, 2006
The kinetics associated with the water gas shift reaction have been modeled as pseudo-first order, power law, and Langmuir-Hinshelwood depending on the proposed mechanisms. Altogether, research performed by multiple groups suggests that the water gas shift reaction occurs largely via three particular reaction mechanisms: The redox mechanism, the formate mechanism, and the carbonate mechanism.

3.1. Proposed Mechanisms

3.1.1. Redox Mechanism of the Water Gas Shift Reaction

The redox mechanism proposed by Temkin’s research group involved two steps where water is first absorbed into the catalyst. The oxygen from the water molecule bonds to the catalyst site releasing hydrogen gas. Carbon monoxide then reacts with the oxygen on the surface of the catalyst and carbon dioxide is desorbed.

<table>
<thead>
<tr>
<th>Two-Step Redox Reaction Mechanism</th>
</tr>
</thead>
<tbody>
<tr>
<td>( H_2O + S \rightleftharpoons O\cdot S + H_2 ) (527)</td>
</tr>
<tr>
<td>( O\cdot S + CO \rightleftharpoons CO_2 + S ) (528)</td>
</tr>
<tr>
<td>( CO + H_2O \rightleftharpoons CO_2 + H_2 ) (OR)</td>
</tr>
</tbody>
</table>

The rate law associated with the redox mechanism takes the following form:

\[
r = kP_{CO} \left( \frac{P_{H_2O}}{P_{H_2}} \right)^{1/2} (1 - \beta) \\
\]

where \( \beta = \frac{P_{CO} P_{H_2}}{K_{P_{CO}} P_{H_2O}} \)

3.1.2. Formate Mechanism of the Water Gas Shift Reaction

The formate mechanism involves absorption of both carbon monoxide and water onto the surface of the catalyst. Two surface reaction mechanisms are proposed. In the first surface reaction a water
molecule already bonded to a single site is ripped apart by another catalyst site to yield an OH group on one site and the remaining H on the other. Once split, a carbon monoxide molecule and the OH group react on the surface of the catalyst to form HCOO. The desorption of HCOO involves a reaction to yield CO\(_2\) and a hydrogen atom combining with the bound hydrogen from the first surface reaction to yield H\(_2\).

The following rate law was developed by the Campbell research group and is associated with the formate mechanism:\(^{103}\)

\[
r = \frac{kP_{CO}P_{H_2O}}{(1 + K_1P_{CO} + K_2P_{H_2O} + K_3P_{CO_2} + K_4P_{H_2})^2(1 - \beta)}
\]

\(^{104}\) Callaghan, 2006

### 3.1.3 Carbonate Mechanism of the Water Gas Shift Reaction

In the carbonate mechanism water molecules are absorbed onto the catalyst surface and ripped apart as OH and H. Once absorbed, the OH groups are pulled apart even more to yield an oxygen atom on one site and a hydrogen atom on a different one. Carbon monoxide in the gas phase is then able to react directly with two oxygen atoms on adjacent sites to yield CO\(_3\). The carbon trioxide is desorbed as carbon dioxide leaving with the other oxygen atom remaining on the site for more reactions with carbon monoxide.

\(^{103}\) Callaghan, 2006
\(^{104}\) Callaghan, 2006
The carbonate mechanism can be described by the following rate law developed by the Lund group:\textsuperscript{105}

\[
r_{\rho} = C_{\rho} \Delta_{\rho} \exp \left( - \frac{E_{\rho}}{R_{\text{gas}} T} \right) \left[ \frac{\tilde{\sigma}_{\rho} \prod_{i=1}^{l} a_i^{-r_{\rho i}} - \exp \left( \frac{\Delta S_{\rho} (T)}{R_{\text{gas}}} \right) \exp \left( - \frac{\Delta H_{\rho} (T)}{R_{\text{gas}} T} \right)}{\exp \left( \frac{\Delta S_{\rho} (T)}{R_{\text{gas}}} \right) \exp \left( - \frac{\Delta H_{\rho} (T)}{R_{\text{gas}} T} \right)} \right]
\]

\[
\Delta S_{\rho} (T) = \sum_{i_{fp}} v_{i_{fp}} S_{i_{fp}}^0 (T) + \sum_{i_{fp}} v_{i_{fp}} S_{i_{fp}}^0 (T_{\text{ref}}) - \sum_{i_{fp}} v_{i_{fp}} S_{i_{fp}}^0 (T_{\text{ref}}) - \sum_{i_{fp}} v_{i_{fp}} S_{i_{fp}}^0 (T_{\text{ref}})
\]

\[
\Delta H_{\rho} (T) = \sum_{i_{fp}} v_{i_{fp}} \Delta H_{i_{fp}}^0 (T) + \sum_{i_{fp}} v_{i_{fp}} \Delta H_{i_{fp}}^0 (T_{\text{ref}}) - \sum_{i_{fp}} v_{i_{fp}} \Delta H_{i_{fp}}^0 (T_{\text{ref}}) - \sum_{i_{fp}} v_{i_{fp}} \Delta H_{i_{fp}}^0 (T_{\text{ref}})
\]

\textsuperscript{105} Callaghan, 2006
\textsuperscript{106} Callaghan, 2006
3.2 Effect of H$_2$S on Water Gas Shift Catalysts

Many metal based catalysts may be used for the water gas shift, including: Cu, Fe, Ni, Pd, Pt, Rh, and Ru. The most common catalyst used for the HTS in industry, however, is magnetite: Fe$_3$O$_4$. One catalyst used commercially is 55 wt% Fe, 6 wt% Cr, with the balance being any typical catalyst substrate such as metals and metal oxides. The catalyst is available as a tablet, ring, or pellet\textsuperscript{107} and was used for the purposes of this study.

In IGCC power plants with integrated water gas shift and pre-combustion CO$_2$, the syngas must be desulfurized to avoid corrosion and fouling in downstream equipment, as well as prevent poisoning of catalysts. Desulfurization is placed upstream of the WGS reactors since H$_2$S will poison catalysts used in both the HTS and LTS reactors.

Hydrogen sulfide, or H$_2$S, causes catalyst fouling by binding with active sites and blocking water and carbon monoxide from binding, and also by changing the structure of the catalyst itself. In the following figure, a comparison of the Fe 2p regions of the X-ray photoelectron spectra of hydrogen sulfide reduced samples of a HTS catalyst are shown\textsuperscript{108}:

\begin{center}
\includegraphics[width=0.7\textwidth]{figure29.png}
\end{center}

\textit{Figure 29: Fe 2p region of an Fe-Al-Cu catalyst after different exposures of hydrogen sulfide.}

\textsuperscript{107} Callaghan, 2006
\textsuperscript{108} Zhang, Oxkan 2007
As seen in Figure 29, increasing the concentration of hydrogen sulfide exposure to the water gas shift catalyst deactivated it by reducing the Fe$_2$O$_3$ present in the catalyst to Fe$_3$O$_4$ on the surface. Not only was the original iron oxide reduced, in high concentrations of hydrogen sulfide (1000 ppm) a clear FeS$_2$ phase is observed. Formation of FeS$_2$ is closely related to a decline in reaction activity$^{109}$.

Usually, the HTS and LTS reactors use different catalysts. Iron-based and copper-based catalysts are typically used industrially for HTS and LTS operations, respectively$^{110}$. In the case of coal-derived syngas where the CO concentration is relatively high (40-60%), a considerably high degree of shifting is required. For our purposes, it was necessary to find catalysts for both the HTS and LTS and kinetic data for each.

Several catalysts were researched in order to find one with a good kinetic model that could be easily entered into UNSIM for design. The following table gives specifications for two commercially available high temperature iron based catalysts for the HTS reactor$^{111}$:

Table 20: HTS Catalyst Specifications

<table>
<thead>
<tr>
<th>Catalyst</th>
<th>HTC1</th>
<th>HTC2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Composition</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fe$_2$O$_3$</td>
<td>80–90%</td>
<td>80–95%</td>
</tr>
<tr>
<td>Cr$_2$O$_3$</td>
<td>8–13%</td>
<td>5–10%</td>
</tr>
<tr>
<td>CuO</td>
<td>1–2%</td>
<td>1–5%</td>
</tr>
<tr>
<td>Shape [commercially available]</td>
<td>Pellet</td>
<td>Pellet</td>
</tr>
<tr>
<td>Size (diameter x height) [commercially available]</td>
<td>6 mm x 6 mm</td>
<td>6 mm x 6 mm</td>
</tr>
</tbody>
</table>

As a result of experimentation, it was found that a power-law rate expression for the WGS reaction over HTC1 and HTC2 at 450 °C under select inlet conditions can be described as follows:

For HTC1

\[
R = 10^{2.845 \pm 0.03} \exp \left( \frac{-111 \pm 2.63}{RT} \right) \left( \frac{p_{CO}^{1.0 \pm 0.03}}{p_{CO}^{2.0 \pm 0.36}} \right) p_{CO}^{0.36 \pm 0.04} p_{H_2}^{0.9 \pm 0.07} \left( 1 - \frac{1}{K} \frac{p_{CO} p_{H_2}}{p_{H_2O}} \right)
\]

For HTC2

\[
R = 10^{0.659 \pm 0.0125} \exp \left( \frac{-88 \pm 2.18}{RT} \right) \left( \frac{p_{CO}^{0.9 \pm 0.041}}{p_{CO}^{2.0 \pm 0.56}} \right) p_{CO}^{-0.31 \pm 0.056} p_{CO_2}^{-0.156 \pm 0.078} p_{H_2}^{-0.05 \pm 0.006} \left( 1 - \frac{1}{K} \frac{p_{CO} p_{H_2}}{p_{H_2O}} \right)
\]

It was determined HTC1 performed better than HTC2 on dry-feed coal derived syngas and slurry-feed coal derived syngas, while HTC2 performed better on the retentive side of a catalytic membrane reactor used in experimentation$^{112}$. We chose the HTC1 kinetic model for our UNSIM design because it

$^{109}$ Zhang, Oskan 2007
$^{110}$ Callaghan, 2006
$^{111}$ Hla, Park, Duffy, et.al 2009
$^{112}$ Hla, Park, Duffy, et.al 2009
performed better on slurry-fed coal derived syngas (which is what we are using for our model) and because a rate law for HTC1 was provided.

The rate law presented for HTC1 is not in a format acceptable to UNISIM. The field requirements include a reversible reaction rate, which needed to be derived from the rate law presented above. It was possible to determine the reverse reaction rate frequency factor and activation energy given the equilibrium constant as a function of temperature.

In order to plot $k_r$ as a function of inverse temperature, $k_f$ was first calculated over a range of temperatures using kinetic data from the rate law of HTC1. Next, the equilibrium constant, $K$, was found as a function of temperature (from a plot shown in Perry’s Chemical Engineering Handbook). Knowing $k_f$ and $K$ as a function of temperature, it was possible to plot $k_r$ as a function of temperature since $k_r$ and $k_f$ are related to $K$. Once $k_r$ was plotted it was possible to get the frequency factor and activation energy of the reverse reaction.

Next, it was necessary to find a catalyst with kinetic data to model the LTS reactor in UNISIM. One copper based catalyst used for low temperature shift is 50 wt% CuO, 45 wt% ZnO, and 5 wt% Al$_2$O$_3$, with a density of approximately 5168 kg/m$^3$. The water gas shift rate law for the copper based catalyst is as follows:

$$R = 4.785 \pm 2.187 \exp \left( \frac{-34.983 \pm 0.014}{RT} \right) \frac{P_{CO}^{0.85} P_{H_2O}^{0.2} P_{H_2}^{-1.9} P_{H_2}^{-0.57}}{1 - \frac{1}{K \frac{P_{CO}}{P_{CO} P_{H_2O}}}}$$

Like the HTS reaction, this rate law model is not an acceptable format to enter into UNISIM. To determine the reverse reaction rate law, the frequency factor and the activation energy for the reverse reaction using the equilibrium constant as a function of temperature.

In order to plot $k_r$ as a function of inverse temperature, $k_f$ was first calculated over a range of temperatures using kinetic data from the rate law of the LTS catalyst. Next, the equilibrium constant, $K$, was found as a function of temperature (from a plot shown in Perry’s Chemical Engineering Handbook). Knowing $k_f$ and $K$ as a function of temperature, it was possible to plot $k_r$ as a function of temperature since $k_r$ and $k_f$ are related to $K$. Once $k_r$ was plotted it was possible to get the frequency factor and activation energy of the reverse reaction. This method was the same used to find the kinetics of the reverse reaction for the HTS reactor.

### 3.3 Reactor Modeling in UNISIM

In order to determine important design details in the water gas shift unit, as well as optimize and test reactor performance, it was necessary to use a software program because of the complex calculations involved in determining design parameters. Thus one of the criteria for choosing catalysts for the WGS unit was the availability of kinetic data that could be adapted to software that could calculate reactor volume, conversion, cooling requirements, amount of catalyst necessary, and other important variables necessary for design. Several software programs are available that could solve the differential equations necessary to find some of the design parameters, but UNISIM was chosen for its speed of calculation, allowable iterations for optimization, and familiarity of use. UNISIM is an industry standard software for design and modeling of chemical and power plants, and was an ideal tool to conduct design and
modeling for our clean coal project. For the water gas shift specifically, it was possible to enter kinetic
data into UNISIM for each of the catalysts and model the chemical reactions taking place in both the HTS
and LTS reactors in as packed bed reactors. Once the two catalyst reaction sets for the HTS and LTS
reactors were added to UNISIM, a model of the two reactors with interstage cooling was created.
Figure 30: PFD for the water gas shift unit with control.
3.4 Detailed Process Description (WGS)

3.4.1 Compressor K-100W

Raw syngas with H₂S removed coming from absorber T-100S is mixed with high pressure steam from heat exchanger E-101S and sent to compressor K-100W. When entering the compressor, the syngas steam mix is at approximately 280 °C (536 °F) and 2758 kPa (400 psia). Using approximately 89 MW of parasitic load, the compressor brings the pressure of the syngas/steam mixture up to approximately 507 °C (945 °F) and 8274 kPa (1200 psia) before entering the HTS reactor.

3.4.2 HTS Reactor R-100W

The syngas enters the HTS reactor at 507 °C (945 °F) and 8274 kPa (1200 psia). The reactor is a packed bed reactor (PBR) that is filled with HTC1 catalyst with a void fraction of 0.4. This void fraction was chosen based on typical literature values for high temperature shift reactors\(^\text{113}\). For modeling purposes, the catalysts (though pellets in reality) were modeled as spheres with a diameter of 6 mm (0.02 ft), having a sphericity of one and a catalyst density of 5193 kg/m\(^3\) (324.2 lb/ft\(^3\)). The reactor volume was set to 300 m\(^3\) (10,594.4 ft\(^3\)), and the length was set to 20 m (65.617 ft). Using this volume and length, UNISIM calculated a reactor diameter of 4.3702 m (14.34 ft) and a void volume of 120 m\(^3\) (4238 ft\(^3\)). The reactor volume and length were set in order to achieve approximately 80% of equilibrium conversion in the HTS reactor. The conversion profile of CO in the HTS reactor is shown in Figure 31.

![CO Mole Fraction Profile of HTS Reactor](image)

**Figure 31:** CO mole fraction as a function of reactor length in the HTS reactor.

\(^{113}\) Haldor Topsoe
UNISIM also determines the temperature profile of the reactor as the water gas shift reaction progresses down the length of the reactor. Since the reaction is exothermic, we expected that the temperature would rise as the reaction progressed down the length of the reactor. The model for the HTS reactor behaved as expected, and the temperature profile shows a significant increase in temperature:

![Temperature Profile of HTS Reactor](image)

Figure 32: Temperature as a function of reactor length in the HTS reactor.

As the syngas travels through the packed bed HTS reactor the pressure of the stream decreases. UNISIM calculates the pressure drop across the packed bed reactor by using the Ergun Equation:

\[
\frac{\Delta P}{L} = 150 \frac{\mu G}{kg \rho D^2} \frac{(1 - \varepsilon)^2}{\varepsilon^3} + 1.75 \frac{G^2}{kg \rho D} \frac{(1 - \varepsilon)}{\varepsilon^3}
\]

Where \( \Delta P \) is the pressure drop, \( L \) is the length, \( G \) is the mass velocity, \( V \) is the superficial linear velocity, \( \rho \) is the fluid density, \( \mu \) is the fluid viscosity, \( D \) is the effective particle diameter, \( \varepsilon \) is the particle void fraction, \( g \) is the gravitational constant, and \( k \) is a conversion factor. According to UNISIM’s calculation, the pressure drop across the reactor is 49.23 kPa (7.14 psi).

### 3.4.3 Interstage Cooler E-100W

Following the HTS reactor the syngas enters an interstage cooler at approximately 646 °C (1195 °F) and 8224 kPa (1193 psia), and exits the heat exchanger at 400 °C (752 °F) and 8224 kPa (1193 psia). The temperature drop in the syngas is 426 °C (799 °F). UNISIM was able to calculate a heat duty, \( Q \), of 109.6 million watts, which corresponds to an overall heat transfer coefficient of 668 W/m²·K and a heat
transfer area of 403.4 m² given the heat transfer fluids, materials of construction, and other design parameters. In order to achieve this cooling, process cooling water at 600 kPa (87 psia) and 30 °C (86 °F) was used, and the flow rate required is 147,000 kg/h (324,000 lb/hr). The outlet of the process cooling water is 200 °C (392 °F) and 791 kPa (115 psia). This hot process water is then used to preheat syngas entering the fuel gas turbine K-100FG.

3.4.4 LTS Reactor R-101W

After exiting the interstage cooler E-100W, the syngas enters the LTS reactor R-101W at 400 °C (752 °F) and 8224 kPa (1193 psia). This reactor is a packed bed reactor that is filled with CuO/ZnO/Al₂O₃ catalyst with a void fraction of 0.4 and a density of 5168 kg/m³ (322.6 lb/ft³).

For this reactor the volume was set to 1600 m³ (56503.5 ft³) and the length was set to 60 m (196.85 ft). UNISIM calculated the diameter to be 5.83 m (19.12 ft) and the void volume to be 640 m³ (22600 ft³). The volume and length were set in order to achieve the maximum amount of shift to attain our target 90% carbon sequestration. The conversion profile of the LTS reactor is shown below:

![CO Mole Fraction Profile of LTS Reactor](image)

And again, we expected to see an increase in temperature over the length of the reactor because the reaction is exothermic. Because the mole fraction of CO in the feed to the LTS reactor is lower than that of the HTS reactor, we expected less temperature increase along the length of the reactor:
Finally, the Ergun Equation was used to calculate the pressure drop across the LTS reactor. According to UNISIM’s calculation, the pressure drop is 41.47 kPa (6.015 psi) across the reactor. For both the HTS and LTS reactors, the feed temperature and reactor pressure were set to the typical ranges for temperature and pressure found in literature.

### 3.4.5 Heat Exchanger E-101W

After leaving the LTS reactor, the shifted syngas enters heat exchanger E-101W at 511 °C (952 °F) and 8183 kPa (1187 psia) and exits at 35 °C (95 °F). The temperature drop across the heat exchanger is 476 °C (889 °F). UNISIM was able to calculate a heat duty, $Q$, of 305.8 million watts, which corresponds to an overall heat transfer coefficient of 933 W/m$^2$K and a heat transfer area of 6646 m$^2$ given the heat transfer fluids, materials of construction, and other design parameters. In order to achieve this cooling, cooling water at 600 kPa (87 psia) and 30 °C (86 °F) was used, and the flow rate required is 21,400 kg/h (47,200 lb/hr). The outlet of the process cooling water is 285 °C (545 °F) and 600 kPa (87 psia). This steam is sent to the HRSG and blended with the steam being sent to the steam turbine K-110ST.

### 3.4.6 Shifted Syngas Separator V-100W

After leaving heat exchanger E-101W, the shifted syngas is sent to V-100W. Here, the liquid water that was unreacted in R-100W and R-101W is separated from the syngas. The pressure of the vessel is 8201 kPa (1189 psia), and the temperature of the vessel is approximately 35 °C (95 °F). From the top of V-100W, the syngas leaves as a gas at approximately 626,000 kg/h (1,380,000 lb/h) while the water leaves
from the bottom of the vessel as a liquid at approximately 163,000 kg/h (359,000 lb/h) and sent to water treatment. The syngas which has now had the water removed is sent to the absorber T-100C to begin the process of CO₂ removal.

3.5 Equipment Sizing

Though the reactors are the most important pieces of equipment in the water gas shift unit, there are other pieces of equipment that needed to be designed and have their capital costs estimated. These pieces of equipment include a single stage compressor, a separation vessel, and two heat exchangers.

For the water gas shift unit, raw syngas from the desulfurization unit is mixed with hot steam from the convective syngas cooler on the outlet of the gasifier. This mixture of syngas and steam is compressed in a single stage compressor (K-100W), and flows into the HTS water gas shift reactor (R-100W). There, CO and water are reacted over the iron based catalyst to make more CO₂ and H₂. The syngas exiting the HTS reactor is very hot, and is cooled by an interstage cooler (E-100W) in order to achieve higher overall conversion. After leaving the interstage cooler, the syngas flows into the LTS reactor (R-101W), where it is shifted further by the copper based catalyst. Shifted syngas leaves the LTS reactor approximately 2.4 mol% CO in composition, achieving an overall conversion of approximately 92% between the HTS and LTS reactors. Finally, the shifted syngas is cooled to 35 °C in a shell and tube heat exchanger (E-101W), and the water is dropped out of the flow in the separator tank (V-100W) and sent to treatment. The shifted syngas is then sent to the CO₂ sequestration unit, where CO₂ is absorbed from the process stream using Selexol. The remaining CO continues through the process, and is combusted in the fuel gas turbine to CO₂. This unshifted CO is what accounts for the 10% CO₂ that is not removed and sequestered by the CO₂ removal unit.

After deciding on the equipment required in the water gas shift unit, it was necessary to determine their size, weight, area, and other design specifications as well as cost. Often UNISIM provided calculated design parameters such as height and diameter. Other methods were necessary to determine the weight and wall thickness of equipment, especially for the reactors. Using the method of Mulet, Corripio, and Evans (1981), the capital cost of the water gas shift reactors were estimated based on the weight of the shell and two 2:1 elliptical heads. The following Mathcad calculations serve as an example to show how the HTS reactor (R-100W) was cost estimated in Mathcad using the above method with added cost of catalyst:

114 Seider, Seader, Lewin, & Widagdo, 2009
Costing Calculations for the Water-Gas Shift Reactors

Capital Costs:

Weight of a pressure vessel, assuming constant shell and head thickness with 2:1 elliptical heads,

\[ \rho := 521.9 \quad \text{density of Inconel 617 (lb/ft}^3) \]

\[ D_i := 14.34 \quad \text{diameter if high temperature reactor (ft)} \]

\[ L := 65.617 \quad \text{length of high temperature reactor (ft)} \]

\[ P_o := 1200 - 14.7 \quad \text{operating pressure of high temperature reactor (psig)} \]

\[ T := 646 \quad \text{operating temperature of high temperature reactor (Celsius)} \]

\[ S := \begin{cases} 15000 & \text{if } T \leq 750 \\ 14750 & \text{if } 750 < T \leq 800 \\ 14200 & \text{if } 800 < T \leq 850 \\ 13100 & \text{if } 850 < T \leq 900 \end{cases} \]

\[ E := 1 \]

\[ P_d := \begin{cases} 1.1P_o & \text{if } P_o > 1000 \\ \exp\left[0.60608 + 0.91615\cdot\ln(P_o) + 0.001565\cdot(\ln(P_o))^2\right] & \text{otherwise} \end{cases} \]

\[ t_p := \frac{(P_d\cdot D_i)}{(2\cdot S\cdot E - 1.2P_d)} \]

\[ t_p = 0.658 \quad \text{increase thickness to 11/16\" due to availability} \]

\[ t_s := \frac{11}{16} \quad \text{wall thickness in inches} \]

\[ W_r := \pi\cdot(t_s + D_i)\cdot(L + 0.8\cdot D_i)\cdot t_s\cdot \rho \]

\[ W = 1305837.15 \quad \text{weight of vessel in lbs} \]
Because the environment is corrosive and includes hydrogen, an acceptable construction material is Inconel 617

\[ Fm := 3.9 \quad \text{See Seader Table 22.26 p. 576} \]

For horizontal vessels \(9,000 < W < 920,000\) lbs

\[ Cv := \exp\left[8.9552 - 0.2330 \cdot \ln(W) + 0.04333 \cdot (\ln(W))^2\right] \]

For horizontal vessels \(3 < D_i < 12\) ft

\[ Cpl := 2005 \cdot D_i^{0.20294} \]

\[ Cp := Fm \cdot Cv + Cpl \]

\[ Cp = 6128372.74 \quad \text{f.o.b. purchase cost at a CE index = 500} \]

assuming an CE cost index of 574 for the year 2015:

\[ Cost := \frac{Cp \cdot \left(\frac{574}{500}\right)}{574} \]

\[ Cost = 7035371.9 \quad \text{Capital Cost for high temperature reactor} \]

Catalyst Costs

\[ rVol := 300 \quad \text{m}^3 \]

\[ \text{voidfrac} := 0.7 \]

\[ \text{catdens} := 5193 \quad \frac{\text{kg}}{\text{m}^3} \]

\[ \text{catweight} := rVol \cdot (1 - \text{voidfrac}) \cdot \text{catdens} \]

\[ \text{catweight} = 4.674 \times 10^5 \]

\[ \text{catcost} := 17.31 \cdot \left(\frac{574}{402}\right) \cdot \text{catweight} \]

\[ \text{catcost} = 1.155 \times 10^7 \]

Total cost := Cost + catcost

Total cost = 18587014.38
Similarly, Mathcad was used to do design calculations on the interstage cooler (E-100W) between the two reactors:

**Shell and Tube Heat Exchanger Design**

- **Shell Diameter**: 37 inches (0.9398 m)
- **Tube Diameter**: 3/4 inches (0.019 m)
- **Triangle Pitch**: 1 inch (0.0254 m)
- **Number of Tubes**: 1074
- **Heat Exchanger Duty (W)**: $Q = 1.096 \times 10^6$

**Tube Side Properties**

- **Tube Inside Diameter (m)**: $D_i = 0.0135$
- **Thermal Conductivity of Tube Side Fluid (W/m-K)**: $k = 0.1150$
- **Mass Heat Capacity of Tube Side Fluid (J/kg-K)**: $C_p = 2040$
- **Mass Flow Rate of 1 Tube (kg/s)**: $F = 219.1$

- **Cross Sectional Area for Flow (m²)**: $A = \pi \left( \frac{D_i}{2} \right)^2$
- **Fluid Mass Velocit (kg/m²·s)**: $\dot{m} = 2.645 \times 10^{-5}$
- **Viscosity (Pa-s)**: $\mu = 0.3$

- **Constant**: $n = 0.4$ if heating, 0.3 if cooling

- **Sieder-Tate Equation (18.13)**:

  $h_f = 0.023 \cdot \frac{F}{D_i} \left( \frac{D_i}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^{0.2}$

- **Convective Heat Transfer Coefficient (W/m²·K)**: $h_f = 7.637 \times 10^3$

**Shell Side Properties**

- **Tube Outside Diameter (m)**: $D_o = 0.0191$
- **Shell Diameter (m)**: $D_s = 0.9398$
- **Thermal Conductivity of Shell Side Fluid (W/m-K)**: $k_s = 0.126$
- **Mass Heat Capacity of Fluid (J/kg-K)**: $C_{Pf} = 3121$
- **Mass Flow Rate of Fluid (kg/s)**: $F_s = 40.89$

- **Cross Sectional Area for Flow (m²)**: $A_s = \pi \left( \frac{D_o}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_s}{2} \right)^2 \right]$

- **Fluid Mass Velocit (kg/m²·s)**: $\dot{m}_s = 4.943 \times 10^{-4}$

- **Viscosity (Pa-s)**: $\mu_s = 0.36$

- **Kern Correlation Constants**:

  - $C_s = 0.36$
  - $n_s = 0.55$

- **Kern Correlation (18.21)**:

  $h_o = C_s \cdot \frac{k}{D_o} \left( \frac{D_o}{\mu} \right)^{0.332} \left( \frac{C_{Pf} \mu}{k} \right)^{0.332}$

- **Convective Heat Transfer Coefficient (W/m²·K)**: $h_o = 1.024 \times 10^3$
Inlet and Outlet Temperatures

Hot Side
\[ T_{hi} := 646.0 \]
\[ T_{ho} := 400 \]

Cold Side
\[ T_{ci} := 30 \]
\[ T_{co} := 200 \]

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

Materials of Construction

Shell Side - Carbon Steel
Tube Side - Inconel 617

\[ k_s := 19.76 \quad \text{Thermal Conductivity of Tubes (W/m-K)} \]
\[ L_t := 6.096 \quad \text{Tube length (m)} \]

Overall Heat Transfer Coefficient and Ft Correction Factor

\[ A_1 := \pi \cdot D_1 \cdot L \quad A_o := \pi \cdot D_o \cdot L \quad A_m := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \]
\[ t_w := D_o - D_i \]

\[ A_1 = 0.259 \quad A_o = 0.366 \]

\[ U := \frac{1}{\frac{1}{h_o} + \frac{t_w \cdot A_o}{k \cdot A_m} + \frac{A_0}{h_o A_1}} \]

\[ U = 719.029 \quad \text{Heat Transfer Coefficient (W/m}^2\text{K)} \]

\[ R_s := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 1.447 \quad S = 0.276 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 486

\[ F_T = 1.0 \quad \text{Ft Correction Factor for 2-4 Shell Pass Heat Exchanger} \]

Heat Transfer Area Needed

\[ A = \frac{Q}{U \cdot F_T \cdot T_{lm}} \]

\[ A = 374.684 \quad \text{Heat Exchange Area Needed (m}^2\text{)} \]

\[ A_{ft} := A \cdot 10.76 = 4.032 \times 10^3 \quad \text{Heat Exchange Area Needed (ft}^2\text{)} \]
3.6 Materials of Construction

As shown in both the reactor and heat exchanger design equations, Inconel 617 was chosen as a construction material when dealing with high temperature and pressure syngas. This is because standard construction materials like carbon steel and stainless steel should not be used with hydrogen because of hydrogen embrittlement, and also experience metal dusting and corrosion when CO and CO\textsubscript{2} are present\textsuperscript{115}. Metal dusting is corrosion caused by carbon, and manifests itself as a break up of bulk metal to metal powder\textsuperscript{116}. Several Inconel and Monel alloys were analyzed to determine their response to exposure to syngas (in the presence of CO and 20\% H\textsubscript{2}) at 621 °C. All experienced pitting due to metal dusting, but Inconel 617 shows good resistance to pitting at long exposure times:

Inconel 693 also shows good resistance to pitting, though it is more expensive than Inconel 617. Thus, Inconel 617 was chosen for reactor linings and heat exchanger tube bundles that contain high temperature and pressure syngas. Unfortunately, our reactors operate hotter than the temperature at which the above data was taken. Therefore, the rate of corrosion in our water gas shift reactors will be higher than projected here since our operating temperatures are slightly higher (645 °C).

\textsuperscript{115} Baker & Smith
\textsuperscript{116} Chun, Mumford, Ramanarayanan
4 Air Separation Block

Most large scale commercial oxygen is produced using one of three well established processes: pressure swing adsorption, cryogenic distillation, and membrane separation.\textsuperscript{117} Each of these processes has unique energy and capital costs and provides oxygen at a wide variety of purities. The oxygen provided to the IGCC plant needs to be high purity to remove inert N\textsubscript{2} so it does not dilute the process streams in the plant but the energy costs associated with achieving high purity must be optimized to reduce the overall cost of energy produced by the plant. Based on the gasifier chosen for the IGCC, it is necessary to have an oxygen stream at least 95 mol\% pure to run the plant efficiently. The main goal of the air separation unit design is to provide an oxygen stream at high enough purity that the plant can run efficiently, but to do it using as little energy as possible to minimize the unit’s impact on the overall plant efficiency. As illustrated in Figure 36, the preferred method of oxygen generation for a specific process depends on both the amount of oxygen needed to run the process and the required purity of the oxygen stream.

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{figure36.png}
\caption{Economic expediency of oxygen supply and manufacture techniques\textsuperscript{118}}
\end{figure}

\textsuperscript{117} Hansel, 2005
\textsuperscript{118} GRASYS JSC, 2011
4.1 Air Separation Process Options

4.1.1 Pressure Swing Adsorption

Noncryogenic air separation processes are often used for small to moderate oxygen production units. The Pressure Swing Adsorption (PSA) cycle is a process where adsorption into a packed bed takes place at a much higher pressure than desorption. It is generally used for bulk separations involving high concentrations of contaminates. PSA plants can be designed to a wide range of sizes ranging from small hospital units to large units producing as much as 165 tons per day per vessel.\textsuperscript{119}

The PSA cycle, shown in Figure 37, has three steps: adsorption, blowdown, and repressurization. After the vessel is pressurized with air, the nitrogen is adsorbed into the bed of molecular sieves. This leaves the remaining components of the gas, which is mostly oxygen but can also contain CO\textsubscript{2} and argon, concentrated. The depressurization step removes the concentrated oxygen stream and reduces the pressure in the adsorbent bed which then causes the weakly adsorbed nitrogen to desorb. This depressurization step can be adjusted to provide the desired purity of the strongly adsorbed component. Next, the blowdown step removes the N\textsubscript{2} at the adsorption inlet of the bed by reducing the pressure of the bed. If the blowdown is extended to a vacuum (vacuum swing adsorption), the yield of the cycle is greatly increased. The bed is then repressurized to the feed pressure and the process is repeated.\textsuperscript{120}

![Figure 37: Process flow diagram of an adsorption oxygen generator\textsuperscript{121}](image)

\textsuperscript{119} Hansel, 2005
\textsuperscript{120} Yon & Sherman, 2003
\textsuperscript{121} GRASYS JSC, 2011
For the bulk separation of O\textsubscript{2} and N\textsubscript{2}, zeolites, clinoptilolite and mordenite, and carbon molecular sieves are commonly used as the adsorbent.\textsuperscript{122} Commercial oxygen produced using the PSA cycle generally has 90 – 93% purity.\textsuperscript{123} The purity is limited because the argon present in the air remains with the O\textsubscript{2} as it passes through the bed.\textsuperscript{124} Residual water and CO\textsubscript{2} may also be present in the oxygen stream in the low ppm range.\textsuperscript{125}

Commonly, the process involves two or three adsorbers alternating between the adsorption and desorption steps to provide a more constant stream of oxygen.\textsuperscript{126} Figure 37 illustrates the process using two adsorbers. Because this is a batch process there is not a constant flow of oxygen from the system. The unsteady flow can be mitigated by using a surge tank to even out the gas flow.\textsuperscript{127} Based on the specifications from existing PSA units, a 35 m\textsuperscript{3} unit is capable of producing up to 2.1 tons of 90% O\textsubscript{2} per day.\textsuperscript{128} The space requirement for the vessels is estimated based on existing occupation area data to be 26 m\textsuperscript{3} per ton of O\textsubscript{2} per day.\textsuperscript{129}

Pressure swing adsorption is a well-established technology capable of producing an oxygen stream of acceptable purity. However, it is most economically viable when applied to processes with low to moderate oxygen requirements. A coal power station requires thousands of tons of oxygen per day and the size and energy requirements for the number of vessels needed to provide that tonnage makes PSA a less attractive candidate for oxygen generation.

4.1.2 Cryogenic Distillation

Cryogenic technology is often used to produce high purity gasses at a low cost.\textsuperscript{130} Cryogenic distillation is a technique commonly used to separate air into is separate components and is capable of producing commercial oxygen up to 99.5% purity. The balance impurity is mostly argon due to the closeness of boiling points.\textsuperscript{131}

During the cryogenic distillation process, shown in Figure 38, the ambient air is compressed and the water and carbon dioxide is removed by adsorption, generally by passing the gas through a bed of zeolite or silica gel type adsorbents. After the impurities have been removed the air is cooled using heat exchangers and is expanded to low pressure through a turbine or a valve.\textsuperscript{132} The resulting liquid, formed at 80 K, is then distilled using a two stage separation process to separate the oxygen and nitrogen. The first stage consists of a high pressure column (500-600kPa) to separate the feed air stream into nitrogen

\textsuperscript{122} Yon & Sherman, 2003
\textsuperscript{123} Hansel, 2005
\textsuperscript{124} Yon & Sherman, 2003
\textsuperscript{125} Hansel, 2005
\textsuperscript{126} Yon & Sherman, 2003
\textsuperscript{127} Hansel, 2005
\textsuperscript{128} IGS Italia, 2009
\textsuperscript{129} ZhuHai GongTong Mechanical Equipment Co., Ltd.
\textsuperscript{130} Agrawal, Herron, Rowles, & Kinard, 2001
\textsuperscript{131} Hansel, 2005
\textsuperscript{132} Hansel, 2005
Cryogenic distillation is the current preferred method to obtain very high purity oxygen streams in large quantities. However, while this is a mature, reliable method of oxygen generation there is significant cost associated with its use due to the extensive equipment list associated with the distillation process. Additionally, cryogenic distillation is a very energy intensive process which and the parasitic energy effect on the plant would decrease net efficiency and increase the cost of electricity produced by the plant. There is a possibility of saving energy on distillation not purifying the to the maximum possible level but instead only purifying to the minimum required level to run the gasifier. To aid in deciding which process to choose, a simple distillation model was constructed using UNISIM to determine the energy difference between processes producing the two different purity levels. The simulation showed that there is approximately a 20% energy savings when the stream is only brought to a target of 90% purity instead of 99% purity. This demonstrates that there is a possibility of reducing the energy

Agrawal, Herron, Rowles, & Kinard, 2001
Agrawal, Herron, Rowles, & Kinard, 2001
Air Products and Chemicals, Inc., 2011
consumption of the distillation process by reducing oxygen purity and it was further investigated through the development and optimization of rigorous distillation model in UNISIM.

4.1.3 Gas Membrane Separation

Membrane separation is a relatively new, emerging technology when it comes of oxygen separation. The production of high purity nitrogen using membrane technology is well established but membrane technology is in the process of evolving to a point where it can be used produce a high purity oxygen stream.

4.1.3.1 Hollow Fiber Membranes

Traditionally, hollow fiber membranes have been used for gas separation to produce pure nitrogen and oxygen enriched air. This membrane technology is capable of removing some nitrogen from a pressurized air stream but is only capable of enriching air up to 45% oxygen content.\textsuperscript{136} In gas separation one of the components of the pressurized feed permeates through a selective membrane much faster than the others (Figure 39). The driving force is a pressure difference between feed gas and the lower pressure permeate.\textsuperscript{137}

\begin{figure}
\centering
\includegraphics[width=\textwidth]{membrane_cartridge.png}
\caption{Distribution of gas flows in a membrane cartridge\textsuperscript{138}}
\end{figure}

\textsuperscript{136} GRASYS JSC, 2011
\textsuperscript{137} Baker, 2005
\textsuperscript{138} GRASYS JSC, 2011
This traditional method of membrane gas separation is capable of producing large quantities of gas while using much less energy compared to cryogenic distillation. The drawback of this method is the limits on oxygen purity. The process was originally developed to produce high purity commercial nitrogen because oxygen and the other components of air permeate through the membrane much more quickly than nitrogen leaving an almost pure stream behind. However, the oxygen rich permeate still contains all the argon, water, and carbon dioxide, as well as still significant quantities of nitrogen that permeated the membrane wall alongside it. With a maximum purity of 45% oxygen right stream, this limit does not meet the goals of this project to provide a >90% oxygen stream to the power plant to prevent the formation of harmful side products in the gasifier or boiler.

4.1.3.2 Ion Transport Membrane

A new membrane separation technology is in the process of being developed by Air Products and Chemicals, USA to provide high purity commercial oxygen in large quantities for future industrial and energy markets.\(^{139}\)

The ceramic Ion Transport Membrane (ITM) is extremely selective and accommodates very fast transport of oxygen. The selectivity is due to the ceramic material composed of a metallic oxide structure incorporates oxygen ion vacancies in its structure when heated greater than 700°C. Compressed air at 800 – 900°C is passed over the membrane wafers and oxygen ions pass through the membrane to form O\(_2\) gas on the permeate side (Figure 40 and Figure 41).\(^{140}\)

![Figure 40: Mechanism of oxygen ion transport across membrane](image)

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\(^{139}\) Air Products and Chemicals, Inc., 2011

\(^{140}\) Fogash, 2007

\(^{141}\) Fogash, 2007
This single stage separation method makes for a compact design and because it is a high temperature process it has more synergy with an energy generation system than a cryogenic process. Air Products and Chemicals, the company designing the ceramic membrane technology, estimates that this new design will take up 50% less plot area and the fewer materials of construction has an estimated capital cost savings of 35% over a cryogenic distillation unit with equivalent capacity. The power required to produce the oxygen is estimated to be 37% less for an IGCC. The potential savings using an ITM rather than cryogenic distillation are substantial but the technology is still being developed and is currently in the pilot plant phase. A five ton per day (TPD) prototype plant was successfully tested demonstrating >99% purity. A 100 TPD intermediate scale test unit is currently under construction and will test the commercial design concept to provide data for larger scale plants. Air Chemicals plans to bring a 2000 TPD on stream by 2015. This is the order of magnitude required to provide sufficient oxygen to a power plant.

4.1.4 Choice of Air Separation Unit

To evaluate the different air separation processes and decide on the most appropriate design to supply the oxygen to the coal plant, we developed a decision matrix taking into consideration all the important factors that would affect the overall design and operation of the plant. Each factor was scored on a scale

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142 Air Products and Chemicals, Inc., 2011
143 Fogash, 2007
144 Air Products and Chemicals, Inc., 2011
of 1 to 10 and then weighted as a percentage of the total score. The process with the overall highest score is the one that best fits the requirements of the project.

The variables chosen as being the most important for determining the best air separation unit to meet our requirements are the capital cost, capacity, operating cost, extent developed, and the purity of the oxygen stream. The capital cost includes all costs incurred on land, materials, and construction of the air separation unit. The capital cost of this unit is a significant portion of the overall plant cost and was given a weight of 15% in the decision matrix. The capacity is the process’s ability to produce the volume of oxygen needed to run the gasification process at optimized conditions. This is a very important design consideration because if a process is not capable of providing enough oxygen it is not adequate for the plant. Because this is such an important consideration, it was given a weight of 25% in the design matrix. The operating cost is also a very important design consideration. The biggest cost of producing oxygen is the energy required to run the air separation unit. Any energy required to produce oxygen will be taken directly from the plant so an increase in energy consumption by the air separation unit will reduce the overall plant efficiency and the amount of electricity available to customers. Because of this the operating cost is considered to be one of the most important design variables and was given a weight of 25%. The extent of the technology’s development is an important decision variable because there is an inherent risk when a design incorporates a new, less developed technology. The project has a risk of failing if new technology is used before it is fully understood and unforeseen design problems prevent the process from operating correctly. This consideration was given a weight of 15%. The final design variable considered was the purity of the oxygen stream produced by the air separation unit. A purer oxygen stream means few inert compounds, such as Ar and N₂, and fewer contaminates, such as CO₂, are introduced to the system and the gasifier will operate with a higher efficiency. This variable was given a weight of 20% because of its importance to the operation of the gasifier.

<table>
<thead>
<tr>
<th>Gas Separation Method</th>
<th>Capital Cost</th>
<th>Capacity</th>
<th>Operating Cost</th>
<th>Extent Developed</th>
<th>Oxygen Purity</th>
<th>Total Score</th>
</tr>
</thead>
<tbody>
<tr>
<td>Weight</td>
<td>15%</td>
<td>25%</td>
<td>25%</td>
<td>15%</td>
<td>20%</td>
<td>100%</td>
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<td>6</td>
<td>10</td>
<td>7</td>
<td>5.9</td>
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<td>4</td>
<td>10</td>
<td>10</td>
<td>7.2</td>
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<td>6</td>
<td>1</td>
<td>6.6</td>
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<tr>
<td>Ion Transport Membrane</td>
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<td>4</td>
<td>7</td>
<td>4</td>
<td>10</td>
<td>6.4</td>
</tr>
</tbody>
</table>

**Capitol Cost (15%)**

Pressure swing adsorption is considered to have a lower capital cost because it only requires several large vessels and compressors to run the process which are relatively simple and inexpensive equipment. However, the number of vessels required does add up to a significant expense to it is given a rating of 6 in the decision matrix. Cryogenic distillation requires several different large vessels such as distillation towers and heat exchangers and also requires several pumps and compressors to pressurize the vessels and move the fluids. Because of all the equipment required for this process the capital cost is relatively large, preliminarily estimated to be $390 million, and this process was scored as 1 on the
decision matrix. Both membrane technologies require a relatively simple vessel and compressors to pressurize the vessel for separation to take place. Because of the simplicity of the equipment, the hollow fiber membrane received a score of 10 in the decision matrix. The ITM additionally needs a heat exchanger to heat the air to the required temperature and the ceramic membrane, because it is new technology, is more expensive than a traditional hollow fiber membrane. The preliminary cost estimate for an ITM plant was $250 million and scored an 8 on the decision matrix.

**Capacity (25%)**

As stated before, the number of vessels required for PSA to provide a large enough oxygen stream to run the power plant is not feasible. This is generally a process used for small and medium scale production of high purity oxygen. Therefore, PSA scored a 1 on the decision matrix for capacity. Cryogenic distillation has been proven to supply the required amount of oxygen for existing IGCC plants so it scored a 10. Hollow fiber membranes are capable of providing a large volume of gas from a relatively small vessel but by volume, that gas is less than 50% oxygen, so it scored a 6 in the matrix. The ion transport membrane has the potential to supply sufficient quantities for the plant with further development but it has not yet been proven with a full scale design the way distillation has so it received a score of 4 in the decision matrix.

**Operating Cost (25%)**

The operating cost of pressure swing adsorption is relatively small because the only equipment requiring energy is the compressors and vacuums on each vessel. However, the number of compressors and vacuums required would be significant so the process was given a score of 6 for operating cost. Cryogenic distillation is very energy intensive because of the energy requirements of the tower and the refrigeration processed so this process scored the lowest with a 4 on the matrix. The hollow fiber membrane only requires a compressor to pressurize the air into the separation vessel. This process requires relatively little energy so it was scored as a 10 on the decision matrix. The ion transport membrane also requires relatively little power compared to distillation but the energy needed to heat the air must be taken into account in addition to the energy needed to compress it. Therefore, it was given a score of 7 on the decision matrix.

**Extent of Technology Development (15%)**

Pressure swing adsorption and cryogenic distillation are both a well-established, reliable methods of oxygen generation. Therefore, both of these processes were given a score of 10 on the decision matrix. The hollow fiber membrane has been established and is used commercially but the technology is still being developed to try and make membranes more selective for oxygen and improve stream purity. With that consideration, this membrane was scored as a 6 on the matrix. The ion transport membrane is still in the small scale prototype phase. The concept has been proven but it has not been implemented on a scale large enough to suit our power plant’s needs so it was given a score of 4 for extent developed.

**Purity of Oxygen Stream (20%)**

The purity of oxygen provided by pressure swing adsorption is between 90 and 93%. This hits the project’s target of a 90% pure oxygen stream but is not as good as the best purity that can be achieved
using some of the other processes. Therefore, this process was scored as a 7 on the decision matrix. Cryogenic distillation and the ion transport membrane are both capable of providing oxygen streams well over 99% purity so these processes scored a 10 on the matrix. Hollow fiber membrane technology does not provide a pure stream, but only an oxygen enriched stream up to 45% oxygen. This does not meet the project’s target and was scored as a 1 on the matrix.

A final score for each process was then calculated by taking the score in each column, multiplying it by its weight percentage, and adding all the resulting values together. The final scores are shown in the right hand column on the decision matrix. Based on the variables considered and the scores assigned to each process, cryogenic distillation is the best option for providing the oxygen stream to the power plant.

4.2 Cryogenic Distillation Unit Design

Once we had evaluated each oxygen generation method and determined the most appropriate method to use for the IGCC plant, we began designing the entire plant. The initial design step was to create a UNISIM model of the cryogenic distillation plant. The major components needed for a cryogenic distillation process were determined primarily using the Kirk-Othmer Encyclopedia and from Linde’s ‘Cryogenic Air Separation – History and Technological Progress.’

4.2.1 Process Overview

The process of cryogenic distillation is relatively simple in concept because fundamentally it is exactly the same as normal distillation of two fluids. The only difference is that cryogenic distillation occurs at extremely low temperatures because the normal boiling point of oxygen is -297°F and the boiling point of nitrogen is -320°F.

The first step of cryogenic distillation is compression. Ambient air is compressed to 6.5 bar using a two stage centrifugal compressor with interstage cooling. Once the air has been compressed to the required pressure and has been cooled back down to ambient temperatures, it is cleaned of CO₂ and water vapor by sending it through a pre-purification unit containing activated alumina and molecular sieves and then it enters the main heat exchanger. This heat exchanger is a brazed aluminum plate-fin heat exchanger that uses the cold oxygen and nitrogen product streams to cool the inlet air stream to cryogenic temperatures. A small stream (about 12% of the inlet air flow) is drawn out of the heat exchanger and sent to a gas expander. This stream is expanded to 1.5 bar to provide additional cooling and is then sent to the low pressure column. The high pressure stream is fed into the bottom of a high pressure column operating at 5.75 bar. The refluxed distillate of this column is nearly pure nitrogen and the bottoms product is a crude, enriched oxygen product. Both of these streams are then fed through a subcooler, which is another brazed aluminum plate-fin heat exchanger, for additional cooling using the outlet streams of the low pressure column as a cooling fluid. Both of these streams are then throttled to 1.5

145 Agrawal, Herron, Rowles, & Kinard, 2001
146 The Linde Group
bar and join the expanded air stream in the low pressure column. The low pressure column does a secondary distillation to provide the final nitrogen and oxygen products. The liquid outlet oxygen product is then pumped to 1000 psi because this is the pressure the oxygen must be delivered to the gasifier. Two product streams are then run through the subcooler and then the main heat exchanger to provide cooling for the inlet stream. The oxygen is then sent to the gasifier and the nitrogen passed through the pre-purification unit to regenerate the alumina and molecular sieves and is then released to the environment.

### 4.2.2 Oxygen Purity Specification

The major design decision for the design of the ASU was to determine the most economical oxygen purity of the stream being fed to the gasifier. As a general rule of thumb, the higher the oxygen purity, the more energy the process requires, but the stream must be at least 95 mol% pure to meet the operating requirement of the gasifier. Oxygen purity does not have a significant impact on capital cost unless the unit is going to be used to provide a pure argon stream as well. This plant will be used only for its oxygen generation capabilities so we decided that including an argon recovery column is unnecessary. After designing the cryogenic distillation process in UNISIM, the model was run specifying several different oxygen product purities to determine the effect of oxygen purity on energy consumption. For the purpose of this analysis, a production basis of 60 kg O$_2$/s was set. The results of this analysis are shown in Figure 42.

![System Energy Requirements](image)

*Figure 42: The power requirement of the distillation unit as a function of oxygen stream purity*

This plot shows that the energy consumption of the unit is relatively constant until the purity gets above 97 mol%. Past 98% purity, the energy consumption begins to increase exponentially. Based on this data, we made a value judgment to set our oxygen purity at 98 mol%. At this purity, the energy consumption is only 1.1% greater than the requirement to meet our minimum threshold of 95% and the increased 3% in oxygen purity will enable the gasifier to run more efficiently due to decreased inert concentration in the process stream.
Figure 43: Process flow diagram for the cryogenic distillation unit
4.2.2 Detailed Process Description

Figure 43 is a comprehensive flow sheet of the air separation process that gives a detailed view of the equipment involved in the distillation process.

4.2.2.1 Main Air Compressor – K-100A and K-101A

The purpose of the main air compressors (MAC) is to compress atmospheric air to 94.3 psi (6.5 bar) to be used for cryogenic distillation. The MAC is comprised of two compressor units K-100A and K-101A with interstage cooling because the overall compression ratio of 6.5 is too large for a single compressor to handle efficiently. Heuristics say that the maximum compression ratio for each stage should be no more than 4, necessitating the use of two compressors. It was then necessary to optimize the interstage pressure which corresponds to equal horsepower for each compressor. The optimal interstage pressure is calculated using Equation 39 where \( n \) is the number of stages and \( P_i \) is the outlet pressure of a particular stage.

\[
P_i = P_{i-1} \left( \frac{P_{\text{final}}}{P_{\text{initial}}} \right)^{1/n}
\]

Eq 39

Using this equation, we determined that the optimal interstage pressure is about 37.3 psi (2.53 bar). However this value needed to be adjusted upwards slightly to counteract the pressure drop across the heat exchanger. After initial compression in K-100A, the air stream leaves the compressor at 312.8°F and 38.4 psi. Heat Exchanger E-100A then uses cooling water to cool the air back down to 95.0°F. The stream then enters K-101A and is compressed to 94.3 psi (6.5 bar) and leaves at 322.3°F. The stream then enters E-101A where it is cooled by the oxygen product stream to 251.7°F. The stream then goes to E-102A for additional cooling against the nitrogen product stream to 95.0°F.

4.2.2.2 Air Pre-Purification Unit – V-100A and V-101A

After being cooled, the air inlet stream enters the pre-purification unit (PPU). This unit is comprised of two horizontal vessels, V-100A and V-101A, and the purpose of these vessels is to remove any CO\(_2\), water vapor, and trace hydrocarbons from the air stream. These trace impurities must be removed from the stream or they will freeze at the cryogenic temperatures in the coldbox. A buildup of impurities can damage equipment and cause dangerous bottlenecks in the system. The first section of each of these vessels is packed with activated alumina which is used to adsorb all the water vapor entering the unit. The second part of the vessel is packed with molecular sieves which are used to adsorb any CO\(_2\) and other impurities in the air stream. Two parallel vessels packed with adsorbent are needed because at any time, one vessel is online and cleaning the inlet stream and the other vessel is offline and being regenerated by running the plant’s nitrogen product stream through the vessel at a high temperature to desorb and remove all the impurities in the adsorbent. The vessels are cycled every three hours.

The vessel size was determined by calculating exactly how much alumina and sieves each vessel needed to effectively clean the inlet stream for 3 hours. F-1 alumina is capable of holding 0.07 lb of water per
pound and 13X zeolite molecular sieves are capable of holding $4 \times 10^3$ kmol CO$_2$ per kg.\textsuperscript{147} Assuming that CO$_2$ accounts for 0.03 vol% air and the air is 3 vol% water vapor, adsorbent must be capable of adsorbing 36.2 kmol CO$_2$ and 3456 kmol water during a 3 hour cycle. Based on these values, the total amount of each adsorbent and the required volume of the vessel were calculated. The results are shown in Table 22. The 3 hour cycle time was chosen to make the volume of the PPU vessels a reasonable size while still having the cycle long enough to sufficiently desorb the contaminants and regenerate the adsorbent.

Table 22: Adsorbent needed to remove impurities from ASU inlet stream and vessel volume

<table>
<thead>
<tr>
<th>Component</th>
<th>Carbon Dioxide</th>
<th>Water</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cyclic Load – 3 hours (mol)</td>
<td>36180</td>
<td>3456000</td>
</tr>
<tr>
<td>Adsorbent Mass (kg)</td>
<td>18090</td>
<td>3920</td>
</tr>
<tr>
<td>Adsorbent Density (kg/m$^3$)</td>
<td>680</td>
<td>830</td>
</tr>
<tr>
<td>Adsorbent Volume (m$^3$)</td>
<td>26.6</td>
<td>4.7</td>
</tr>
<tr>
<td>Total Vessel Volume</td>
<td>31.3 m$^3$</td>
<td>1105 ft$^3$</td>
</tr>
<tr>
<td>Vessel Diameter</td>
<td>2.0 m</td>
<td>6.6 ft</td>
</tr>
<tr>
<td>Vessel Height</td>
<td>10.0 m</td>
<td>32.8 ft</td>
</tr>
</tbody>
</table>

Both V-100A and V-101A are 1105 ft$^3$ and house enough adsorbent to hold the impurities of the inlet stream for a 3 hour cycle with a 33% safety factor in case of conditions of abnormally high concentrations of impurities in the inlet stream.

4.2.2.3 Main Heat Exchanger – E-103A

Once the stream has had all trace impurities remove it enters the coldbox section of the distillation plant via the main heat exchanger. This is a brazed aluminum plate fin heat exchanger that used the two product streams from the distillation column to cool the inlet stream down to cryogenic temperatures. Plate fin heat exchangers are commonly used in cryogenic distillation because they have up to 1500 m$^2$/m$^3$ of heat transfer area enable very close temperature approaches between heating and cooling streams while keeping the unit relatively small. A diagram of a plate fin heat exchanger is shown in Figure 39. There are several types of fins that are used in these exchangers, as shown in Figure 40. We decided that a serrated strip fin would be the best configuration to use for the exchangers because they generally have the highest heat transfer coefficients due the very turbulent flow they create. However, they do have the drawback of a higher pressure drop than other fin configurations.

\textsuperscript{147} CACHE Corporation
Figure 44: Diagram of a 4 stream plate-fin heat exchanger

(a)  (b)  (c)  (d)  (e)  (f)

Figure 45: Types of plate fin surfaces: (a) plain rectangular (b) plain trapezoidal (c) wavy (d) serrated or offset strip fin (f) perforated

148 Linde Engineering
149 Dewatwal, 2009
Plate fin heat exchangers are capable of incorporating several streams into one unit for heating and cooling and are utilized to cool the inlet air stream. In the main heat exchanger E-103A, the compressed air stream at 77.0°F first encounters the cooling nitrogen stream flowing counter currently and is cooled to -69.4°F. It is then contacts the countercurrent cooling oxygen stream and is cooled further to -171.6°F. After contacting the oxygen stream, 12% of the flow is withdrawn from the exchanger and sent to gas expander K-102A. The later expansion of this side stream in K-102A provides refrigeration necessary to operate the distillation unit. The remaining air stream is then contacted again with the nitrogen stream for further cooling to -260.3°F. A concept diagram of the flow inside the plate fin heat exchanger is shown in Figure 46.

![Concept diagram showing the flow of hot and cold streams in plate fin heat exchanger E-103A](image)

For each section of the heat exchanger where two specific streams are contacting, it was necessary to model the flow and heat exchange properties to determine a heat transfer coefficient and the area required for the needed heat exchange to take place. Several correlations have been developed for estimating flow through these exchangers and the correlation used for our design was the Magahanic correlation. Heat exchanger specifications for E-103A are shown in Table 23. Once the compressed air stream has been cooled, a portion of the flow is sent to gas expander K-102A and the remainder is sent to T-100A.

<table>
<thead>
<tr>
<th>Specifications for E-103A</th>
</tr>
</thead>
<tbody>
<tr>
<td>Height</td>
</tr>
<tr>
<td>Width</td>
</tr>
<tr>
<td>Length</td>
</tr>
<tr>
<td>Heat Exchange Area</td>
</tr>
<tr>
<td>Duty</td>
</tr>
</tbody>
</table>
4.2.2.4 Gas Expander – K-102A

The purpose of the gas expander is to take the side stream sent to it from E-103A and expand it to 21.76 psi (1.5 bar) to drop the temperature of the stream to provide needed refrigeration to run the plant. The stream enters K-102A at 85.4 psi and -171.6°F and leaves the expander at 21.76 psi and -246.5°F. This stream is then sent to the low pressure distillation column T-101A and is fed into the column at stage 2.

4.2.2.5 High Pressure Distillation Column – T-100A

The initial separation of the two components, nitrogen and oxygen, occurs in the high pressure distillation column. The main flow leaving E-103A is fed into the bottom stage of T-100A which is a packed, refluxed column. As the incoming gas rises through the column, gas condenses as it contacts the packing and liquid flowing down from above and falls into an oxygen enriched, liquid pool at the bottom of the column. The vapor that leaves the top of the column is condensed by contacting in a condenser with the bottoms product to T-101A. The major design component of the column distillation design was determining the appropriate number of theoretical equilibrium stages to achieve maximum separation of the components while keeping the column as small as possible. We used UNISIM to simulate tower with a wide range of stages and then plotted the resulting composition of the bottoms oxygen enriched stream and the reflux ratio required in the tower. The results of the analysis are shown in Figure 47.

![High Pressure Column Performance](image)

*Figure 47: Performance of T-100A as the number of theoretical stages in the tower is varied*

This analysis shows that the maximum oxygen content of the bottoms product is approximately 41 mol% oxygen and uses a reflux ratio of about 1.3. We decided that the optimal height of the tower would consist of 30 stages because this puts the column at near maximum separation and minimum reflux. Adding more stages would result in diminishing, almost nonexistent, increases in product purity. However, fewer stages would result in significantly decreased product quality. Therefore, we made a
judgment call and decided that 30 stages was the optimal number of stages for the high pressure column.

The next step in the design was to determine the type of packing in the column and to size the column to accommodate the needed liquid and vapor flow rates. The UNISIM tray sizing utility was used to size the column and the dimensions are shown in Table 24.

<table>
<thead>
<tr>
<th>High Pressure Column T-100A Specifications</th>
</tr>
</thead>
<tbody>
<tr>
<td>Internals</td>
</tr>
<tr>
<td>Column Diameter</td>
</tr>
<tr>
<td>Flooding</td>
</tr>
<tr>
<td>Cross Sectional Area</td>
</tr>
<tr>
<td>Column Height</td>
</tr>
<tr>
<td>Column Pressure Drop</td>
</tr>
<tr>
<td>HETP</td>
</tr>
<tr>
<td>Packing Type</td>
</tr>
</tbody>
</table>

The type of packing used in the column is a Norton Intalox packing that is used because of its large contact surface area between the vapor and liquid and the low pressure drop it causes across the column. An example of Intalox High performance packing is shown in Figure 48. The packing in the cryogenic tower is made of stainless steel to withstand the low temperature. The column walls are also made of stainless steel. The pressure drop of packed columns is significantly less than that of trayed columns and because of the large contacting area, the needed height to achieve an equilibrium stage is much smaller than with trays.

Figure 48: An example of the Norton Intalox packing in T-100A

---

150 Kock-Glitsch
The bottoms stream leaves T-100A as a crude product at 41.1 mol% oxygen at -279.1°F and the distillate stream leaves 99.5 mol% nitrogen at -287.2°F. Both of these streams are then sent to E-104A for further cooling.

4.2.2.6 Subcooler – E-104A

The subcooler is another brazed aluminum plate fin heat exchanger that is used to further cool the product streams leaving T-100A before they are throttled and sent into T-101A. The product overhead stream of nitrogen leaving T-101A is used to cool these two stream. A concept diagram showing the stream flows E-104A is shown in Figure 49.

![Figure 49: Concept diagram showing the flow of the hot and cold stream through E-104A](image)

The design for E-104A followed used the same correlation and design procedure as E-103A. The specifications for E-104A are shown in Table 25. After the streams exit the subcooler they are sent through throttling valves to drop the pressure to 21.76 psi (1.5 bar) which is the operating pressure of the low pressure column T-101A.

<table>
<thead>
<tr>
<th>Specifications for E-104A</th>
</tr>
</thead>
<tbody>
<tr>
<td>Height</td>
</tr>
<tr>
<td>Width</td>
</tr>
<tr>
<td>Length</td>
</tr>
<tr>
<td>Heat Exchange Area</td>
</tr>
<tr>
<td>Duty</td>
</tr>
</tbody>
</table>

4.2.2.7 Low Pressure Distillation Column – T-101A

Once the nitrogen and crude oxygen streams leaving T-100A have been throttled to 21.8 psi (1.5 bar), they are fed into T-101A, the low pressure distillation column, for separation into the final two product streams. The nitrogen stream from T-100A is fed into the top stage of T-101A, the stream from the gas expander K-103 A is fed into stage 2, and the enriched oxygen stream from T-100A is fed into T-101A at stage 12. T-101A is also a packed column using the same Norton Intalox packing as T-100A. The inlets enter the column and the vapor and liquid contacting throughout the column results in two product streams of high purity oxygen and nitrogen. The nitrogen stream exits the top of the column and the
oxygen stream collects as a liquid in the bottom of the column where a portion is reboiled by contacting it with the condensing stream in T-100A. The same design process used for T-100A was used for this column. The first task was to determine the number of stages needed in the column to achieve the desired separation while keeping column size as small as possible. For the analysis, the outlet composition of the oxygen stream was set at 98 mol% oxygen and the resulting boilup ratios and overhead oxygen composition were plotted. It is desired to minimize boilup in the column and to minimize the oxygen composition of the overhead stream because this is wasted oxygen that is ultimately lost to the environment. The results of this analysis are shown in Figure 50.

![Low Pressure Column Performance](image)

**Figure 50: Performance of T-101A as the number of theoretical stages in the tower is varied**

The plot shows that at 30 stages, both the boilup ratio and the overhead oxygen composition begin to level off at minimum values so we decided that 30 stages was the optimal number of stages for the column.

The next step of the process was to size the column appropriately using the tray sizing utility in UNISIM. The specifications for T-101A are shown in Table 26.
Table 26: Specifications for low pressure column T-101A

<table>
<thead>
<tr>
<th>Low Pressure Column T-101A Specifications</th>
</tr>
</thead>
<tbody>
<tr>
<td>Internals</td>
</tr>
<tr>
<td>Column Diameter</td>
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<tr>
<td>Flooding</td>
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<tr>
<td>Cross Sectional Area</td>
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<tr>
<td>Column Height</td>
</tr>
<tr>
<td>Column Pressure Drop</td>
</tr>
<tr>
<td>HETP</td>
</tr>
<tr>
<td>Packing Type</td>
</tr>
</tbody>
</table>

The overhead stream is 98.0 mol% nitrogen leaves T-101A at -313.8°F and the bottoms stream leaves the T-101A with a composition of 98% oxygen at -290.8°F. The overhead stream is next sent to the subcooler E-104A to cool two streams leaving T-100A and the bottoms product is sent to P-100A to increase its pressure to the gasifier operating pressure.

4.2.2.8 Liquid Oxygen Pump – P-100A

The purpose of P-100A is to increase the pressure of the oxygen product stream from 21.8 psi (1.5 bar) to 1000 psi (69.0 bar) because that is the pressure that it must be delivered to the gasifier. A centrifugal pump made of stainless steel is used with an electric, explosion proof motor. The motor must be explosion proof because the pump is operating in a high oxygen environment and a spark could provide a source of ignition to a potentially explosive liquid. Once the pressure of the liquid oxygen stream has been increased to the desired pressure, the stream is sent to the main heat exchanger E-103A where it is used to cool the incoming compressed air streams.

4.2.2.9 System Exit Stream

After exiting T-101A, the two product streams are used to cool warmer incoming streams. First, the overhead N₂ product stream is sent through E-104A to cool the two streams leaving T-100A. Then the N₂ product stream along with the O₂ product stream are used to cool the incoming air stream in the main heat exchanger E-103A. The oxygen stream is vaporized and leaves at 212°F for insertion into the gasifier. After the two product streams have been used to cool the incoming compressed air stream in the E-103A, the nitrogen stream is then directed through either V-100A or V-101A to regenerate the adsorbent in the vessel and then released into the atmosphere. The oxygen stream is sent from the air separation unit to the gasification block where it is fed into R-100G.
5 Turbine Block

The turbine block is the area of the power plant that uses the products from the water gas shift reactor to create work and subsequently generate electricity. Team 16 was responsible for the turbine design and specifications and detailed design information can be found in their report. We were responsible for modeling the turbine block in UNISIM to determine total power output of the combine cycle and designing equipment associated with the turbines, such as compressors, pumps, and heat exchangers.

5.1 Process Overview

The turbine block is a combined cycle, meaning that first the hydrogen gas from the water gas shift reactors is combusted in a fuel gas turbine and then the heat from this turbine’s exit stream is used to create steam to turn a secondary steam turbine. Using two turbines increases the overall efficiency of the plant by producing more energy by recovering as much heat as possible.

The first step in the combined cycle involves preparing the input streams for combustion in the fuel gas turbine. The cleaned hydrogen stream and an input air stream are compressed to 2180 psi and introduced to the turbine where the H₂ gas is combusted to H₂O. This stream leaves the turbine at 14.7 psi and 1400°F. This hot gas stream is then cooled to 158°F by exchanging heat to produce steam for the second turbine.

The steam for the second turbine is heated from 125°F to 1100°F and is at 2200 psi. This steam is sent through the second turbine and leaves at 165°F and 2.9 psi. This stream is then condensed in a heat exchanger using cooling water and is then pumped back up to 2200 psi to be recycled to be heated and sent through the turbine again.
This sketch is supplied for the convenience of engineers and does not represent the plant design in entirety.

Figure 51: Process flow diagram of the Turbine Block
5.2  Detailed Process Description

Figure 51 is a comprehensive flowsheet that gives a detailed view of the equipment involved in the combined cycle process.

5.2.1  Hydrogen Stream Compressor – K-100T

The purpose of the hydrogen compressor is to take the hydrogen stream from T-100C and compress it up to 2200 psi for injection into the fuel gas turbine. This stream leaves the compressor at 245°F. It is then heated in E-100T up to 320°F. After heating, it is combined with the air stream incoming from the compressors and sent to K-100FG for combustion.

5.2.2  Inlet Air Compressors – K-101T, K-102T, K-103T, K-104T

The inlet air compressors take atmospheric air and compress it up to 2200 psi for injection into the fuel gas turbine. The inlet air is at an atmospheric 14.7 psi. The overall compression ratio for the inlet stream is about 150, necessitating the use of a four stage compressor with interstage cooling. The optimal pressure leaving each stage was calculated using Equation 39. The inlet air is initially compressed from 14.7 psi and 77°F to 51.3 psi and 378°F in K-101T. After cooling the stream down to 95°F in E-101T, the stream is compressed a second time in K-102T to 179 psi. This stream is then cooled in E-102T and compressed a third time in K-103T to 625 psi. After a third cooling in E-103T, it is finally compressed to 2200 psi and leaves the compressor at 405°F. This stream is then mixed with the compressed hydrogen stream and sent to the K-100FG for combustion.

5.2.3  Fuel Gas Turbine – K-100FG

The fuel gas turbine accepts the mixed streams from K-100T and K-104T at 2200 psi and the mixture is combusted to create work in the turbine. The resulting mixture of N₂, O₂, and H₂O is expanded to 14.7 psi and leaves the turbine at 1400°F. The work on the turbine produces 990 MW of power. The hot exit stream is then cooled to 166°F in E-104T and the heat is used to create stream for K-100ST. The exit stream is then released to the environment.

5.2.4  Steam Turbine – K-100ST

The steam leaving E-104T is at 2200 psi and 1100°F. This steam is sent into turbine K-100ST where it is expanded to 2.9 psi and leaves at 165°F. This process creates 126 MW of power. The expanded vapor stream is condensed in E-105T by cooling the stream to 122°F in a heat exchanger using cooling water. This liquid stream is then pumped by P-100T back up to 2200 psi and is sent back to E-104T for reheating.

5.3  Parasitic Power Consumption

Overall, each turbine unit in the IGCC plant is outputting 1100 MW of power leading to an overall energy production of 2200 MW. However, significant portions of this energy are consumed by the plant itself running pumps, compressors, and other equipment. In the integrated UNISIM model we took all these
parasitic energy sources into account and the overall sellable energy output of the IGCC plant is 1000 MW.
Figure 52: Process flow diagram of the desulfurization block
6 Separations Blocks

The separations units for removing hydrogen sulfide and carbon dioxide were designed by Team 14 using out specifications for the outlet streams from the gasifier and water gas shift reactors. The following detailed process descriptions are excerpts from Team 14’s design report and are meant to provide an overview of the desulfurization process and the CO₂ sequestration process. For more detailed design information, refer to the Team 14 report.²⁵¹

6.1 Desulfurization Block Detailed Process Description

The desulfurization block was designed by Team 14 and optimized using the output specifications of our gasifier design. A comprehensive flowsheet of the desulfurization process is shown in Figure 51.

6.1.1 Absorber – T-100S

The separation of the hydrogen sulfide from the exit stream from the gasifier occurs in the high pressure absorption column. The solvent being used is a poly-ethylene glycol derivative called Selexol produced by Dow. The main flow leaving the gasifier is fed into the bottom stage of T-100S which is a sieve trayed column. As the incoming gas rises through the column, it contacts the falling liquid at each sieve tray transferring the hydrogen sulfide from the gas stream to the liquid stream. The gas that leaves the top of the column contains only 3% of the hydrogen sulfide that entered from the bottom.

There are multiple glycol feeds with varying levels of solvent purity at various stages to ensure the maximum amount of hydrogen sulfide is absorbed while maintaining low capital costs.

For the detailed design of the absorption column, the main independent design parameters that we examined were the number of stages in the column, the temperature of the solvent entering the column, the mass flow rate of solvent entering the column, and the fraction split occurring after flash tank 2 (V-101C). The stream after V-101C is due to the stream being split into a lean and super lean glycol stream, where the super lean glycol stream enters the column at a higher stage in order to remove any residual hydrogen sulfide not removed by the lean stream re-entering the column.

The major design component of the column absorber design was determining the appropriate number of theoretical equilibrium stages to achieve maximum separation of the components while keeping the column as small as possible. We varied the number of stages in the absorption column to see how the hydrogen sulfide separation was affected, and the required number needed to achieve our desired purity level. This study showed that 95% hydrogen sulfide removal, or greater, could be accomplished by using 20 stages or greater as shown in the Figure 53. The column did not converge below 5 stages or above 40 stages, but since at least removal to 70 ppm was being achieved this was seen as unnecessary.

²⁵¹ DeRose, Hoeksema, Markosky, & Reese, 2012
Figure 53: Fraction removal of hydrogen sulfide as the number of stages in the tower increased

Figure 53 also shows the fraction hydrogen sulfide removal for our particular process, examining how much is removed by the absorber, and the fraction hydrogen sulfide removal for the entire plant. The additional hydrogen sulfide being added to the entire plant comes from the CO in the incoming stream, which is not removed by Selexol, and which is combusted to hydrogen sulfide in the combined cycle. The optimal number of stages was found to be 20 in order to remove as much of the hydrogen sulfide as possible in the case of our specific absorption process and the overall plant.

The effect of glycol temperature entering the inlet of the absorption column also has a significant effect on the effectiveness of the separation. Our process achieved a better separation if the Selexol was as cold as possible, as shown in the figure below. However, we assumed that 35°C was the lowest temperature that could be used for the inlet Selexol as any lower temperatures would require refrigeration, which would require additional capital.
Figure 54: Fraction removal of hydrogen sulfide as a function of Selexol feed temperature

In Figure 54, the fraction removal for the overall plant versus the absorber is different for the same reasons mentioned before; more hydrogen sulfide is being created in portions of the plant outside of our individual group’s project scope.

Figure 55: Fraction removal of hydrogen sulfide as a function of mass flow rate of Selexol
The disadvantage with increasing the flow rate of Selexol solvent through the process is that the overall power output of the plant decreases significantly as more power has to be recycled into the pumps in the lean and super lean glycol streams to get the additional fluid to an acceptable pressure. We kept the flow rate as low as possible, while altering the other independent design parameters to obtain the desired amount of hydrogen sulfide separation.

6.1.2 Solvent Regeneration Flash Tank Series – V-100S and V-101S

The separation of the hydrogen sulfide from the Selexol stream leaving from E-102S-B occurs in the high pressure flash tanks. The main flow leaving E-102S-B is fed into the bottom of V-100S which is a simple pressure vessel that separates hydrogen sulfide from the Selexol. The pressure is dropped before each vessel to change the equilibrium vapor pressure which allows the hydrogen sulfide to escape from the solvent. The temperature and pressure are set by preheating heat exchangers and valves.

6.1.3 Absorber Pre-Cooling – E-101S

The purpose of E-101S heat exchangers is to decrease the temperature of the flue gas from the gasifier 2552°F to 95°F because that is the optimum temperature that the absorber works. The steam being generated is being used to preheat the gas going into the fuel gas turbine to maximize power generation efficiency.

6.1.4 Flash Tank Preheat – E-102S-A and E-102S-B

The purpose of the E-102S series of heat exchangers is to increase the temperature of the hydrogen sulfide rich glycol stream from 115 °F to 475 °F in order to raise the vapor pressure of the hydrogen sulfide for solvent regeneration in the flash tanks. The first exchanger, E-102S-A, is being heated by the stream exiting the second flash tank (V-101S). The second, E-102SC-B, has process steam running through it to get the stream up to the temperature needed for solvent regeneration.

6.1.5 Solvent Recovery Cooling – E-104S

The purpose of the E-104S heat exchanger is to decrease the temperature of the hydrogen sulfide rich stream exiting the two flash tanks from 467 °F to 185 °F in order to lower the vapor pressure of the Selexol for solvent recovery in the flash tank (V-102C). E-104S has cooling water running through it to get the stream down to the temperature needed for solvent recovery.

6.1.6 Solvent Recycle Cooling – E-103S

The purpose of E-103S of heat exchangers is to decrease the temperature of the lean glycol stream leaving the two flash tanks from 474 °F to 95 °F because that is the temperature needed in the Absorber for hydrogen sulfide removal. E-103C, is being cooled by cooling water and being sent to preheat hydrogen sulfide rich glycol stream into the flash tanks.

6.1.7 Selexol Recycle Pump – P-100S

The purpose of P-100S is to increase the pressure of the recycled Selexol stream from 87 psi to 609 psi because that is the pressure that it must be delivered to the absorption column. A centrifugal pump
made of stainless steel is used with an electric motor. Once the pressure of the Selexol stream has been increased to the desired pressure, the stream is sent to the main heat exchanger E-102S-A where it is cooled to the desired temperature.

6.1.8 Selexol Recovery Pump – P-101S

The purpose of P-101S is to increase the pressure of the recycled Selexol stream from 87 psi to 609 psi because that is the pressure that it must be delivered to the absorption column. A centrifugal pump made of stainless steel is used with an electric motor. Once the pressure of the Selexol stream has been increased to the desired pressure, the stream is sent to the absorption column.
Figure 56: Process flow diagram of the CO₂ sequestration unit
6.2 CO₂ Sequestration Block

The desulfurization block was designed by Team 14 and optimized using the output specifications of our water gas shift reactor design. A comprehensive flowsheet of the desulfurization process is shown in Figure 56.

6.2.1 Absorber – T-100C

The separation of the carbon dioxide from the exit stream from the water gas shift reactor occurs in the high pressure absorption column. The solvent being used is a poly-ethylene glycol derivative called Selexol produced by Dow. The main flow leaving V-100W is fed into the bottom stage of T-100C which is a sieve trayed column. As the incoming gas rises through the column, it contacts the falling liquid at each sieve tray transferring the carbon dioxide from the gas stream to the liquid stream. The gas that leaves the top of the column contains only 3% of the carbon dioxide that entered from the bottom.

There are multiple glycol feeds with varying levels of solvent purity at various stages to ensure the maximum amount of carbon dioxide is absorbed while maintaining low capital costs.

For the detailed design of the absorption column, the main independent design parameters that we examined were the number of stages in the column, the temperature of the solvent entering the column, the mass flow rate of solvent entering the column, and the fraction split occurring after flash tank 2 (V-101C). The stream after V-101C is due to the stream being split into a lean and super lean glycol stream, where the super lean glycol stream enters the column at a higher stage in order to remove any residual carbon dioxide not removed by the lean stream re-entering the column.

The major design component of the column absorber design was determining the appropriate number of theoretical equilibrium stages to achieve maximum separation of the components while keeping the column as small as possible. We varied the number of stages in the absorption column to see how the carbon dioxide separation was affected, and the required number needed to achieve our desired purity level. This study showed that 95% carbon dioxide removal, or greater, could be accomplished by using 12 stages or greater as shown in Figure 57. The column did not converge below 4 stages or above 13 stages, but since at least 95% removal was being achieved this was seen as unnecessary.
Figure 57: Fraction removal of carbon dioxide as a function of the number of stages in the column

Figure 57 also shows the fraction carbon dioxide removal for our particular process, examining how much is removed by the absorber, and the fraction carbon dioxide removal for the entire plant. The additional carbon dioxide being added to the entire plant comes from the CO in the incoming stream, which is not removed by Selexol, and which is combusted to carbon dioxide in the combined cycle. Since the CO is converted to carbon dioxide outside of our process, either in the water gas shift reactors or the combined cycle, we focused on separating as much of the carbon dioxide we could in our physical absorption process. The optimal number of stages was found to be 13, in order to remove as much of the carbon dioxide as possible in the case of our specific absorption process and the overall plant.

The effect of glycol temperature entering the inlet of the absorption column also has a significant effect on the effectiveness of the separation. Our process achieved a better separation if the Selexol was as cold as possible, as shown in the figure below. However, we assumed that 35°C was the lowest temperature that could be used for the inlet Selexol as any lower temperatures would require refrigeration, which would require additional capital.
In Figure 58, the fraction removal for the overall plant versus the absorber is different for the same reasons mentioned before; more CARBON DIOXIDE is being created in portions of the plant outside of our individual group’s project scope.

The mass flow rate of Selexol was found to increase the amount of carbon dioxide being removed from the flue gas stream. The mass flow rate of the super lean glycol stream was varied, which was deemed sufficient for a flow rate study since the lean and super lean streams have a directly proportional flow rate. The Figure 59 shows increase in carbon dioxide removal for a given mass flow rate of Selexol.

The disadvantage with increasing the flow rate of Selexol solvent through the process is that the overall power output of the plant decreases significantly as more power has to be recycled into the pumps in the lean and super lean glycol streams to get the additional fluid to an acceptable pressure. We kept the flow rate as low as possible, while altering the other independent design parameters to obtain the desired amount of carbon dioxide separation.
6.2.2 Flash Tank Preheat – E-100C-A and E-100C-B

The purpose of the E-100C series of heat exchangers is to increase the temperature of the carbon dioxide rich glycol stream from 115 °F to 475 °F in order to raise the vapor pressure of the carbon dioxide for solvent regeneration in the flash tanks. The first exchanger, E-100C-A, is being heated by steam generated in the solvent recycle cooling streams (E-102C & E-103C). The second, E-100C-B, has process steam running through it to get the stream up to the temperature needed for solvent regeneration.

6.2.3 Solvent Recovery Cooling – E-101C-A and E-101C-B

The purpose of the E-101C series of heat exchangers is to decrease the temperature of the carbon dioxide stream exiting the three flash tanks from 467 °F to 185 °F in order to lower the vapor pressure of the Selexol for solvent recovery in the flash tank (V-105C). The first exchanger, E-101C-A, is being cooled by steam generated in the solvent recycle cooling streams (E-102C & E-103C). The second, E-101C-B, has cooling water running through it to get the stream down to the temperature needed for solvent recovery.

6.2.4 Solvent Recycle Cooling – E-102C-A and E-103C-B

The purpose of E-102C and E-103C of heat exchangers is to decrease the temperature of the lean and super lean glycol stream leaving the three flash tanks from 474 °F to 95 °F because that is the temperature needed in the Absorber for carbon dioxide removal. The first exchanger, E-102C, is being cooled by cooling water and being sent to preheat carbon dioxide rich glycol stream into the flash tanks.
The second, E-103C, is being cooled by cooling water and being sent to preheat carbon dioxide rich glycol stream into the flash tanks.

### 6.2.5 Solvent Regeneration Flash Tank Series – V-100C, V-101C, and V-102C

The separation of the carbon dioxide from the Selexol stream leaving from E-100C-B occurs in the high pressure flash tanks. The main flow leaving E-100C-B is fed into the bottom of V-100C which is a simple pressure vessel that separates carbon dioxide from the Selexol. The pressure is dropped before each vessel to change the equilibrium vapor pressure which allows the carbon dioxide to escape from the solvent. The temperature and pressure are set by preheating heat exchangers and valves.

The way that the stream was split, into the lean glycol and super lean glycol streams, also had a significant effect on the amount of carbon dioxide being removed. The amount that needs to be split into each stream is a function of the additional separation that is obtained from sending the super lean glycol through the third flash tank. Different combinations of splits were examined, more specifically how much is being sent to the lean glycol stream, which can be seen in Figure 60. It can be observed that there is a definite peak in carbon dioxide removal for the ratio of 45% being directly recycled and 55% being sent to V-102C prior to recycling.

![Column Performance - Flow Split](image)

*Figure 60: Fraction removal of carbon dioxide as a function of lean glycol stream ratio*

### 6.2.6 Selexol Recycle Pump – P-100C

The purpose of P-100C is to increase the pressure of the recycled Selexol stream from 71.6 psi to 1160 psi because that is the pressure that it must be delivered to the absorption column. A centrifugal pump made of stainless steel is used with an electric motor. Once the pressure of the Selexol stream has been
increased to the desired pressure, the stream is sent to the main heat exchanger E-102C where it is cooled to the desired temperature.

6.2.7 Selexol Recycle Pump – P-101C

The purpose of P-101C is to increase the pressure of the recycled Selexol stream from 144 psi to 1160 psi because that is the pressure that it must be delivered to the absorption column. A centrifugal pump made of stainless steel is used with an electric motor. Once the pressure of the Selexol stream has been increased to the desired pressure, the stream is sent to the main heat exchanger E-103C where it is cooled to the desired temperature.

6.2.8 Selexol Recovery Pump – P-102C

The purpose of P-102C is to increase the pressure of the recovered Selexol stream from 72.5 psi to 1160 psi because that is the pressure that it must be delivered to the absorption column. A centrifugal pump made of stainless steel is used with an electric motor. Once the pressure of the Selexol stream has been increased to the desired pressure, the stream is sent to the absorption tower.

6.2.9 Solvent Recovery Flash Tank – V-105C

V-105C is a crucial in Selexol recovery. Coming out of V-100C, V-101C, V-102C, there is a considerable amount of Selexol being lost. The pressure remains constant but the temperature is set by E-101C-B which lowers the equilibrium vapor pressure of the Selexol allowing it to condense. V-105C is merely there to separate the liquid solvent from the gas where it can then be recycled to T-100C.
7 UNISIM Model Integration

During initial design work, each team developed their own UNISIM models for the specific processes they were designing. Once design work was nearing completion, it was necessary for up to integrate all the models for the different unit operations into one comprehensive flowsheet to find the total energy output, energy consumption, and efficiency to the plant. Team 15 contributed design models for the cryogenic distillation unit, the gasification unit, the water gas shift reactors, and the turbine unit. Team 14 contributed their designs for the sulfur removal unit and the CO\textsubscript{2} removal unit.

After all the models were integrated into one UNISIM flowsheet, we integrated as many process streams as possible to provide heating and cooling inside the plant and determined the amount of cooling water that would need to be drawn from Lake Michigan to run the plant. Once the energy integration for heat exchange was complete, we summed up all the parasitic energy streams and subtracted them from the turbine output to determine an overall plant output. The UNSIM model was then optimized by adjusting the coal input to provide 1000 MW of sellable power.
8 Cost Analysis

8.1 Chemical Engineering Plant Cost Index

When calculating costs for equipment and determining overall costs for a chemical plant, it is essential that all dollar amounts be scaled to the same year to make sure analysis is performed on a consistent basis. For this IGCC plant, the purchase cost of most processing equipment was determined using equations in Warren Seider’s “Product and Process Design Principles.” However, costs are not static and generally increase due to inflation. Therefore, it is necessary to use a cost index to scale costs from an earlier date using Equation 40.

\[
Cost = BaseCost \left( \frac{I}{I_{base}} \right) \quad \text{Eq 40}
\]

There are several different cost indexes commonly used by engineers, but for the purposes for this design we chose to use the Chemical Engineering (CE) Plant Cost Index because this is the index applied to the cost equations in Seider.\(^{152}\)

All costs associated with the IGCC plant are scaled to the year 2015 because this is the estimated year that the plant would come online. While cost indexes are very useful for scaling past costs to present values, the index does not provide predicted values for future years. Therefore, we took all the available data for the CE index through 2010 and then, by observing the cost trends, made a prediction of what the values will be through year 2015.\(^{153}\) A plot of all CE index values from 1980 to 2010 as well as our predicted values through 2015 is shown in Figure 61. As shown on the plot, the index values increased at a relatively constant rate from 1981 to 2003. From 2004 to 2008, the index increased much faster and then dropped sharply in 2009. For our index prediction, we realize that the sharp increase from 2004 to 2008 was not sustainable and not representative of the overall trend. We therefore predicted for the years after 2010, the index will return to its linear trend observed during the 1981 to 2003 period. This prediction sets the 2015 CE Index value at 574.

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\(^{152}\) Seider, Deader, Lewin, & Widagdo, 2009

\(^{153}\) Seider, Deader, Lewin, & Widagdo, 2009
8.2 Equipment Cost

To determine the total capital investment of the IGCC plant, it was necessary to determine a cost for each piece of equipment. This estimation was performed using the Guthrie method described in Chapter 22 of Seider’s book. The majority of equipment costs were calculated using equations in Section 22.5. All costing equations in this book are correlations developed for process equipment costs using 2006 dollars (CE Index=500). Therefore, all equipment costs calculated using these equations needed to be scaled up to estimated 2015 dollars using the factor 574/500. Any costs found using other sources were scaled accordingly using the proper index value for the base cost year. The costs calculated using these equations include all costs for direct materials and labor, shipping, installation, insurance, taxes, overhead, and engineering.\textsuperscript{154} The costs for each piece of equipment are shown in Table 27. The calculations performed to obtain these numbers are shown in Appendix F.

\textsuperscript{154} Seider, Deader, Lewin, & Widagdo, 2009
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</tbody>
</table>
The IGCC plant being design involves two equipment trains each capable of producing 500 MW. Therefore, the table above shows the equipment needed for one 500 MW train and all values must be doubled to obtain the cost of the entire plant. These doubled costs are shown in Table 28.

<table>
<thead>
<tr>
<th>Area Name</th>
<th>Total Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Air Separation Unit</td>
<td>$ 111,764,000.00</td>
</tr>
<tr>
<td>Gasification Block</td>
<td>$ 369,774,000.00</td>
</tr>
<tr>
<td>CO2 Sequestration Unit</td>
<td>$ 36,764,000.00</td>
</tr>
<tr>
<td>Desulfurization Unit</td>
<td>$ 47,066,000.00</td>
</tr>
<tr>
<td>Turbine Unit</td>
<td>$ 612,600,000.00</td>
</tr>
<tr>
<td>Water Gas Shift Unit</td>
<td>$ 412,906,000.00</td>
</tr>
<tr>
<td><strong>Total 1000 MW Bare Module Cost</strong></td>
<td><strong>$ 1,590,874,000.00</strong></td>
</tr>
</tbody>
</table>

### 8.3 Total Capital Investment

While the value reported in Table 28 accounts for the bare module cost for all the equipment, it does not represent the true cost of the power plant. Other factors such as the cost of land, buildings, offsite facilities, and working capital must be considered. The total capital investment of the plant is approximated by Equation 41.\(^{155}\)

\[
C_{TCI} = 1.18\left(C_{TBM} + C_{site} + C_{buildings} + C_{offsite \ facilities}\right) + C_{WC} \
\text{Eq 41}
\]

According to Seider, \(C_{site}\) is approximately 10% to \(C_{TBM}\), \(C_{buildings}\) is approximately 20% of \(C_{TBM}\), and \(C_{WC}\) is approximately 17.6% of \(C_{TCI}\). There are no offsite facilities with this plant so \(C_{offsite \ facilities}\) is eliminated. Using Equation X and the previous assumptions, the total capital investment for the 1000 MW IGCC plant was calculated and is presented in Table 29.

| 1000 MW IGCC Total Capital Investment | $ 2,962,000,000 |

\(^{155}\) Seider, Deader, Lewin, & Widagdo, 2009
8.4 Operating Cost

In addition to the total capital investment, it was also necessary to determine the cost of operating the plant. Utilities, labor, and other costs were estimated using operating cost information in Chapter 23 of ‘Product and Process Design Principles.’ Using the information here as well as recommendations from Dr. Wayne Wentzheimer, we estimated total operating costs of the plant per year. These costs are summarized in Table 31. These costs represent the operating costs for one 500 MW train of the IGCC. Therefore, the total amount must be doubled to reflect the total operating cost of the 1000 MW plant, as shown in the final row of Table 31.

8.5 Cost of Electricity

Finally, after determining the total capital investment needed to build the plant and the operating expenses for the plant, it is possible to determine the breakeven price and a selling price that will provide a reasonable rate of return for the plant. Based on research on operating IGCC plants, a reasonable internal rate of return (IRR) for an IGCC plant is 12%.

Initially, a breakeven cost was determined using a method shown in Chapter 7 of ‘Thermal Design and Optimization’ by Adrian Bejan. These calculations were done using the Engineering Equation Solver (EES) and are shown in Appendix G. A cash flow analysis was then performed to determine the price electricity would need to be sold at to generate a 12% IRR. This sheet is also shown in Appendix G. The summary of these results is shown in Table 30.

| Table 30: Breakeven price and selling price of electricity from the 1000 MW IGCC plant |
|-----------------------------------------------|------------------|
| Break Even Price                              | $0.14/kWh        |
| Selling Price                                 | $0.18/kWh        |
Table 31: Production costs associated with a 500 MW equipment train of IGCC plant and total 1000 MW production cost

<table>
<thead>
<tr>
<th>Yearly 500 MW IGCC Operating Costs</th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Feedstock</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Coal</td>
<td>$257,800,000</td>
<td></td>
</tr>
<tr>
<td>Selexol -make up</td>
<td>$62,000</td>
<td></td>
</tr>
<tr>
<td><strong>Utilities</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Cooling Water</td>
<td></td>
<td></td>
</tr>
<tr>
<td>E-100A</td>
<td>$335,000</td>
<td></td>
</tr>
<tr>
<td>Combined IGCC</td>
<td>$545,000</td>
<td></td>
</tr>
<tr>
<td>Process Water</td>
<td>$568,000</td>
<td></td>
</tr>
<tr>
<td><strong>Operations</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Direct Wages and Benefits</td>
<td>$4,502,000</td>
<td></td>
</tr>
<tr>
<td>Direct Salaries and Benefits</td>
<td>$675,000</td>
<td></td>
</tr>
<tr>
<td>Operating Supplies and Services</td>
<td>$270,000</td>
<td></td>
</tr>
<tr>
<td><strong>Maintenance</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Wages and Benefits</td>
<td>$35,795,000</td>
<td></td>
</tr>
<tr>
<td>Salaries and Benefits</td>
<td>$8,949,000</td>
<td></td>
</tr>
<tr>
<td>Materials and Services</td>
<td>$35,795,000</td>
<td></td>
</tr>
<tr>
<td>Maintenance Overhead</td>
<td>$1,790,000</td>
<td></td>
</tr>
<tr>
<td><strong>Operating Overhead</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>General Plant Overhead</td>
<td>$6,232,000</td>
<td></td>
</tr>
<tr>
<td>Mechanical Department Services</td>
<td>$2,107,000</td>
<td></td>
</tr>
<tr>
<td>Employee Relations Department</td>
<td>$5,179,000</td>
<td></td>
</tr>
<tr>
<td>Business Services</td>
<td>$6,495,000</td>
<td></td>
</tr>
<tr>
<td>Property Taxes and Insurance</td>
<td>$15,909,000</td>
<td></td>
</tr>
<tr>
<td><strong>Depreciation</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Direct Plant</td>
<td>$36,113,000</td>
<td></td>
</tr>
<tr>
<td><strong>General Expenses</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Selling expense</td>
<td>$6,160,000</td>
<td></td>
</tr>
<tr>
<td>Direct Research</td>
<td>$6,160,000</td>
<td></td>
</tr>
<tr>
<td>Allocated Research</td>
<td>$3,080,000</td>
<td></td>
</tr>
<tr>
<td>Administrative Expense</td>
<td>$12,320,000</td>
<td></td>
</tr>
<tr>
<td>Management Incentive Compensation</td>
<td>$7,700,000</td>
<td></td>
</tr>
<tr>
<td><strong>Total Production Cost</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Total Production Cost 500 MW</td>
<td>$454,538,000</td>
<td></td>
</tr>
<tr>
<td>Total Production Cost 1000 MW</td>
<td>$909,077,000</td>
<td></td>
</tr>
</tbody>
</table>
9 Conclusion

This report outlining the design on an IGCC power plant with CO$_2$ sequestration was compiled with the ultimate goal of determining what the cost of electricity consumers would have to pay to feasibly implement clean coal power in the United States. With its vast resources, the U.S. is likely to continue to use coal as a cheap energy feedstock to provide electricity but with recent EPA restrictions on CO$_2$ emissions, it has become impossible to build a new coal fired power plant without integrating some type of CO$_2$ sequestration technology into the plant design.

Preliminary design, modeling, and analysis for each of the major units in an IGCC plant, including the air separation unit, the gasifier, the water gas shift reactors, the separations units, and the turbine unit, were included as part of this report to determine an estimate of the total capital investment that would be required to construct a new IGCC plant with sequestration technology in West Olive, MI. The resulting power plant is capable of providing 1000 MW to the community while sequestering over 90% of the CO$_2$ emissions created from the gasification of coal.

In addition to providing capital investment estimation, we also created a cost sheet outlining all the associated costs of operating the plant every year. This cost sheet includes utilities, operations, maintenance, overhead, taxes and insurance, depreciation, and general expenses to determine a total production cost.

Using both the operating and capital costs for the IGCC plant, a breakeven price of electricity was calculated to be $0.14/kWh. After performing a cash flow analysis, we calculated that a 12% internal rate of return would be achieved with a price of $0.18/kWh. Currently, the selling price of electricity is approximately $0.10/kWh in West Michigan. If an IGCC plant were to be built in West Olive, MI and came online in 2015, consumers would see an 80% increase in their cost of electricity over today's prices. However, as restrictions are placed on the coal power industry, this may become a necessary cost to continue to meet this country's every increasing need for energy. Coal power is one of the few sources currently capable of producing enough energy to meet consumer demand and is not likely to disappear in the near future.

As illustrated through this design project, technology currently exists to produce electricity cleanly from coal and within the guidelines of any current and foreseeable future regulations. However, there is a higher cost associated with this newer technology compared to the old. It is up to the American consumers to decide if they are willing to pay for it.
10 Bibliography


11 Acknowledgements

Dr. Wayne Wentzheimer
Dr. J. Aubrey Sykes
Mr. Randall Elenbaas
Calvin College Engineering Department
Teams 14 and 16
Family, Friends, and Spouses
12 Appendices

Appendix A: Coal Feed Heuristics
Appendix B: Equipment Specification Sheets
Appendix C: Polymath Models for Gasification Kinetics
Appendix D: First Semester Work Breakdown Schedule
Appendix E: Second Semester Work Breakdown Schedule
Appendix F: Equipment Cost Calculation Sheets
Appendix G: Break Even and Selling Price Calculation Sheets
Appendix H: Heat Exchanger Design Sheets
Appendix I: Radiant Syngas Cooler Design Sheets
Appendix A: Coal Feed Heuristics

Conveying granular solids

If the solid particles are small in size, low in particle density and not sticky or abrasive, use pneumatic conveying with air at 1 to 7 ft³/ft³ of solids and 35 to 120 ft/s velocity for distances up to 400 feet. Otherwise, for sticky and/or abrasive solids of any size and density, use a screw conveyor and/or bucket elevator for distances up to 150 ft. For solid particles of any size and shape, and not sticky, use a belt conveyor, with inclination up to 30° if necessary, for long distances up to a mile or more.

Solid particle size change and separation

Crushing of coarse solids. Use a jaw crusher to reduce lumps of hard, abrasive, and/or sticky materials of 4 inches to 3 feet in diameter to slabby particles of 1 to 4 inches in size. Use a gyratory crusher to reduce slabby materials of 8 inches to 6 feet in size to rounded particles of 1 to 10 inches in diameter. Use a cone crusher to reduce less hard and less sticky materials of 2 inches to 1 foot in diameter to particles of 0.2 inch (4 mesh) to 2 inches in diameter.

Grinding to fine solids. Use a rod mill to take particles of intermediate hardness as large as 20 mm and reduce them to particles in the range of 10 to 35 mesh. Use a ball mill to reduce particles of low to intermediate hardness of 1 to 10 mm in size to very small particles of less than 140 mesh.

Particle size enlargement. Use compression with rotary compression machines to convert powders and granules into tablets of up to 1.5 inches in diameter. Use extruders with cutters to make pellets and wafers from pastes and melts. Use roll compactors to produce sheets from finely divided materials; the sheets are then cut into any desired shape. Use rotary drum granulators and rotary disk granulators with binders to produce particles in the size range of 2 to 25 mm.

Size separation of particles. Use a grizzly of spaced, inclined, vibrated parallel bars to remove large particles greater than 2 inches in diameter. Use a revolving cylindrical perforated screen to remove intermediate-sized particles in the size range of 0.25 inch to 1.5 inches in diameter. Use flat, inclined woven screens (U.S. standard) that are vibrated, shaken, or impacted with bouncing balls to separate small particles in the size range of 3 to 80 mesh. Use an air classifier to separate fine particles smaller than 80 mesh.

Use a cyclone separator to remove, from a gas, droplets or solid particles of diameter down to 10 microns (0.01 mm). Use a hydroclone separator to remove, from a liquid, insoluble liquid droplets or solid particles of diameter down to 5 microns (0.005 mm). However, small amounts of entrained liquid droplets are commonly removed from gases by vertical knock-out drums equipped with mesh pads to help coalesce the smallest droplets.
Appendix B: Equipment Specification Sheets

The following equipment specification sheets provide summary details for all equipment in the IGCC power plant.

Major unit operations with specification sheets:

1) Air Separation Unit
2) Gasification Unit
3) Desulfurization Unit
4) CO$_2$ Sequestration Unit
5) Water Gas Shift Reactor Unit
6) Turbine Unit
### Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
<th>E-100A</th>
</tr>
</thead>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>CW-100A</td>
<td>4683000</td>
<td>2452000</td>
<td>2452000</td>
<td>4683000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>O2</th>
<th>N2</th>
<th>Ar</th>
<th>Water</th>
<th>CO2</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>4683000</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>567000</td>
<td>1852000</td>
<td>32000</td>
<td>0</td>
<td>1200</td>
</tr>
</tbody>
</table>

**T**\_Hot In (F): 312.8

**T**\_Hot Out (F): 95.0

**T**\_Cold In (F): 86.0

**T**\_Cold Out (F): 113.0

**Pressure (psi):** 38.4

**Design Data:**

- Type: Fixed Head Shell and Tube
- Materials of Construction: Carbon Steel Shell/Carbon Steel Tube
- Heat Duty: 38000 kW
- Shell Diameter: 37 in
- Heat Transfer Coefficient: 4000 Btu/(F-ft^2-hr)
- Heat Transfer Area: 18500 ft^2

**Comments:** E-100A is used to cool the compressed ambient air coming from K-100A. This interstage cooler increases compressor efficiency by cooling the stream before additional compression in K-101A.
## Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
<th>5/1/2012</th>
<th>E-101A</th>
</tr>
</thead>
</table>

### Function:
To exchange energy between hot and cold streams.

### Operation:
Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-118A</td>
<td>539000</td>
<td>567000</td>
<td>567000</td>
<td>539000</td>
</tr>
<tr>
<td>S-103A</td>
<td>2452000</td>
<td>2452000</td>
<td>2452000</td>
<td>552398</td>
</tr>
<tr>
<td>S-104A</td>
<td>0</td>
<td>1852000</td>
<td>1852000</td>
<td>0</td>
</tr>
<tr>
<td>S-119A</td>
<td>14000</td>
<td>32000</td>
<td>32000</td>
<td>14000</td>
</tr>
<tr>
<td>S-103A</td>
<td>0</td>
<td>1200</td>
<td>1200</td>
<td>0</td>
</tr>
</tbody>
</table>

### Composition:

<table>
<thead>
<tr>
<th></th>
<th>O2</th>
<th>N2</th>
<th>Ar</th>
<th>CO2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cold in</td>
<td>539000</td>
<td>0</td>
<td>14000</td>
<td>0</td>
</tr>
<tr>
<td>Hot In</td>
<td>567000</td>
<td>1852000</td>
<td>32000</td>
<td>1200</td>
</tr>
<tr>
<td>Hot Out</td>
<td>567000</td>
<td>1852000</td>
<td>32000</td>
<td>1200</td>
</tr>
<tr>
<td>Cold Out</td>
<td>539000</td>
<td>0</td>
<td>14000</td>
<td>0</td>
</tr>
</tbody>
</table>

### Temperature:

- $T_{Hot \ In}$ (F): 322.3
- $T_{Hot \ Out}$ (F): 251.7
- $T_{Cold \ In}$ (F): -91.5
- $T_{Cold \ Out}$ (F): 212.0

### Pressure (psi):
94.3

### Design Data:

<table>
<thead>
<tr>
<th>Type:</th>
<th>Fixed Head Shell and Tube</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon Steel Shell/Stainless Steel Tube</td>
<td></td>
</tr>
</tbody>
</table>

- Heat Duty: 12500 kW
- Shell Diameter: 37 in
- Heat Transfer Coefficient: 2000 Btu/(F-ft^2-hr)
- Heat Transfer Area: 10600 ft^2

### Comments:
E-101A is used to cool the compressed air coming from K-101A. This interstage cooler cools the compressed stream before pre-purification of CO2 in V-100A or V-101A and sent to the main heat exchanger.
## Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Heat Exchanger</th>
<th>Date:</th>
<th>5/1/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td>E-102A</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

### Function:
To exchange energy between hot and cold streams.

### Operation:
Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-122A</td>
<td>1898000</td>
<td>2452000</td>
<td>2452000</td>
<td>1898000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>29000</td>
<td>567000</td>
<td>567000</td>
<td>29000</td>
</tr>
<tr>
<td>N2</td>
<td>1852000</td>
<td>1852000</td>
<td>1852000</td>
<td>1852000</td>
</tr>
<tr>
<td>Ar</td>
<td>18000</td>
<td>32000</td>
<td>32000</td>
<td>18000</td>
</tr>
<tr>
<td>CO2</td>
<td>0</td>
<td>1200</td>
<td>1200</td>
<td>0</td>
</tr>
</tbody>
</table>

| Temperature | Value     | |
|-------------|-----------|
| $T_{\text{Hot In (F)}}$ | 251.7     |
| $T_{\text{Hot Out (F)}}$ | 95.0      |
| $T_{\text{Cold In (F)}}$ | 32.0      |
| $T_{\text{Cold Out (F)}}$ | 230.8     |

<table>
<thead>
<tr>
<th>Pressure (psi):</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>92.3</td>
</tr>
</tbody>
</table>

### Design Data:

<table>
<thead>
<tr>
<th>Type:</th>
<th>Fixed Head Shell and Tube</th>
</tr>
</thead>
<tbody>
<tr>
<td>Materials of Construction:</td>
<td>Carbon Steel Shell/Stainless Steel Tube</td>
</tr>
<tr>
<td>Heat Duty:</td>
<td>27500 kW</td>
</tr>
<tr>
<td>Shell Diameter:</td>
<td>37 in</td>
</tr>
<tr>
<td>Heat Transfer Coefficient:</td>
<td>86.1 Btu/(F-ft$^2$-hr)</td>
</tr>
<tr>
<td>Heat Transfer Area:</td>
<td>14000 ft$^2$</td>
</tr>
</tbody>
</table>

### Comments:
E-102A is used to cool the compressed air coming from K-101A. This interstage cooler cools the compressed stream before pre-purification of CO2 in V-100A or V-101A and sent to the main heat exchanger.
# Heat Exchanger

**Identification:** Heat Exchanger  
**Item Number:** E-103A  
**Date:** 4/18/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Cold 1 In</th>
<th>Cold 1 Out</th>
<th>Cold 2 In</th>
<th>Cold 2 Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-121A</td>
<td>1932000</td>
<td>1932000</td>
<td>563000</td>
<td>563000</td>
</tr>
<tr>
<td>S-122A</td>
<td>29000</td>
<td>29135</td>
<td>549000</td>
<td>549000</td>
</tr>
<tr>
<td>S-117A</td>
<td>1885000</td>
<td>1885103</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>A-118A</td>
<td></td>
<td></td>
<td>14000</td>
<td>14000</td>
</tr>
</tbody>
</table>

- **Composition:**
  - O₂: 29000, 29135, 549000, 549000
  - N₂: 1885000, 1885103, 0, 0
  - Ar: 18000, 18041, 14000, 14000

- **Temperature (F):**
  - Cold: -279.0, 32.0, -284.9, -138.7

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Hot 1 In</th>
<th>Hot 1 Out</th>
<th>Hot 2 Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-106A</td>
<td>2495000</td>
<td>2196000</td>
<td>299000</td>
</tr>
<tr>
<td>S-107A</td>
<td>578000</td>
<td>508000</td>
<td>69000</td>
</tr>
<tr>
<td>S-114A</td>
<td>1885000</td>
<td>1659000</td>
<td>226000</td>
</tr>
<tr>
<td></td>
<td>32000</td>
<td>28000</td>
<td>4000</td>
</tr>
</tbody>
</table>

- **Composition:**
  - O₂: 578000, 508000, 69000
  - N₂: 1885000, 1659000, 226000
  - Ar: 32000, 28000, 4000

- **Temperature (F):**
  - Hot: 77.0, -260.3, -171.6

## Design Data:

- **Type:** Plate Fin Heat Exchanger
- **Materials of Construction:** Brazed Aluminum
- **Heat Duty:** 58000 kW
- **Length:** 10.2 ft
- **Width:** 3.3 ft
- **Height:** 3.3 ft
- **Heat Transfer Area:** 101000 ft²

**Comments:** E-103A is a brazed aluminum plate fin heat exchanger that uses two cold process streams from T-101A to cool the inlet compressed air stream after an initial cooling in E-102A. The cooled air stream is split with 12% of flow sent to K-103A and the balance flow to T-100A.
### Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Heat Exchanger</th>
<th>Date:</th>
<th>5/3/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td></td>
<td>E-104A</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold Streams</th>
<th>Hot Streams</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Cold 1 In</td>
<td>Cold 1 Out</td>
</tr>
<tr>
<td></td>
<td>1129000</td>
<td>1129000</td>
</tr>
<tr>
<td></td>
<td>1028000</td>
<td>1028000</td>
</tr>
<tr>
<td></td>
<td>1129000</td>
<td>1129000</td>
</tr>
<tr>
<td></td>
<td>1028000</td>
<td>1028000</td>
</tr>
<tr>
<td></td>
<td>1129000</td>
<td>1129000</td>
</tr>
<tr>
<td></td>
<td>1028000</td>
<td>1028000</td>
</tr>
</tbody>
</table>

**Composition:**

- **O2**
  - Cold Streams: 29000, 29000
  - Hot Streams: 498000, 498198
  - Comments: E-103A is a brazed aluminum plate fin heat exchanger that uses the cold overhead nitrogen stream from T-101A to cool the overhead and distillate product streams from T-100A.

**Temperature (F):**

- Cold Streams: -313.8, -280.9
- Hot Streams: -279.1, -290.2, -287.2, -306.4

**Design Data:**

- **Type:** Plate Fin Heat Exchanger
- **Materials of Construction:** Brazed Aluminum
- **Heat Duty:** 4600 kW
- **Length:** 6.2 ft
- **Width:** 3.3 ft
- **Height:** 3.3 ft
- **Heat Transfer Area:** 30000 ft^2
## Compressor

**Identification:** Item Name: Compressor  
**Item Number:** K -100A  
**Date:** 5/1/2012

**Function:** To compress atmosphereic air for cryogenic distillation

**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-100A</td>
<td>2452000</td>
<td>2452000</td>
</tr>
<tr>
<td>S-101A</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Material</th>
<th>Inlet (lb/hr)</th>
<th>Out (lb/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>567000</td>
<td>567000</td>
</tr>
<tr>
<td>N2</td>
<td>1852000</td>
<td>1852000</td>
</tr>
<tr>
<td>Ar</td>
<td>32000</td>
<td>32000</td>
</tr>
<tr>
<td>CO2</td>
<td>1200</td>
<td>1200</td>
</tr>
</tbody>
</table>

- **T In (F):** 86.0
- **T Out (F):** 312.8
- **Pressure In (psi):** 14.7
- **Pressure Out(psi):** 38.4

### Design Data:

- **Type:** Centrifugal
- **Volumetric Flow Rate:** 33714000 ft³/hr
- **Pressure Change:** 23.8 psi
- **Power Required:** 40000 kW
- **Efficiency:** 0.75
- **Materials of Construction:** Carbon Steel

**Comments:** K-100A is a centrifugal that takes atmospheric air and compresses it for cryogenic distillation to 38.43 psi as the first stage of a 2-stage compression system.
# Compressor

**Identification:** Compressor  
**Item Number:** K-101A  
**Date:** 5/1/2012

**Function:** To compress air for cryogenic distillation  
**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>Stream ID:</td>
<td>S-102A</td>
<td>S-103A</td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>2452000</td>
<td>2452000</td>
</tr>
<tr>
<td>Composition:</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>O2</td>
<td>567000</td>
</tr>
<tr>
<td></td>
<td>N2</td>
<td>1852000</td>
</tr>
<tr>
<td></td>
<td>Ar</td>
<td>32000</td>
</tr>
<tr>
<td></td>
<td>CO2</td>
<td>1200</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th></th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>T In (F)</td>
<td>95.0</td>
<td></td>
</tr>
<tr>
<td>T Out (F)</td>
<td>322.2</td>
<td></td>
</tr>
<tr>
<td>Pressure In (psi):</td>
<td>36.4</td>
<td></td>
</tr>
<tr>
<td>Pressure Out (psi):</td>
<td>94.3</td>
<td></td>
</tr>
</tbody>
</table>

## Design Data:

<table>
<thead>
<tr>
<th>Type: Centrifugal</th>
<th>Volumetric Flow Rate: 13814000 ft³/hr</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure Change:</td>
<td>57.8 psi</td>
</tr>
<tr>
<td>Power Required:</td>
<td>40000 kW</td>
</tr>
<tr>
<td>Efficiency:</td>
<td>0.75</td>
</tr>
<tr>
<td>Materials of Construction: Carbon Steel</td>
<td></td>
</tr>
</tbody>
</table>

**Comments:** K-101A is the second compressor stage used to compress air for cryogenic distillation. Cooled air at 36.4 psi is received from E-100A and compressed to 94.27 psi. The compressed air is then sent through V-100A or V-101A for air prepurification.
**Turbine**

**Identification:** Turbine  
**Item Name:** Turbine  
**Item Number:** K-102A  
**Date:** 5/3/2012

**Function:** To expand compressed air  
**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-114A</td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-115A</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>294000</td>
<td>294000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>68000</td>
<td>68000</td>
</tr>
<tr>
<td>N2</td>
<td>222000</td>
<td>222000</td>
</tr>
<tr>
<td>Ar</td>
<td>3800</td>
<td>3800</td>
</tr>
</tbody>
</table>

| Inlet Temp (F): | -171.6 |
| Outlet Temp (F): | -246.5 |
| Pressure In (psi): | 85.4 |
| Pressure Out (psi): | 21.8 |

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Radial-Flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>355000 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>-63.6 psi</td>
</tr>
<tr>
<td>Power Generated:</td>
<td>1857 hp</td>
</tr>
<tr>
<td>Efficiency:</td>
<td>0.75</td>
</tr>
<tr>
<td>Materials of Construction:</td>
<td>Stainless Steel</td>
</tr>
</tbody>
</table>

**Comments:** K-103A expands a portion of the air compressed by K-101A from 85.4 psi to 21.8 psi. This process significantly drops the temperature of the stream and provides the necessary cryogenic refrigeration for the plant to operate. After expansion, the cold gas is fed into T-101A.
### Pump & Motor

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Pump</td>
<td>5/3/2012</td>
<td>P-100A</td>
</tr>
</tbody>
</table>

**Function:** To pump from T-101A to R-100G

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-116A</td>
<td>552000</td>
<td></td>
</tr>
<tr>
<td>S-117A</td>
<td>552000</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>539000</td>
<td>539000</td>
</tr>
<tr>
<td>N2</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>Ar</td>
<td>14000</td>
<td>14000</td>
</tr>
</tbody>
</table>

**Temperature (F):** -290.8

**Pressure In (psi):** 21.8

**Pressure Out (psi):** 1000

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Centrifugal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>7800 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>978.2 psi</td>
</tr>
<tr>
<td>Power Required:</td>
<td>742 hp</td>
</tr>
<tr>
<td>Efficiency:</td>
<td>0.78</td>
</tr>
</tbody>
</table>

**Comments:** P-100A increases the pressure of the O2 product out of T-101A from 21.8 psi to 1000 psi which is the operating pressure of R-100G. P-100A uses an explosion proof electric motor to mitigate the possibility of ignition in a high oxygen environment.
## Distillation Column

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Distillation Column</th>
<th>Date:</th>
<th>5/3/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Item Number:</td>
<td>T-100A</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To separate N2 from O2

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-107A</td>
<td>S-111A</td>
<td>S-108A</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>2157000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
</tr>
<tr>
<td>N2</td>
</tr>
<tr>
<td>Ar</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th></th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>499000</td>
<td>1200</td>
<td>498000</td>
</tr>
<tr>
<td>N2</td>
<td>1630000</td>
<td>1021000</td>
<td>609000</td>
</tr>
<tr>
<td>Ar</td>
<td>28000</td>
<td>6000</td>
<td>22000</td>
</tr>
</tbody>
</table>

| Temp Top (F): | -287.2 |
| Temp Bott (F): | -279.1 |

**Design Data:**

<table>
<thead>
<tr>
<th>Materials of Construction:</th>
<th>Stainless Steel</th>
</tr>
</thead>
<tbody>
<tr>
<td>Stages:</td>
<td>30</td>
</tr>
<tr>
<td>Pressure Drop:</td>
<td>0.54 psi</td>
</tr>
<tr>
<td>Stage Pressure Drop:</td>
<td>0.54 psi</td>
</tr>
<tr>
<td>Diameter:</td>
<td>22.5 ft</td>
</tr>
<tr>
<td>Height:</td>
<td>39.9 ft</td>
</tr>
<tr>
<td>Packing Type:</td>
<td>Norton Intalox</td>
</tr>
<tr>
<td>Theoretical Plate Height:</td>
<td>1.33 ft</td>
</tr>
</tbody>
</table>

**Comments:** T-100A is a high pressure, packed distillation column column used to do an initial separation of N2 and O2 to get a crude O2 product. It uses a structured metal packing to achieve separation effective seperation with very little pressure drop. The two product streams are then sent to T-101A for a second seperation.
# Distillation Column

**Identification:**

<table>
<thead>
<tr>
<th>Item Name</th>
<th>Distillation Column</th>
<th>Date</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number</td>
<td>T-101A</td>
<td>5/3/2012</td>
</tr>
</tbody>
</table>

**Function:** To separate N2 from O2

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed Streams</th>
<th>S-110A</th>
<th>S-113A</th>
<th>S-115A</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quantity (lb/hr)</td>
<td></td>
<td>1129000</td>
<td>1028000</td>
<td>294000</td>
</tr>
<tr>
<td>Composition:</td>
<td></td>
<td>O2: 498000</td>
<td>1100</td>
<td>68000</td>
</tr>
<tr>
<td></td>
<td></td>
<td>N2: 609000</td>
<td>1021000</td>
<td>222000</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Ar: 22000</td>
<td>6000</td>
<td>3800</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>S-120A</th>
<th>S-116A</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quantity (lb/hr)</td>
<td>1898000</td>
<td>552000</td>
</tr>
<tr>
<td>Composition:</td>
<td>O2: 29000</td>
<td>539000</td>
</tr>
<tr>
<td></td>
<td>N2: 1852000</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>Ar: 18000</td>
<td>14000</td>
</tr>
</tbody>
</table>

## Design Data:

- Materials of Construction: Stainless Steel
- Stages: 30
- Pressure Drop: 0.54 psi
- Diameter: 25.5 ft
- Height: 37.6 ft
- Packing Type: Structured
- Packing: Norton Intalox
- Theoretical Plate Height: 1.25 ft

**Comments:** T-101A is a low pressure, packed distillation column used to do a final separation of N2 and O2 to get a 98mol% O2 product. It uses a structured metal packing to achieve separation effective separation with very little pressure drop. The oxygen product is then sent to P-100A to raise the pressure injection to the gasifier.
**Valve**

**Identification:** Item Name: Pressure Valve  
Item Number: PV-100A  
Date: 5/3/2012

**Function:** To separate high pressure and low pressure processes

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Quantity (lb/hr)</th>
<th>O2</th>
<th>N2</th>
<th>Ar</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-109A</td>
<td>1129000</td>
<td>498000</td>
<td>609000</td>
<td>22000</td>
</tr>
</tbody>
</table>

**Temp In (F):** -290.2

**Temp Out (F):** -307.4

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Pressure Drop:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Diaphram</td>
<td>61.4 psi</td>
</tr>
</tbody>
</table>

**Comments:** SV-100A separates the high pressure of T-100A from the low pressure of T-101A. The stream going through this valve contains the crude O2 product from the bottoms of T-100A and will be fed into the middle of T-101A.
### Valve Specifications

**Identification:** | **Item Name:** | **Pressure Valve** | **Item Number:** | **PV-101A** | **Date:** | **5/3/2012**
---|---|---|---|---|---|

**Function:** | To separate high pressure and low pressure processes

**Operation:** | Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Composition:</th>
<th>Quantity (lb/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-112A</td>
<td>O2</td>
<td>1200</td>
</tr>
<tr>
<td></td>
<td>N2</td>
<td>1040000</td>
</tr>
<tr>
<td></td>
<td>Ar</td>
<td>6200</td>
</tr>
<tr>
<td></td>
<td></td>
<td>1047000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-112A</td>
<td>O2</td>
</tr>
<tr>
<td></td>
<td>N2</td>
</tr>
<tr>
<td></td>
<td>Ar</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Temp In (F):** | -306.4 |
**Temp Out (F):** | -314.1 |

**Design Data:**

<table>
<thead>
<tr>
<th>Pressure Drop:</th>
<th>60.8 psi</th>
</tr>
</thead>
</table>

**Type:** Diaphram

**Comments:** SV-101A separates the high pressure of T-100A from the low pressure of T-101A. The stream going through this valve contains the overhead N2 product from T-100A which will be fed into the top of T-101A.
**Vessel**

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Item Number:</th>
<th>Date:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pre-Purification Vessel</td>
<td>V-100A</td>
<td>5/1/2012</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To remove CO2 and other trace impurities from compressed air stream

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Quantity (lb/hr)</th>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-105A</td>
<td>2452000</td>
<td>O2 567000, N2 1852000, Ar 32000, CO2 1200</td>
</tr>
<tr>
<td>S-106A</td>
<td>2451000</td>
<td></td>
</tr>
</tbody>
</table>

**Temperature (F):** 95.0

**Pressure (psi):** 91.2

**Design Data:**

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Type:</td>
<td>Horizontal Closed Pressure Vessel</td>
</tr>
<tr>
<td>Volume</td>
<td>1260 ft³</td>
</tr>
<tr>
<td>Diameter</td>
<td>6.6 ft</td>
</tr>
<tr>
<td>Height</td>
<td>39.0 ft</td>
</tr>
<tr>
<td>Material</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Molecular Sieves</td>
<td>49000 lb</td>
</tr>
<tr>
<td>Alumina</td>
<td>8800 lb</td>
</tr>
</tbody>
</table>

**Comments:** V-100A is a prepurification unit used for removing CO2 and other trace impurities from the compressed air stream before it enters the coldbox to prevent freezing. V-100A is cycled in conjunction with V-101A every 3 hours to regenerate the adsorbants using the product N2 stream from the ASU.
# Vessel

**Identification:** Pre-Purification Vessel  
**Item Number:** V-101A  
**Date:** 5/1/2012

**Function:** To remove CO2 and other trace impurities from compressed air stream  
**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-105A</td>
<td>567000</td>
<td>567000</td>
</tr>
<tr>
<td>S-106A</td>
<td>1852000</td>
<td>1852000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>2452000</td>
<td>2451000</td>
</tr>
<tr>
<td>N2</td>
<td>567000</td>
<td>567000</td>
</tr>
<tr>
<td>Ar</td>
<td>1852000</td>
<td>1852000</td>
</tr>
<tr>
<td>CO2</td>
<td>32000</td>
<td>32000</td>
</tr>
</tbody>
</table>

**Temperature (F):** 95.0  
**Pressure (psi):** 91.2

**Design Data:**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Type</td>
<td>Horizontal Closed Pressure Vessel</td>
</tr>
<tr>
<td>Volume</td>
<td>1260 ft³</td>
</tr>
<tr>
<td>Diameter</td>
<td>6.6 ft</td>
</tr>
<tr>
<td>Height</td>
<td>39.0 ft</td>
</tr>
<tr>
<td>Material</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Molecular Sieves</td>
<td>49000 lb</td>
</tr>
<tr>
<td>Alumina</td>
<td>8800 lb</td>
</tr>
</tbody>
</table>

**Comments:** V-101A is a prepurification unit used for removing CO2 and other trace impurities from the compressed air stream before it enters the coldbox to prevent freezing. V-101A is cycled in conjunction with V-100A every 3 hours to regenerate the adsorbants using the product N2 stream from the ASU.
## Radiant Syngas Cooler

<table>
<thead>
<tr>
<th><strong>Identification:</strong></th>
<th><strong>Item Name:</strong> Radiant Syngas Cooler</th>
<th><strong>Date:</strong> 4/18/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Item Number:</strong></td>
<td>E-100G</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot syngas and water to make steam

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-105G</td>
<td>794000</td>
<td>1130000</td>
<td>1130000</td>
<td>794000</td>
</tr>
</tbody>
</table>

**Composition:**

<table>
<thead>
<tr>
<th>Component</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO2</td>
<td>0</td>
<td>294000</td>
<td>294000</td>
<td>0</td>
</tr>
<tr>
<td>CO</td>
<td>0</td>
<td>768000</td>
<td>768000</td>
<td>0</td>
</tr>
<tr>
<td>H2S</td>
<td>0</td>
<td>6000</td>
<td>6000</td>
<td>0</td>
</tr>
<tr>
<td>H2</td>
<td>0</td>
<td>38000</td>
<td>38000</td>
<td>0</td>
</tr>
<tr>
<td>H2O</td>
<td>794000</td>
<td>20000</td>
<td>20000</td>
<td>794000</td>
</tr>
</tbody>
</table>

**T\text{Hot} In (F):** 2550

**T\text{Hot} Out (F):** 1400

**T\text{Cold} In (F):** 86

**T\text{Cold} Out (F):** 930

**Pressure (psi):** 580.2

**Design Data:**

- **Type:** Vertical Tube Lined Pressure Vessel
- **Materials of Construction:** ZICHROM 90, Haynes HR-160 Alloy, Croloy, Carbon Steel
- **Heat Duty:** 566900000 BTU/hr
- **Tube Diameter**
  - OD: 2.5 in
  - Wall Division Spacing: 36 in
  - Heat Transfer Coefficient: 25 BTU/hr-F-ft^2
  - Heat Transfer Area: 17000 ft^2
  - Height: 139 ft
  - Diameter: 25.9 ft
  - Weight: 1630000 lb

**Comments:** E-100G is a radiant syngas cooler that exchanges heat between water and syngas. The vertical tube lined pressure vessel is positioned directly below the gasifier and cools syngas from 2550 F to 1382 F. US Patent No. US7,587,995 B2 provides more detailed schematics of design.
<table>
<thead>
<tr>
<th>Identification</th>
<th>Item Name: Heat Exchanger</th>
<th>Date: 4/18/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td>E-101G</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between clean syngas and water.

**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-108G</td>
<td>794000</td>
<td>1130000</td>
<td>1130000</td>
<td>794000</td>
</tr>
</tbody>
</table>

**Composition:**

- CO: 0
- CO2: 768000
- H2: 294000
- H2O: 794000
- H2S: 6000

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
<th>794000</th>
<th>1130000</th>
<th>1130000</th>
<th>794000</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hot In</td>
<td>768000</td>
<td>768000</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Hot Out</td>
<td>294000</td>
<td>294000</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Cold In</td>
<td>38000</td>
<td>38000</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Cold Out</td>
<td>19815.66</td>
<td>19815.66</td>
<td>794000</td>
<td></td>
</tr>
<tr>
<td></td>
<td>6000</td>
<td>6000</td>
<td>0</td>
<td></td>
</tr>
</tbody>
</table>

**T**

- **Hot In (F):** 2550
- **Hot Out (F):** 95.0
- **Cold In (F):** 86.0
- **Cold Out (F):** 932.0

**Pressure (psi):** 580.2

### Design Data:

- **Type:** Shell and Tube, One Pass, Triangle Pitch
- **Materials of Construction:** Tubes: Inconel 617 Shell: Carbon Steel
- **Heat Duty:** 526000000 BTU/h
- **Heat Transfer Coefficient:** 244 BTU/hr-ft²K
- **Number of Tubes:** 1100
- **Triangle Pitch:** 1 in
- **Heat Transfer Area:** 8900 ft²

**Comments:** E-101G is the low temperature syngas cooler positioned in between radiant syngas cooler E-100G and the H2S removal columns. It cools syngas from 2550 F to 95 F in order to make H2S and CO2 separation realizable.
**Gasification Unit Equipment Specification Sheets**

### Pump & Motor

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>P-100G</th>
</tr>
</thead>
<tbody>
<tr>
<td>Date:</td>
<td></td>
<td>4/18/2012</td>
</tr>
</tbody>
</table>

**Function:** To pump coal slurry from the wet rod mill to the gasifier

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-101T</td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-102T</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>894000</td>
<td>894000</td>
</tr>
<tr>
<td>Composition:</td>
<td></td>
<td></td>
</tr>
<tr>
<td>COAL</td>
<td>592000</td>
<td>592000</td>
</tr>
<tr>
<td>H2O</td>
<td>302000</td>
<td>302000</td>
</tr>
</tbody>
</table>

**Temp (F):** 518

**P_{in} (psi):** 14.7

**P_{out} (psi):**

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Horizontal Split Case Centrifugal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>11000 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>500 psi</td>
</tr>
<tr>
<td>Power Required:</td>
<td>506 hp</td>
</tr>
<tr>
<td>Efficiency:</td>
<td>0.79</td>
</tr>
<tr>
<td>Materials of Construction:</td>
<td>Stainless Steel</td>
</tr>
</tbody>
</table>

**Comments:** P-100G is a stainless steel horizontal split case pump that pressurized the coal slurry from the wet rod mill from 14.7 psi to 518 psi to be introduced into the gasifier. Due to the highly corrosive nature of the coal slurry, the pump motor is a totally enclosed, fan cooled enclosure.
**Gasifier**

<table>
<thead>
<tr>
<th><strong>Identification:</strong></th>
<th><strong>Item Name:</strong></th>
<th>Vertical Reactor</th>
<th><strong>Date:</strong></th>
<th>4/18/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Item Number:</strong></td>
<td>R-100G</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To react coal water and oxygen to produce syngas

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-102G</td>
<td>1434000</td>
<td></td>
</tr>
<tr>
<td>S-103G</td>
<td>1434000</td>
<td></td>
</tr>
</tbody>
</table>

**Composition:**

- **COAL:** 592000 lb/hr, 15015.6 lb/hr
- **H2O:** 302000 lb/hr, 19000 lb/hr
- **OXYGEN:** 539000 lb/hr, 0 lb/hr
- **ARGON:** 1700 lb/hr, 1700 lb/hr
- **CO:** 0 lb/hr, 749000 lb/hr
- **CO2:** 0 lb/hr, 279000 lb/hr
- **NH3:** 0 lb/hr, 4200 lb/hr
- **H2S:** 0 lb/hr, 6000 lb/hr
- **H2:** 0 lb/hr, 35000 lb/hr
- **ASH:** 0 lb/hr, 326000 lb/hr

**Temp (F):** 2552

**Pressure (psi):** 580.2

**Design Data:**

- **Type:** PFR
- **Residence Time:** 5.0 s
- **Materials of Construction:** ZIRCHROM 90, ZIRCHROM 60, CHROMCHOR 12, Carbon Steel
- **Height:** 32.4 ft
- **Diameter:** 16.4 ft
- **Volume:** 1800 ft³

**Comments:** R-100G is a gasifier vertical reaction vessel that receives a 65 wt% coal slurry and 98mol% Oxygen feed at 212F and 1000 psi. The gasifier operates at a pressure of 580.2 psi and 2552 F. The coal coal undergoes pyrolysis and multiple gasification reactions to produce syngas.
# Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Heat Exchanger</td>
<td>5/8/2012</td>
<td>E-100C-A</td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Hot in</th>
<th>Cold In</th>
<th>Cold Out</th>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-108C</td>
<td></td>
<td>18652000</td>
<td>18652000</td>
<td>0</td>
</tr>
<tr>
<td>S-106C</td>
<td>2.47E+06</td>
<td></td>
<td>2.01E+07</td>
<td>2.47E+06</td>
</tr>
<tr>
<td>S-107C</td>
<td>2.01E+07</td>
<td></td>
<td>2.01E+07</td>
<td>2.47E+06</td>
</tr>
<tr>
<td>S-109C</td>
<td>2.47E+06</td>
<td>2.01E+07</td>
<td>2.47E+06</td>
<td>2.47E+06</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
<th></th>
<th>2.47E+06</th>
<th>2.01E+07</th>
<th>2.01E+07</th>
<th>2.47E+06</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>0</td>
<td>18652000</td>
<td>18652000</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>CO2</td>
<td>0</td>
<td>1382000</td>
<td>1382000</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>CO</td>
<td>0</td>
<td>9000</td>
<td>9000</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>H2S</td>
<td>0</td>
<td>1200</td>
<td>1200</td>
<td>0</td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>2.47E+06</td>
<td>2200</td>
<td>2200</td>
<td>2.47E+06</td>
<td></td>
</tr>
<tr>
<td>H2</td>
<td>2.47E+06</td>
<td>2100</td>
<td>2100</td>
<td>0</td>
<td></td>
</tr>
</tbody>
</table>

**Temperature (F):**
- $T_{\text{Hot In}}$: 419.0
- $T_{\text{Hot Out}}$: 318.1
- $T_{\text{Cold In}}$: 115.7
- $T_{\text{Cold Out}}$: 392.0

**Pressure (psi):** 1160

## Design Data:

- **Type:** Fixed Head Shell and Tube
- **Materials of Construction:** Carbon Steel Shell/Monel Tube
- **Heat Duty:** 2.15E+09 Btu/hr
- **Shell Diameter:** 37 in
- **Heat Transfer Coefficient:** 16434 Btu/(F-ft^2-hr)
- **Heat Transfer Area:** 1267 ft^2
**Heat Exchanger**

**Identification:** Heat Exchanger  
**Item Name:** Heat Exchanger  
**Item Number:** E-100C-B  
**Date:** 5/8/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Hot in</th>
<th>Cold In</th>
<th>Cold Out</th>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-111C</td>
<td>S-107C</td>
<td>S-110C</td>
<td>S-112C</td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>6.27E+05</td>
<td>1.03E+05</td>
<td>1.03E+05</td>
<td>6.27E+05</td>
</tr>
</tbody>
</table>

**Composition:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>SELEXOL</th>
<th>CO2</th>
<th>CO</th>
<th>H2S</th>
<th>Water</th>
<th>H2</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>95000</td>
<td>7000</td>
<td>48</td>
<td>6</td>
<td>11</td>
<td>11</td>
</tr>
<tr>
<td></td>
<td>95000</td>
<td>7000</td>
<td>48</td>
<td>6</td>
<td>11</td>
<td>11</td>
</tr>
<tr>
<td></td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
</tbody>
</table>

**T黄石 In (F):** 572.0  
**T黄石 Out (F):** 419.0  
**T冷 In (F):** 392.0  
**T冷 Out (F):** 482.0  
**Pressure (psi):** 1160

**Design Data:**

- **Type:** Fixed Head Shell and Tube  
- **Materials of Construction:** Carbon Steel Shell/Monel Tube  
- **Heat Duty:** 8.35E+08 Btu/hr  
- **Shell Diameter:** 37 in  
- **Heat Transfer Coefficient:** 1753 Btu/(F-ft^2-hr)  
- **Heat Transfer Area:** 870.8 ft^2

**Comments:** E-100C-B is used to further heat the carbon dioxide rich solvent stream coming out of the absorption column (T-100C) in order to increase the vapor pressure of the carbon dioxide where it can then be flashed out of solution.
CO\textsubscript{2} Sequestration Unit Equipment Specification Sheets

# Heat Exchanger

**Identification:** Heat Exchanger
**Item Name:** Heat Exchanger
**Item Number:** E-101C-A
**Date:** 5/8/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>SELEXOL</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>N2</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>CO</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>H2S</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td></td>
<td>Water</td>
<td>2.47E+06</td>
<td>400</td>
<td>400</td>
</tr>
<tr>
<td></td>
<td>H2</td>
<td>0</td>
<td>3000</td>
<td>3000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.47E+06</td>
</tr>
<tr>
<td>1.69E+06</td>
</tr>
<tr>
<td>1.69E+06</td>
</tr>
<tr>
<td>2.47E+06</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
</tr>
<tr>
<td>N2</td>
</tr>
<tr>
<td>CO</td>
</tr>
<tr>
<td>H2S</td>
</tr>
<tr>
<td>Water</td>
</tr>
<tr>
<td>H2</td>
</tr>
</tbody>
</table>

| T\text{Hot In (F)}: | 467.4  |
| T\text{Hot Out (F)}: | 338.0  |
| T\text{Cold In (F)}: | 318.1  |
| T\text{Cold Out (F)}: | 318.1  |

| Pressure (psi): | 72.52  |

## Design Data:

<table>
<thead>
<tr>
<th>Type:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fixed Head Shell and</td>
</tr>
<tr>
<td>Tube</td>
</tr>
<tr>
<td>Materials of Construction:</td>
</tr>
<tr>
<td>Heat Duty:</td>
</tr>
<tr>
<td>Shell Diameter:</td>
</tr>
<tr>
<td>Heat Transfer Coefficient:</td>
</tr>
<tr>
<td>Heat Transfer Area:</td>
</tr>
</tbody>
</table>

**Comments:** E-101C-A is used to cool the carbon dioxide coming out of the flash tanks in preparation for compression and sequestration.
# Heat Exchanger

**Identification:** Item Name: Heat Exchanger  
**Item Number:** E-101C-B  
**Date:** 5/8/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-132C</td>
<td>0</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Hot In</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-130C</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-134C</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-133C</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>7.12E+04</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
</tr>
<tr>
<td>CO2</td>
</tr>
<tr>
<td>CO</td>
</tr>
<tr>
<td>H2S</td>
</tr>
<tr>
<td>Water</td>
</tr>
<tr>
<td>H2</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>1315000</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>9300</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>1100</td>
</tr>
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</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>7.12E+04</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>400</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>400</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>7.12E+04</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>3000</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>3000</td>
</tr>
</tbody>
</table>

<p>| |</p>
<table>
<thead>
<tr>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
</tr>
</tbody>
</table>

**T\text{Hot In (F)}:** 338  
**T\text{Hot Out (F)}:** 185.0  
**T\text{Cold In (F)}:** 86.0  
**T\text{Cold Out (F)}:** 320.0  
**Pressure (psi):** 1160

**Design Data:**

- **Type:** Fixed Head Shell and Tube  
- **Materials of Construction:** Carbon Steel Shell/Monel Tube  
- **Heat Duty:** 7.98E+07 Btu/hr  
- **Shell Diameter:** 37 in  
- **Heat Transfer Coefficient:** 4017 Btu/(F-ft^2-hr)  
- **Heat Transfer Area:** 697 ft^2

**Comments:** E-101C-B is used to further cool the carbon dioxide coming out of the flash tanks in preparation for compression and sequestration.
# Heat Exchanger

**Identification:** Heat Exchanger  
**Item Name:** Heat Exchanger  
**Item Number:** E-102C  
**Date:** 5/8/2012

## Function:
To exchange energy between hot and cold streams.

## Operation:
Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Quantity (lb/hr)</th>
<th>Composition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cold in</td>
<td>1.11E+06</td>
<td>SELEXOL</td>
</tr>
<tr>
<td>Hot In</td>
<td>8.28E+06</td>
<td>CO2</td>
</tr>
<tr>
<td>Hot Out</td>
<td>8.28E+06</td>
<td>CO</td>
</tr>
<tr>
<td>Cold Out</td>
<td>1.11E+06</td>
<td>H2S</td>
</tr>
</tbody>
</table>

| Cold in    | 1.11E+06        | Water       |
| Hot In     | 8.28E+06        | H2          |
| Hot Out    | 8.28E+06        |             |
| Cold Out   | 1.11E+06        |             |

- **T\_Hot In (F):** 484.5  
- **T\_Hot Out (F):** 419.0  
- **T\_Cold In (F):** 95.0  
- **T\_Cold Out (F):** 86.0  
- **Pressure (psi):** 1160

## Design Data:
- **Type:** Fixed Head Shell and Tube  
- **Materials of Construction:** Carbon Steel Shell/Monel Tube  
- **Heat Duty:** 1.30E+09 Btu/hr  
- **Shell Diameter:** 37 in  
- **Heat Transfer Coefficient:** 7156 Btu/(F-ft\(^2\)-hr)  
- **Heat Transfer Area:** 650 ft\(^2\)

## Comments:
E-102C is used to cool the regenerated lean glycol stream going back into the absorption column (T-100C) as a better separation is obtained in the column when the solvent is at a lower temperature.
# Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Name:</td>
<td>Heat Exchanger</td>
<td>5/8/2012</td>
</tr>
<tr>
<td>Item Number:</td>
<td>E-103C</td>
<td></td>
</tr>
</tbody>
</table>

## Function:
To exchange energy between hot and cold streams.

## Operation:
Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-128C</td>
<td>1.35E+06</td>
<td>1.01E+07</td>
<td>1.01E+07</td>
<td>1.35E+06</td>
</tr>
<tr>
<td>S-123C</td>
<td>0</td>
<td>150000</td>
<td>150000</td>
<td>0</td>
</tr>
<tr>
<td>S-102C</td>
<td>0</td>
<td>151</td>
<td>151</td>
<td>0</td>
</tr>
<tr>
<td>S-127C</td>
<td>0</td>
<td>473</td>
<td>473</td>
<td>0</td>
</tr>
</tbody>
</table>

### Composition:

- **SELEXOL**: 0, 9919000, 9919000, 0
- **CO2**: 0, 150000, 150000, 0
- **CO**: 0, 151, 151, 0
- **H2S**: 0, 473, 473, 0
- **Water**: 1.35E+06, 0, 1, 1.35E+06
- **H2**: 0, 222, 222, 0

### Temperature:
- **T_{Hot In} (F)**: 484.5
- **T_{Hot Out} (F)**: 419.0
- **T_{Cold In} (F)**: 95.0
- **T_{Cold Out} (F)**: 86.0

### Pressure (psi):
- 1160

### Design Data:

- **Type**: Fixed Head Shell and Tube
- **Materials of Construction**: Carbon Steel Shell/Monel Tube
- **Heat Duty**: 1.58E+09 Btu/hr
- **Shell Diameter**: 37 in
- **Heat Transfer Coefficient**: 8795 Btu/(F-ft^2-hr)
- **Heat Transfer Area**: 195 ft^2

### Comments:
E-102C is used to cool the regenerated lean glycol stream going back into the absorption column (T-100C) as a better separation is obtained in the column when the solvent is at a lower temperature.
## CO2 Pump & Motor

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Pump</td>
<td>5/8/2012</td>
<td>P-100C</td>
</tr>
</tbody>
</table>

### Function:
To pump from Solvent Dehydrator to Absorption Column

### Operation:
Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-120C</td>
<td>8.28E+06</td>
<td>8.28E+06</td>
</tr>
<tr>
<td>S-122C</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

| Quantity (lb/hr) | 8.28E+06 | 8.28E+06 |
| Composition (lb/hr): |
| SELEXOL        | 8.24E+06 | 8.24E+06 |
| CO2           | 43000    | 43000    |
| CO            | 55       | 55       |
| H2            | 9        | 9        |
| WATER        | 0        | 0        |

### Temperature (F):

<table>
<thead>
<tr>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>474.4</td>
<td>474.4</td>
</tr>
</tbody>
</table>

### Pressure (psi):

<table>
<thead>
<tr>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>144.6</td>
<td>144.6</td>
</tr>
</tbody>
</table>

### Design Data:

| Type:      | Centrifugal |
| Volumetric Flow Rate: | 128900 ft³/hr |
| Pressure Change: | 1016 psi |
| Power Required: | 11.93 MW |
| Efficiency: | 0.75 |

### Materials of Construction: Inconel

### Comments:
Solvent recycle pump in order to raise the pressure or recycled Selexol to put back into the absorber.
<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO2 Pump &amp; Motor</td>
<td>Pump</td>
<td>5/8/2012</td>
<td>P-101C</td>
</tr>
</tbody>
</table>

**Function:**
To pump from Solvent Dehydrator to Absorption Column

**Operation:**
Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Quantity (lb/hr)</th>
<th>Composition (lb/hr):</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-121C</td>
<td>1.01E+07</td>
<td>SELEXOL 1.00E+07 CO2 25000 CO 15 H2 2 WATER 0</td>
</tr>
<tr>
<td>S-123C</td>
<td>1.01E+07</td>
<td>SELEXOL 1.00E+07 CO2 25000 CO 15 H2 2 WATER 0</td>
</tr>
</tbody>
</table>

**Temperature (F):**

<table>
<thead>
<tr>
<th>Temperature (F):</th>
<th>Inlet</th>
<th>Out</th>
<th>Design Data:</th>
</tr>
</thead>
<tbody>
<tr>
<td>473.6</td>
<td>473.6</td>
<td>473.6</td>
<td></td>
</tr>
<tr>
<td>71.65</td>
<td>71.65</td>
<td>71.65</td>
<td></td>
</tr>
</tbody>
</table>

**Type:**
Centrifugal

**Volumetric Flow Rate:**
156600 ft³/hr

**Pressure Change:**
1089 psi

**Power Required:**
15.63 MW

**Efficiencys:**
0.75

**Materials of Construction:**
Inconel

**Comments:**
Solvent recycle pump in order to raise the pressure or recycled Selexol to put back into the absorber.
## CO2 Absorption Column

<table>
<thead>
<tr>
<th><strong>Identification:</strong></th>
<th><strong>Item Name:</strong></th>
<th><strong>Date:</strong></th>
<th><strong>Item Number:</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Absorption Column</td>
<td>5/8/2012</td>
<td>T-100C</td>
</tr>
</tbody>
</table>

**Function:** To separate chemical Carbon dioxide from Shifted Gas

**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Quantity (lb/hr)</th>
<th>Composition (lb/hr):</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed</td>
<td>Overhead</td>
<td>Bottoms</td>
</tr>
<tr>
<td>S-100C</td>
<td>S-105C</td>
<td>S-106C</td>
</tr>
<tr>
<td>1.54E+06</td>
<td>2.09E+05</td>
<td>2.01E+07</td>
</tr>
<tr>
<td>SELEXOL</td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.36E+06</td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO</td>
<td></td>
<td></td>
</tr>
<tr>
<td>86000</td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>88000</td>
<td></td>
<td></td>
</tr>
<tr>
<td>WATER</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1900</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Quantity (lb/hr)</th>
<th>Composition (lb/hr):</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liq Feed(2)</td>
<td>Liq Feed(1)</td>
<td>Recycle</td>
</tr>
<tr>
<td>S-102C</td>
<td>S-101C</td>
<td>S-104C</td>
</tr>
<tr>
<td>8.35E+06</td>
<td>1.00E+07</td>
<td>3.63E+06</td>
</tr>
<tr>
<td>SELEXOL</td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>9.99E+06</td>
<td>8.31E+06</td>
<td>37000</td>
</tr>
<tr>
<td>CO</td>
<td></td>
<td></td>
</tr>
<tr>
<td>25000</td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2</td>
<td></td>
<td></td>
</tr>
<tr>
<td>WATER</td>
<td></td>
<td></td>
</tr>
<tr>
<td>0</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Temp Top (F):** 95.6

**Temp Bott (F):** 115.7

### Design Data:

- **Materials of Construction:** Inconel
- **Stages:** 12
- **Pressure Drop:** 11 psi
- **Stage Pressure Drop:** 0.92 psi
- **Diameter:** 26 ft
- **Height:** 65 ft
- **Tray Spacing:** 2 ft
- **Tray Type:** Sieve
- **Wall Thickness:** 4.51 in

**Comments:** This absorption column (T-100C) is used in the removal of CO2 using a poly-ethylene glycol solvent called Selexol. There are multiple glycol feeds with varying levels of solvent purity at various stages.
## CO2 Flash Drum

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Vertical Flash Drum</td>
<td>5/8/2012</td>
<td>V-100C</td>
</tr>
</tbody>
</table>

**Function:** To separate CO2 from Selexol  
**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-110C</td>
<td>2.01E+07</td>
<td>1.50E+06</td>
<td>1.86E+07</td>
</tr>
<tr>
<td>S-112C</td>
<td>1.87E+07</td>
<td>1.87E+07</td>
<td>1.84E+07</td>
</tr>
<tr>
<td>S-113C</td>
<td>1.38E+06</td>
<td>1.19E+06</td>
<td>1.92E+06</td>
</tr>
</tbody>
</table>

**Composition (lb/hr):**

- **SELEXOL:** 1.87E+07, 1.87E+07, 1.84E+07
- **CO2:** 1.38E+06, 1.19E+06, 1.92E+06
- **CO:** 9400, 8900, 500
- **H2:** 3000, 3000, 100
- **WATER:** 2000, 400, 1800

**Temp (F):** 475.6, 475.6, 475.6  
**Pressure (psi):** 290.1, 290.1, 290.1

### Design Data:

- **Materials of Construction:** Inconel  
- **Diameter:** 30 ft  
- **Height:** 105 ft  
- **Shell Thickness:** 4.03 in

**Comments:** Solvent regeneration consisting of simply a flash drum. The temperature and pressure are set by preheating heat exchangers and valves.
### CO2 Flash Drum

**Identification:** Vertical Flash<br>
**Item Name:** Drum<br>
**Item Number:** V-101C<br>
**Date:** 5/8/2012

**Function:** To separate CO2 from Selexol

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-113C</td>
<td>S-114C</td>
<td>S-116C</td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>1.86E+07</td>
<td>1.50E+06</td>
<td>1.86E+07</td>
</tr>
<tr>
<td>Composition (lb/hr):</td>
<td>SELEXOL</td>
<td>CO2</td>
<td>CO</td>
</tr>
<tr>
<td></td>
<td>1.84E+07</td>
<td>1.92E+06</td>
<td>500</td>
</tr>
<tr>
<td></td>
<td>4.18E+04</td>
<td>9.69E+04</td>
<td>400</td>
</tr>
<tr>
<td></td>
<td>1.83E+07</td>
<td>9.55E+04</td>
<td>100</td>
</tr>
</tbody>
</table>

**Temp (F):** 474.5 | 474.5 | 474.5 | 474.5

**Pressure (psi):** 145 | 145 | 145 | 145

**Design Data:**

- **Materials of Construction:** Inconel
- **Diameter:** 21 ft
- **Height:** 115.5 ft
- **Shell Thickness:** 1.58 in

**Comments:** Solvent regeneration consisting of simply a flash drum. The temperature and pressure are set by preheating heat exchangers and valves.
## CO₂ Flash Drum

<table>
<thead>
<tr>
<th><strong>Identification:</strong></th>
<th><strong>Item Name:</strong></th>
<th><strong>Date:</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Vertical Flash</td>
<td>5/8/2012</td>
</tr>
<tr>
<td><strong>Item Number:</strong></td>
<td>V-102C</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To separate CO₂ from Selexol

**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-116C</td>
<td>1.86E+07</td>
<td>5.10E+04</td>
<td>1.01E+07</td>
</tr>
</tbody>
</table>

### Composition (lb/hr):

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>SELEXOL</th>
<th>CO2</th>
<th>CO</th>
<th>H2</th>
<th>WATER</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-116C</td>
<td>1.83E+07</td>
<td>9.55E+04</td>
<td>100</td>
<td>19</td>
<td>1750</td>
</tr>
<tr>
<td>S-118C</td>
<td>2.32E+04</td>
<td>2.77E+04</td>
<td>52</td>
<td>8</td>
<td>29</td>
</tr>
<tr>
<td>S-119C</td>
<td>1.01E+07</td>
<td>2.48E+04</td>
<td>16</td>
<td>2</td>
<td>900</td>
</tr>
</tbody>
</table>

### Temp (F):

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-116C</td>
<td>473.1</td>
<td>473.1</td>
<td>473.1</td>
</tr>
<tr>
<td>S-118C</td>
<td>473.1</td>
<td>473.1</td>
<td>473.1</td>
</tr>
<tr>
<td>S-119C</td>
<td>473.1</td>
<td>473.1</td>
<td>473.1</td>
</tr>
</tbody>
</table>

### Pressure (psi):

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-116C</td>
<td>72.5</td>
<td>72.5</td>
<td>72.5</td>
</tr>
<tr>
<td>S-118C</td>
<td>72.5</td>
<td>72.5</td>
<td>72.5</td>
</tr>
<tr>
<td>S-119C</td>
<td>72.5</td>
<td>72.5</td>
<td>72.5</td>
</tr>
</tbody>
</table>

### Design Data:

<table>
<thead>
<tr>
<th>Materials of Construction:</th>
<th>Inconel</th>
</tr>
</thead>
<tbody>
<tr>
<td>Diameter</td>
<td>17 ft</td>
</tr>
<tr>
<td>Height</td>
<td>93 ft</td>
</tr>
<tr>
<td>Shell Thickness</td>
<td>0.70 in</td>
</tr>
</tbody>
</table>

**Comments:** Solvent regeneration consisting of simply a flash drum. The temperature and pressure are set by preheating heat exchangers and valves.
**Heat Exchanger**

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Heat Exchanger</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td></td>
<td>E-101S</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Hot In</th>
<th>Cold In</th>
<th>Cold Out</th>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-127S</td>
<td>1.13E+06</td>
<td>7.94E+05</td>
<td>7.94E+05</td>
<td>1.13E+06</td>
</tr>
</tbody>
</table>

**Composition:**

<table>
<thead>
<tr>
<th></th>
<th>SELEXOL</th>
<th>CO2</th>
<th>CO</th>
<th>H2S</th>
<th>Water</th>
<th>H2</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0</td>
<td>2.94E+05</td>
<td>7.68E+05</td>
<td>6000</td>
<td>20000</td>
<td>38000</td>
</tr>
<tr>
<td></td>
<td>739000</td>
<td>55000</td>
<td>400</td>
<td>50</td>
<td>90</td>
<td>80</td>
</tr>
<tr>
<td></td>
<td>739000</td>
<td>55000</td>
<td>400</td>
<td>50</td>
<td>90</td>
<td>80</td>
</tr>
<tr>
<td></td>
<td>0</td>
<td>2.94E+05</td>
<td>7.68E+05</td>
<td>6000</td>
<td>20000</td>
<td>38000</td>
</tr>
</tbody>
</table>

**T\text{Hot In} (F):** 2552.0

**T\text{Hot Out} (F):** 95.0

**T\text{Cold In} (F):** 86.0

**T\text{Cold Out} (F):** 932.0

**Pressure (psi):** 580.2

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Fixed Head Shell and Tube</th>
</tr>
</thead>
<tbody>
<tr>
<td>Materials of Construction:</td>
<td>Carbon Steel Shell/Monel Tube</td>
</tr>
<tr>
<td>Heat Duty:</td>
<td>1.12E+09 Btu/hr</td>
</tr>
<tr>
<td>Shell Diameter:</td>
<td>37 in</td>
</tr>
<tr>
<td>Heat Transfer Coefficient:</td>
<td>240 Btu/(F-ft^2-hr)</td>
</tr>
<tr>
<td>Heat Transfer Area:</td>
<td>57500 ft^2</td>
</tr>
</tbody>
</table>

**Comments:** E-101S is used to cool the high temperature stream coming out of the gasifier so that a better separation can be obtained in the desulfurization unit.
# Heat Exchanger

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Heat Exchanger</td>
<td>5/8/2012</td>
</tr>
<tr>
<td>Item Number:</td>
<td>E-102S-A</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Hot In</th>
<th>Cold In</th>
<th>Cold Out</th>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>S-113S</td>
<td>S-105S</td>
<td>S-106S</td>
<td>S-114S</td>
</tr>
<tr>
<td></td>
<td>1.22E+07</td>
<td>1.26E+07</td>
<td>1.26E+07</td>
<td>1.22E+07</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Hot In</th>
<th>Cold In</th>
<th>Cold Out</th>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>1.21E+07</td>
<td>1.24E+07</td>
<td>1.24E+07</td>
<td>1.21E+07</td>
</tr>
<tr>
<td>CO2</td>
<td>19000</td>
<td>1.51E+05</td>
<td>1.51E+05</td>
<td>19000</td>
</tr>
<tr>
<td>CO</td>
<td>900</td>
<td>27000</td>
<td>27000</td>
<td>900</td>
</tr>
<tr>
<td>H2S</td>
<td>1500</td>
<td>6800</td>
<td>6800</td>
<td>1500</td>
</tr>
<tr>
<td>Water</td>
<td>20000</td>
<td>24000</td>
<td>24000</td>
<td>20000</td>
</tr>
<tr>
<td>H2</td>
<td>8</td>
<td>400</td>
<td>400</td>
<td>8</td>
</tr>
</tbody>
</table>

| \( T_{\text{Hot In}} \) (F): | 572.4 |
| \( T_{\text{Hot Out}} \) (F): | 122.0 |
| \( T_{\text{Cold In}} \) (F): | 98.8  |
| \( T_{\text{Cold Out}} \) (F) | 544.5 |

| Pressure (psi): | 580.2 |

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Fixed Head Shell and Tube</th>
</tr>
</thead>
<tbody>
<tr>
<td>Materials of Construction:</td>
<td>Carbon Steel Shell/Monel Tube</td>
</tr>
<tr>
<td>Heat Duty:</td>
<td>2.35E+09 Btu/hr</td>
</tr>
<tr>
<td>Shell Diameter:</td>
<td>37 in</td>
</tr>
<tr>
<td>Heat Transfer Coefficient:</td>
<td>320 Btu/(F-ft^2-hr)</td>
</tr>
<tr>
<td>Heat Transfer Area:</td>
<td>17400 ft^2</td>
</tr>
</tbody>
</table>

**Comments:** E-101S-A is used to heat the hydrogen sulfide rich solvent stream coming out of the absorption column (T-100S) in order to increase the vapor pressure of the hydrogen sulfide where it can then be flashed out of solution.
**Heat Exchanger**

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Date:</th>
<th>Item Number:</th>
<th>E-103S</th>
</tr>
</thead>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Hot in</th>
<th>Cold In</th>
<th>Cold Out</th>
<th>Hot Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-108C</td>
<td>3.90E+06</td>
<td>12129000</td>
<td>12129000</td>
<td>0</td>
</tr>
<tr>
<td>S-106C</td>
<td>1.22E+07</td>
<td>12129000</td>
<td>12129000</td>
<td>0</td>
</tr>
<tr>
<td>S-107C</td>
<td>1.22E+07</td>
<td>12129000</td>
<td>12129000</td>
<td>0</td>
</tr>
<tr>
<td>S-109C</td>
<td>3.90E+06</td>
<td>12129000</td>
<td>12129000</td>
<td>0</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>0</td>
<td>19000</td>
<td>19000</td>
<td>0</td>
</tr>
<tr>
<td>CO2</td>
<td>0</td>
<td>900</td>
<td>900</td>
<td>0</td>
</tr>
<tr>
<td>CO</td>
<td>0</td>
<td>1500</td>
<td>1500</td>
<td>0</td>
</tr>
<tr>
<td>H2S</td>
<td>0</td>
<td>20000</td>
<td>20000</td>
<td>3.90E+06</td>
</tr>
<tr>
<td>Water</td>
<td>3.90E+06</td>
<td>12</td>
<td>12</td>
<td>0</td>
</tr>
<tr>
<td>H2</td>
<td>0</td>
<td>12</td>
<td>12</td>
<td>0</td>
</tr>
</tbody>
</table>

\[ T_{\text{Hot In (F)}}: \] 122.0
\[ T_{\text{Hot Out (F)}}: \] 95.0
\[ T_{\text{Cold In (F)}}: \] 86.0
\[ T_{\text{Cold Out (F)}}: \] 113.0

**Pressure (psi):** 609.2

**Design Data:**

- **Type:** Fixed Head Shell and Tube
- **Materials of Construction:** Carbon Steel Shell/Carbon Steel Tube
- **Heat Duty:** 1.06E+08 Btu/hr
- **Shell Diameter:** 37 in
- **Heat Transfer Coefficient:** 123 Btu/(F-ft^2-hr)
- **Heat Transfer Area:** 18500 ft^2

**Comments:** E-100C-A is used to heat the carbon dioxide rich solvent stream coming out of the absorption column (T-100C) in order to increase the vapor pressure of the carbon dioxide where it can then be flashed out of solution.
**H2S Pump & Motor**

**Identification:** Item Name: Pump  
Date: 5/8/2012

**Item Number:** P-100S

**Function:** To pump from Solvent Dehydrator to Absorption Column

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-111S</td>
<td></td>
<td>S-113S</td>
</tr>
<tr>
<td>Quantity (lb/hr):</td>
<td>1.22E+07</td>
<td>1.22E+07</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition (lb/hr):</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>1.21E+07</td>
<td>1.21E+07</td>
</tr>
<tr>
<td>CO2</td>
<td>19000</td>
<td>19000</td>
</tr>
<tr>
<td>CO</td>
<td>900</td>
<td>900</td>
</tr>
<tr>
<td>H2</td>
<td>8</td>
<td>8</td>
</tr>
<tr>
<td>WATER</td>
<td>20000</td>
<td>20000</td>
</tr>
</tbody>
</table>

**Temperature (F):**  
Inlet: 567  
Out: 572.1

**Pressure (psi):**  
Inlet: 87  
Out: 609.2

**Design Data:**

- **Type:** Centrifugal
- **Volumetric Flow Rate:** 1.89E+05 ft³/hr
- **Pressure Change:** 522 psi
- **Power Required:** 9.6 MW
- **Efficiencies:** 0.75
- **Materials of Construction:** Inconel

**Comments:** Solvent recycle pump in order to raise the pressure or recycled selexol to put back into the absorber.
### H2S Pump & Motor

**Identification:**
- **Item Name:** Pump
- **Item Number:** P-101S
- **Date:** 5/8/2012

**Function:** To pump from Solvent Dehydrator to Absorption Column

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-123S</td>
<td>S-102S</td>
<td></td>
</tr>
<tr>
<td></td>
<td>1400</td>
<td>1400</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>1100</td>
<td>1100</td>
</tr>
<tr>
<td>CO2</td>
<td>57</td>
<td>57</td>
</tr>
<tr>
<td>CO</td>
<td>0.8</td>
<td>0.8</td>
</tr>
<tr>
<td>H2</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>WATER</td>
<td>250</td>
<td>250</td>
</tr>
</tbody>
</table>

**Temperature (F):**
- Inlet: 95
- Out: 101

**Pressure (psi):**
- Inlet: 87
- Out: 609

**Design Data:**
- **Type:** Centrifugal
- **Volumetric Flow Rate:** 4700 ft³/hr
- **Pressure Change:** 522 psi
- **Power Required:** 0.2 MW
- **Efficiency:** 0.75
- **Materials of Construction:** Inconel

**Comments:** Solvent recycle pump in order to raise the pressure of recovered selexol to put back into the absorber.
### H2S Absorption Column

<table>
<thead>
<tr>
<th><strong>Identification:</strong></th>
<th>Item Name:</th>
<th>Absorption Column</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Item Number:</strong></td>
<td>T-100S</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To separate chemical Sulfur hydroxide from Shifted Gas

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Feed</th>
<th>Recycle</th>
<th>Bottoms</th>
<th>Make Up</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-100S</td>
<td>1.13E+06</td>
<td>3.02E+05</td>
<td>1.26E+07</td>
<td>5.83</td>
</tr>
<tr>
<td>S-102S</td>
<td>2.94E+05</td>
<td>2499.9</td>
<td>1.51E+05</td>
<td>0</td>
</tr>
<tr>
<td>S-105S</td>
<td>7.68E+05</td>
<td>24</td>
<td>27000</td>
<td>0</td>
</tr>
<tr>
<td>S-101S</td>
<td>38000</td>
<td>0</td>
<td>3.86E+02</td>
<td>0</td>
</tr>
<tr>
<td>S-10S</td>
<td>6000</td>
<td>300</td>
<td>6800</td>
<td>0</td>
</tr>
<tr>
<td>S-101S</td>
<td>20000</td>
<td>4600</td>
<td>24000</td>
<td>0</td>
</tr>
</tbody>
</table>

**Composition (lb/hr):**

- **SELEXOL**
  - Feed: 0
  - Recycle: 2.95E+05
  - Bottoms: 1.24E+07
  - Make Up: 5.83

- **CO2**
  - Feed: 2.94E+05
  - Recycle: 2499.9
  - Bottoms: 1.51E+05
  - Make Up: 0

- **CO**
  - Feed: 7.68E+05
  - Recycle: 24
  - Bottoms: 27000
  - Make Up: 0

- **H2**
  - Feed: 38000
  - Recycle: 0
  - Bottoms: 3.86E+02
  - Make Up: 0

- **H2S**
  - Feed: 6000
  - Recycle: 300
  - Bottoms: 6800
  - Make Up: 0

- **WATER**
  - Feed: 20000
  - Recycle: 4600
  - Bottoms: 24000
  - Make Up: 0

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Overhead</th>
<th>Liq Feed(1)</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-104S</td>
<td>9.45E+05</td>
<td>1.22E+07</td>
</tr>
<tr>
<td>S-103S</td>
<td>30</td>
<td>1.21E+07</td>
</tr>
</tbody>
</table>

**Composition (lb/hr):**

- **SELEXOL**
  - Overhead: 30
  - Liq Feed(1): 1.21E+07

- **CO2**
  - Overhead: 1.64E+05
  - Liq Feed(1): 1.91E+04

- **CO**
  - Overhead: 7.42E+05
  - Liq Feed(1): 930

- **H2**
  - Overhead: 37000
  - Liq Feed(1): 8

- **WATER**
  - Overhead: 0
  - Liq Feed(1): 0

**Temp Top (F):** 96.91
**Temp Bot (F):** 98.78

**Design Data:**

- **Materials of Construction:** Inconel
- **Stages:** 20
- **Pressure Drop:** 12 psi
- **Stage Pressure Drop:** 0.4 psi
- **Diameter:** 18 ft
- **Height:** 75 ft
- **Tray Spacing:** 2 ft
- **Tray Type:** Sieve

**Comments:** This Absorption tower is used in the removal of H2S using a polyethylene glycol solvent called Selexol. There are multiple glycol feeds with varying levels of solvent purity at various stages.
# H2S Flash Drum

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Item Number:</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
</table>

**Function:** To separate H2S from Selexol  
**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S107-S</td>
<td>1.26E+07</td>
<td>1.87E+05</td>
<td>1.45E+07</td>
</tr>
<tr>
<td>S-108S</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-109S</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition (lb/hr):</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>1.24E+07</td>
<td>8.69E+04</td>
<td>1.23E+07</td>
</tr>
<tr>
<td>CO2</td>
<td>1.51E+05</td>
<td>7.63E+04</td>
<td>7.48E+04</td>
</tr>
<tr>
<td>CO</td>
<td>27000</td>
<td>19000</td>
<td>7700</td>
</tr>
<tr>
<td>H2</td>
<td>400</td>
<td>300</td>
<td>87</td>
</tr>
<tr>
<td>H2S</td>
<td>6800</td>
<td>2900</td>
<td>4200</td>
</tr>
<tr>
<td>WATER</td>
<td>24000</td>
<td>1400</td>
<td>23000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Temp (F):</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>570.5</td>
<td>570.5</td>
<td>570.5</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Pressure (psi):</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>290.1</td>
<td>290</td>
<td>290.1</td>
</tr>
</tbody>
</table>

### Design Data:

- **Materials of Construction:** Inconel  
- **Diameter:** 19 ft  
- **Height:** 104.5 ft  
- **Shell Thickness:** 2.7 in

**Comments:** Solvent regeneration consisting of simply a flash drum. The temperature and pressure are set by preheating heat exchangers and valves.
# H2S Flash Drum

**Identification:**  | **Item Name:**  | **Date:**  | **Item Number:**  |  
--- | --- | --- | --- |  
 | Vertical Flash Drum | 5/8/2012 | V-101S |  

**Function:** To separate H2S from Selexol

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-109S</td>
<td>1.25E+07</td>
<td>2.76E+05</td>
<td>1.22E+07</td>
</tr>
<tr>
<td>S-110S</td>
<td>1.23E+07</td>
<td>2.08E+05</td>
<td>1.21E+07</td>
</tr>
<tr>
<td>S-111S</td>
<td>7.48E+04</td>
<td>5.58E+04</td>
<td>1.90E+04</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition (lb/hr)</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>1.23E+07</td>
<td>2.08E+05</td>
<td>1.21E+07</td>
</tr>
<tr>
<td>CO2</td>
<td>7.48E+04</td>
<td>5.58E+04</td>
<td>1.90E+04</td>
</tr>
<tr>
<td>CO</td>
<td>7700</td>
<td>6800</td>
<td>900</td>
</tr>
<tr>
<td>H2</td>
<td>87</td>
<td>79</td>
<td>8</td>
</tr>
<tr>
<td>H2S</td>
<td>4200</td>
<td>2700</td>
<td>1500</td>
</tr>
<tr>
<td>WATER</td>
<td>2.30E+04</td>
<td>3200</td>
<td>1.98E+04</td>
</tr>
</tbody>
</table>

**Temp (F):**  
Feed: 567.2  
Distillate: 567.2  
Bottoms: 567.2

**Pressure (psi):**  
Feed: 87.0  
Distillate: 87.0  
Bottoms: 87.0

**Design Data:**

- **Materials of Construction:** Inconel
- **Diameter:** 19 ft
- **Height:** 102 ft
- **Shell Thickness:** 0.9 in

**Comments:** Solvent regeneration consisting of simply a flash drum. The temperature and pressure are set by preheating heat exchangers and valves.
# H2S Flash Drum

**Identification:** Vertical Flash  
**Item Name:** Drum  
**Item Number:** V-102S  
**Date:** 5/8/2012

**Function:** Separate Selexol for recycling

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-121S</td>
<td>4.63E+05</td>
<td>1.61E+05</td>
<td>3.02E+05</td>
</tr>
</tbody>
</table>

## Composition (lb/hr):

<table>
<thead>
<tr>
<th>Component</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>SELEXOL</td>
<td>2.95E+05</td>
<td>5.8437</td>
<td>2.95E+05</td>
</tr>
<tr>
<td>CO2</td>
<td>3.21E+04</td>
<td>1.30E+05</td>
<td>2500</td>
</tr>
<tr>
<td>CO</td>
<td>2.58E+04</td>
<td>2.58E+04</td>
<td>24</td>
</tr>
<tr>
<td>H2</td>
<td>380</td>
<td>380</td>
<td>0</td>
</tr>
<tr>
<td>H2S</td>
<td>5300</td>
<td>5000</td>
<td>300</td>
</tr>
<tr>
<td>WATER</td>
<td>4600</td>
<td>30</td>
<td>4600</td>
</tr>
</tbody>
</table>

## Temp (F):

<table>
<thead>
<tr>
<th>Temp (F)</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>95</td>
<td>95</td>
<td>95</td>
<td>95</td>
</tr>
</tbody>
</table>

## Pressure (psi):

<table>
<thead>
<tr>
<th>Pressure (psi)</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>87.0</td>
<td>87.0</td>
<td>87.0</td>
<td>87.0</td>
</tr>
</tbody>
</table>

**Design Data:**

- **Materials of Construction:** Inconel
- **Diameter:** 7.5 ft
- **Height:** 26.3 ft
- **Shell Thickness:** 0.4 in

**Comments:** Solvent regeneration consisting of simply a flash drum. The temperature and pressure are set by preheating heat exchangers and valves.
# Heat Exchanger

<table>
<thead>
<tr>
<th>Identification: Item Name:</th>
<th>Heat Exchanger</th>
</tr>
</thead>
<tbody>
<tr>
<td>Date:</td>
<td>4/18/2012</td>
</tr>
<tr>
<td>Item Number:</td>
<td>E-100W</td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between clean syngas and water.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-105W</td>
<td>325000</td>
<td>1740000</td>
<td>1740000</td>
<td>325000</td>
</tr>
<tr>
<td>S-104W</td>
<td>1740000</td>
<td>325000</td>
<td>325000</td>
<td></td>
</tr>
<tr>
<td>S-107W</td>
<td></td>
<td></td>
<td>1740000</td>
<td>325000</td>
</tr>
<tr>
<td>S-106W</td>
<td></td>
<td></td>
<td></td>
<td>325000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
</tr>
<tr>
<td>CO2</td>
</tr>
<tr>
<td>H2</td>
</tr>
<tr>
<td>H2O</td>
</tr>
<tr>
<td>H2S</td>
</tr>
<tr>
<td>SELEXOL</td>
</tr>
</tbody>
</table>

| T<sub>Hot</sub> In (F): | 1195 |
| T<sub>Hot</sub> Out (F): | 752.0 |
| T<sub>Cold</sub> In (F): | 86.0 |
| T<sub>Cold</sub> Out (F): | 392.0 |

| Pressure (psi): | 1193 |

**Design Data:**

- **Type:** Shell and Tube, One Pass, Triangle Pitch
- **Materials of Construction:** Tubes: Inconel 617 Shell: Carbon Steel
- **Heat Duty:** 3.74E+08 BTU/hr
- **Tube Diameter:** 0.75 in
- **Shell Diameter:** 37 in
- **Heat Transfer Coefficient:** 117.61 BTU/hr-ft²K
- **Number of Tubes:** 1074
- **Triangle Pitch:** 1 in
- **Heat Transfer Area:** 4341 ft²

**Comments:** E-100W is an interstage cooling heat exchanger positioned between high temperature water gas shift reactor R-100W and low temperature water gas shift reactor R-101W. The one pass shell and tube heat exchanger cools syngas and makes steam that preheats fuel entering the gas turbine.
# Heat Exchanger

**Identification:** Item Name:
**Heat Exchanger**
**Item Number:**
**Date:**
4/18/2012

**Function:** To exchange energy between clean syngas and water.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-109W</td>
<td>849000</td>
<td>1740000</td>
<td>1740000</td>
<td>849000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
</tr>
<tr>
<td>CO2</td>
</tr>
<tr>
<td>H2</td>
</tr>
<tr>
<td>H2O</td>
</tr>
<tr>
<td>H2S</td>
</tr>
<tr>
<td>SELEXOL</td>
</tr>
</tbody>
</table>

**T**

| Hot In (F): | 952 |
| Hot Out (F):| 95.0|
| Cold In (F):| 86.0|
| Cold Out (F):| 545.0|

**Pressure (psi):**

1187

**Design Data:**

- **Type:** Shell and Tube, One Pass, Triangle Pitch
- **Materials of Construction:**
  - Tubes: Inconel 617
  - Shell: Carbon Steel
- **Heat Duty:** $1.04E+09$ BTU/hr
- **Tube Diameter:** 0.75 in
- **Shell Diameter:** 37 in
- **Heat Transfer Coefficient:** 164.3 BTU/hr-ft$^2$K
- **Number of Tubes:** 1074
- **Triangle Pitch:** 1 in
- **Heat Transfer Area:** 71500 ft$^2$

**Comments:** E-101W is an interstage cooling heat exchanger positioned between low temperature water gas shift reactor R-101W and vertical pressure vessel V-104S. The one pass shell and tube heat exchanger cools syngas and makes steam that is used for the HRSG.
## Compressor

<table>
<thead>
<tr>
<th>Identification</th>
<th>Item Name:</th>
<th>Date:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td>Compressor</td>
<td>4/17/2012</td>
</tr>
</tbody>
</table>

**Function:** To compress from T-100S to R-100W

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-102W</td>
<td>S-103W</td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>1740000</td>
<td>1740000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO2</td>
<td>164000</td>
<td>164000</td>
</tr>
<tr>
<td>CO</td>
<td>742000</td>
<td>742000</td>
</tr>
<tr>
<td>H2S</td>
<td>1200</td>
<td>1200</td>
</tr>
<tr>
<td>H2</td>
<td>37000</td>
<td>37000</td>
</tr>
<tr>
<td>H2O</td>
<td>794000</td>
<td>794000</td>
</tr>
<tr>
<td>Selexol</td>
<td>30</td>
<td>30</td>
</tr>
</tbody>
</table>

| Temperature in (F): | 536.1 |
| Pressure in (psi):  | 400   |
| Temperature out (F): | 944.6 |
| Pressure out (psi): | 1200  |

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Centrifugal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>39400 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>800 psi</td>
</tr>
<tr>
<td>Power Required:</td>
<td>89000 kW</td>
</tr>
<tr>
<td>Efficiency:</td>
<td>0.75</td>
</tr>
</tbody>
</table>

**Comments:** K-100W compresses the sweet syngas to 1200 psia which is the reaction pressure in the HTS water gas shift reactor. It should be modeled as a multi-stage compressor. Once updated, the power requirement (88.97 MW) will be greatly diminished.
### Reactor

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Vertical Reactor</th>
<th>Date: 4/17/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td></td>
<td>R-100W</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To react unshifted syngas to produce hydrogen and water (HTS)

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-103W</td>
<td>742000</td>
<td>356000</td>
</tr>
<tr>
<td>S-104W</td>
<td>164000</td>
<td>780000</td>
</tr>
</tbody>
</table>

| Quantity (lb/hr) | 1739000 | 1739000 |
| CO            | 742000  | 356000  |
| CO2           | 164000  | 780000  |
| H2            | 37000   | 65000   |
| H2O           | 794000  | 546000  |
| H2S           | 1200    | 1200    |
| Selexol       | 30      | 30      |

| Inlet Temp (F): | 944.6 |
| Outlet Temp (F): | 1195 |
| Pressure (psi): | 1200 |

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Packed Bed Reactor</th>
</tr>
</thead>
<tbody>
<tr>
<td>Catalyst Type:</td>
<td>Pellet: 6mm x 6mm</td>
</tr>
<tr>
<td>Materials of Construction:</td>
<td>Inconel 617 (Refractory Lined)</td>
</tr>
<tr>
<td>Volume:</td>
<td>10600 ft³</td>
</tr>
<tr>
<td>Length:</td>
<td>65.6 ft</td>
</tr>
<tr>
<td>Diameter:</td>
<td>14.3 ft</td>
</tr>
<tr>
<td>Heat of Reaction:</td>
<td>-1.779E+04 btu/lbmol</td>
</tr>
</tbody>
</table>

**Comments:** The high temperature shift (HTS) water gas shift reactor uses a Fe₂O₃ (80-90%), Cr₂O₃ (8-13%), CuO (1-2%) pellet catalyst to shift CO and H₂O to H₂ and CO₂. The reaction is exothermic and reversible, and the rate is equilibrium limiting.
### Reactor

**Identification:** Vertical Reactor  
**Item Name:** Vertical Reactor  
**Item Number:** R-101W  
**Date:** 4/17/2012

**Function:** To react unshifted syngas to produce hydrogen and water (LTS)

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-107W</td>
<td>356000</td>
<td>62000</td>
</tr>
<tr>
<td>S-108W</td>
<td>780000</td>
<td>1233000</td>
</tr>
</tbody>
</table>

**Composition:**

- **CO:** 356000 lb/hr  
- **CO2:** 780000 lb/hr  
- **H2:** 65000 lb/hr  
- **H2O:** 546000 lb/hr  
- **H2S:** 1200 lb/hr  
- **Selexol:** 30 lb/hr

**Inlet Temp (F):** 752  
**Outlet Temp (F):** 951.8  
**Pressure (psi):** 1193

**Design Data:**

- **Type:** Packed Bed Reactor  
- **Catalyst Type:** Pellet: 6mm x 6mm  
- **Materials of Construction:** Inconel 617 (Refractory Lined)  
- **Volume:** 56500 ft³  
- **Length:** 197 ft  
- **Diameter:** 19.1 ft  
- **Heat of Reaction:** -1.779E+04 btu/lbmol

**Comments:** The low temperature shift (LTS) water gas shift reactor uses a CuO (50%), Al₂O₃/ZnO (50%) pellet catalyst to shift CO and H₂O to H₂ and CO₂. The reaction is exothermic and reversible, and the rate is equilibrium limiting.
### Separator

**Identification:** Item Name: Vertical Separator  | Date: 4/17/2012
**Item Number:** V-100W

**Function:** To separate water from shifted syngas to be sent to CO2 absorber

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-111W</td>
<td>1739000</td>
<td>13800000</td>
<td>3590000</td>
</tr>
</tbody>
</table>

**Composition:**

<table>
<thead>
<tr>
<th>Component</th>
<th>Feed</th>
<th>Distillate</th>
<th>Bottoms</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO2</td>
<td>1232000</td>
<td>12280000</td>
<td>4000</td>
</tr>
<tr>
<td>CO</td>
<td>62000</td>
<td>62000</td>
<td>0</td>
</tr>
<tr>
<td>H2S</td>
<td>1200</td>
<td>1100</td>
<td>30</td>
</tr>
<tr>
<td>H2</td>
<td>86000</td>
<td>86000</td>
<td>0</td>
</tr>
<tr>
<td>H2O</td>
<td>357000</td>
<td>1800</td>
<td>3550000</td>
</tr>
<tr>
<td>Selexol</td>
<td>30</td>
<td>0</td>
<td>30</td>
</tr>
</tbody>
</table>

**Temp (F):** 95  
**Pressure (psi):** 1189

**Design Data:**

- Materials of Construction: Inconel 617
- Volume: 6700 ft³
- Diameter: 11.5 ft
- Length: 63.3 ft

**Comments:** V-100W Separates water and selexol from the syngas as it heads to the CO2 Absorber. This water selexol mixture is sent to water treatment.
**Heat Exchanger**

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Heat Exchanger</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td>E-100T</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in (lb/hr)</th>
<th>Hot In (lb/hr)</th>
<th>Hot Out (lb/hr)</th>
<th>Cold Out (lb/hr)</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-101T</td>
<td>209000</td>
<td>325000</td>
<td>325000</td>
<td>209000</td>
</tr>
<tr>
<td>S-102T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-103T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-104T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Quantity (lb/hr):**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Selexol</th>
<th>CO2</th>
<th>CO</th>
<th>H2S</th>
<th>H2</th>
<th>H2O</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-101T</td>
<td>13</td>
<td>46000</td>
<td>79000</td>
<td>26</td>
<td>83000</td>
<td>325000</td>
</tr>
<tr>
<td>S-102T</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>325000</td>
</tr>
<tr>
<td>S-103T</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>325000</td>
</tr>
<tr>
<td>S-104T</td>
<td>13</td>
<td>46000</td>
<td>79000</td>
<td>26</td>
<td>83000</td>
<td>325000</td>
</tr>
</tbody>
</table>

**T<sub>Hot</sub> In (F):** 312.8

**T<sub>Hot</sub> Out (F):** 95.0

**T<sub>Cold</sub> In (F):** 244.8

**T<sub>Cold</sub> Out (F):** 320.0

**Pressure (psi):** 2200

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Fixed Head Shell and Tube</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heat Duty:</td>
<td>7000 kW</td>
</tr>
<tr>
<td>Shell Diameter:</td>
<td>37 in</td>
</tr>
<tr>
<td>Heat Transfer Coefficient:</td>
<td>70 Btu/(F-ft^2-hr)</td>
</tr>
<tr>
<td>Heat Transfer Area:</td>
<td>4000 ft^2</td>
</tr>
</tbody>
</table>

**Materials of Construction:** Carbon Steel Shell/Carbon Steel Tube

**Comments:** E-100T is used to heat the compressed hydrogen stream coming from K-100T. This heater prepares the hydrogen stream for injection into turbine K-100FG.
**Heat Exchanger**

**Identification:** Heat Exchanger

**Item Name:** Heat Exchanger

**Item Number:** E-101T

**Date:** 5/8/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-115T</td>
<td>9625000</td>
<td>3753000</td>
<td>3753000</td>
<td>9625000</td>
</tr>
<tr>
<td>S-117T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-114T</td>
<td>9625000</td>
<td>3753000</td>
<td>3753000</td>
<td>9625000</td>
</tr>
<tr>
<td>S-116T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Composition:**

- **N2:** 0 2878913 2878913 0
- **O2:** 0 874201 874201 0
- **H2O:** 9625000 0 0 9625000

**T_{Hot In} (F):** 378.1

**T_{Hot Out} (F):** 95.0

**T_{Cold In} (F):** 86.0

**T_{Cold Out} (F):** 113.0

**Pressure (psi):** 51.3

**Design Data:**

- **Type:** Fixed Head Shell and Tube
- **Materials of Construction:** Carbon Steel Shell/Carbon Steel Tube
- **Heat Duty:** 77000 kW
- **Shell Diameter:** 37 in
- **Heat Transfer Coefficient:** 328 Btu/(F-ft^2-hr)
- **Heat Transfer Area:** 10000 ft^2

**Comments:** E-101T is used to cool the compressed ambient air coming from K-101T. This interstage cooler increases compressor efficiency by cooling the stream before additional compression in K-102T.
**Heat Exchanger**

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Heat Exchanger</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td></td>
<td>E-102T</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold In</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-111T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-113T</td>
<td>10649000</td>
<td>3753000</td>
<td>3753000</td>
<td>10649000</td>
</tr>
<tr>
<td>S-110T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-112T</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Cold In</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>N2</td>
<td>0</td>
<td>2878913</td>
<td>2878913</td>
<td>0</td>
</tr>
<tr>
<td>O2</td>
<td>0</td>
<td>874201</td>
<td>874201</td>
<td>0</td>
</tr>
<tr>
<td>H2O</td>
<td>10649000</td>
<td>0</td>
<td>0</td>
<td>10649000</td>
</tr>
</tbody>
</table>

| T<sub>Hot</sub> In (F): | 404.8 |
| T<sub>Hot</sub> Out (F): | 95.0  |
| T<sub>Cold</sub> In (F): | 86.0  |
| T<sub>Cold</sub> Out (F):| 113.0 |

| Pressure (psi): | 179.0 |

**Design Data:**

| Type:                  | Fixed Head Shell and Tube |
|                       | Carbon Steel Shell/Carbon Steel Tube |
| Materials of Construction: |                         |
| Heat Duty:             | 85000 kW                 |
| Shell Diameter:        | 37 in                    |
| Heat Transfer Coefficient: | 333 Btu/(F-ft^2-hr)  |
| Heat Transfer Area:    | 10800 ft^2               |

**Comments:** E-102T is used to cool the compressed ambient air coming from K-102T. This interstage cooler increases compressor efficiency by cooling the stream before additional compression in K-103T.
# Heat Exchanger

**Identification:** Heat Exchanger  
**Item Number:** E-103T  
**Date:** 5/8/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-111T</td>
<td>0</td>
<td>2878913</td>
<td>2878913</td>
<td>0</td>
</tr>
<tr>
<td>S-113T</td>
<td>3753000</td>
<td>874201</td>
<td>874201</td>
<td>0</td>
</tr>
<tr>
<td>S-110T</td>
<td>3753000</td>
<td>0</td>
<td>0</td>
<td>3753000</td>
</tr>
<tr>
<td>S-112T</td>
<td>3753000</td>
<td>0</td>
<td>0</td>
<td>3753000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>N2</td>
<td>0</td>
<td>2878913</td>
<td>2878913</td>
<td>0</td>
</tr>
<tr>
<td>O2</td>
<td>0</td>
<td>874201</td>
<td>874201</td>
<td>0</td>
</tr>
<tr>
<td>H2O</td>
<td>3753000</td>
<td>0</td>
<td>0</td>
<td>3753000</td>
</tr>
</tbody>
</table>

- **T<sub>Hot In (F)</sub>:** 405.6
- **T<sub>Hot Out (F)</sub>:** 95.0
- **T<sub>Cold In (F)</sub>:** 86.0
- **T<sub>Cold Out (F)</sub>:** 113.0
- **Pressure (psi):** 624.5

### Design Data:

| Type: Fixed Head Shell and Tube  
| Carbon Steel Shell/Carbon Steel Tube |
| Materials of Construction: |
| Heat Duty: 88000 kW |
| Shell Diameter: 37 in |
| Heat Transfer Coefficient: 340 Btu/(F-ft^2-hr) |
| Heat Transfer Area: 11000 ft^2 |

**Comments:** E-103T is used to cool the compressed ambient air coming from K-103T. This interstage cooler increases compressor efficiency by cooling the stream before additional compression in K-104T.
# Heat Exchanger

**Identification:** Heat Exchanger  
**Item Number:** E-104T  
**Date:** 5/8/2012

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Cold in</th>
<th>Hot In</th>
<th>Hot Out</th>
<th>Cold Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-121T</td>
<td>1089000</td>
<td>3962000</td>
<td>3962000</td>
<td>1089000</td>
</tr>
<tr>
<td>S-122T</td>
<td>2879000</td>
<td>2879000</td>
<td>2879000</td>
<td>0</td>
</tr>
<tr>
<td>S-123T</td>
<td>0</td>
<td>1678000</td>
<td>168000</td>
<td>0</td>
</tr>
<tr>
<td>S-120T</td>
<td>1089000</td>
<td>745000</td>
<td>745000</td>
<td>1089000</td>
</tr>
<tr>
<td>Selexol</td>
<td>0</td>
<td>13</td>
<td>13</td>
<td>0</td>
</tr>
<tr>
<td>CO2</td>
<td>0</td>
<td>170000</td>
<td>170000</td>
<td>0</td>
</tr>
<tr>
<td>H2S</td>
<td>0</td>
<td>26</td>
<td>26</td>
<td>0</td>
</tr>
</tbody>
</table>

**T\textsubscript{Hot} In (F):** 1400  
**T\textsubscript{Hot} Out (F):** 158.0  
**T\textsubscript{Cold} In (F):** 125.1  
**T\textsubscript{Cold} Out (F):** 1100.0  
**Pressure (psi):** 2200.0

## Design Data:

- **Type:** Fixed Head Shell and Tube  
- **Materials of Construction:** Carbon Steel Shell/Carbon Steel Tube  
- **Heat Duty:** 450000 kW  
- **Shell Diameter:** 37 in  
- **Heat Transfer Coefficient:** 130 Btu/(F-ft\(^2\)-hr)  
- **Heat Transfer Area:** 148000 ft\(^2\)

**Comments:** E-104T is used to cool the hot stream coming from K-100FG. This heat exchanger produces steam to be used by K-100ST.
# Heat Exchanger

<table>
<thead>
<tr>
<th>Identification: Item Name:</th>
<th>Heat Exchanger</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td>E-105T</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To exchange energy between hot and cold streams.

**Operation:** Continuous

<table>
<thead>
<tr>
<th>Materials Handled:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Stream ID:</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Composition:</td>
</tr>
<tr>
<td></td>
</tr>
</tbody>
</table>

| T<sub>Hot In</sub> (F): 165.4 |
| T<sub>Hot Out</sub> (F): 122.0 |
| T<sub>Cold In</sub> (F): 86.0 |
| T<sub>Cold Out</sub> (F): 113.0 |

| Pressure (psi): 2200.0 |

<table>
<thead>
<tr>
<th>Design Data:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Type: Fixed Head Shell and Tube</td>
</tr>
<tr>
<td>Carbon Steel Shell/Carbon Steel Tube</td>
</tr>
<tr>
<td>Heat Duty: 326000 kW</td>
</tr>
<tr>
<td>Shell Diameter: 37 in</td>
</tr>
<tr>
<td>Heat Transfer Coefficient: 212 Btu/(F-ft^2-hr)</td>
</tr>
<tr>
<td>Heat Transfer Area: 125000 ft^2</td>
</tr>
</tbody>
</table>

**Comments:** E-105T is used to condense the vapor stream leaving K-100ST so it can be pumped by P-100T.
## Compressor

**Identification:** Item Name: Compressor  
**Item Number:** K-100T  
**Date:** 5/8/2012

### Function:
To compress syngas from T-100C to fuel gas turbine pressure

### Operation:
Continuous

### Materials Handled:
<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-100T</td>
<td>S-101T</td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>209000</td>
<td>209000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO2</td>
<td>46000</td>
<td>46000</td>
</tr>
<tr>
<td>CO</td>
<td>79000</td>
<td>79000</td>
</tr>
<tr>
<td>H2</td>
<td>83000</td>
<td>83000</td>
</tr>
</tbody>
</table>

| T In (F) | 95.6 |
| T Out (F) | 244.8 |
| Pressure In (psi): | 1160 |
| Pressure Out (psi): | 2176 |

### Design Data:
- **Type:** Centrifugal
- **Volumetric Flow Rate:** 240,000 ft³/hr
- **Pressure Change:** 1015 psi
- **Power Required:** 14000 kW
- **Efficiency:** 0.75

### Comments:
K-100T is a centrifugal compressor that takes air and compresses it for the fuel gas turbine pressure
### Compressor

<table>
<thead>
<tr>
<th><strong>Identification:</strong></th>
<th>Item Name: Compressor</th>
<th>Date: 5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Item Number:</strong></td>
<td>K-101T</td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To compress air from the atmosphere to fuel gas turbine pressure

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>S-118T</td>
<td>S-117T</td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>3753000</td>
<td>3753000</td>
</tr>
<tr>
<td>Composition:</td>
<td>N2 2879000</td>
<td>O2 874000</td>
</tr>
<tr>
<td></td>
<td>2879000</td>
<td>874000</td>
</tr>
</tbody>
</table>

| T In (F)  | 77.0 |
| T Out (F) | 378.1|
| Pressure In (psi): | 14.7 |
| Pressure Out (psi):  | 51.3 |

**Design Data:**

<table>
<thead>
<tr>
<th>Type:</th>
<th>Centrifugal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>50,950,000 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>36.7 psi</td>
</tr>
<tr>
<td>Power Required:</td>
<td>81000 kW</td>
</tr>
<tr>
<td>Efficiency</td>
<td>0.75</td>
</tr>
</tbody>
</table>

**Comments:** K-101T is a centrifugal compressor that takes air and compresses it for the fuel gas turbine
**Compressor**

**Identification:** Item Name: Compressor  
**Item Number:** K-102T  
**Date:** 5/8/2012

**Function:** To compress air from the atmosphere to fuel gas turbine pressure  
**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>Quantity (lb/hr)</td>
<td>S-114T</td>
<td>S-113T</td>
</tr>
<tr>
<td>Composition:</td>
<td>3753000</td>
<td>3753000</td>
</tr>
<tr>
<td>N2</td>
<td>2879000</td>
<td>2879000</td>
</tr>
<tr>
<td>O2</td>
<td>874000</td>
<td>874000</td>
</tr>
</tbody>
</table>

| T In (F) | 95.0 |
| T Out (F) | 404.8 |
| Pressure In (psi): | 51.3 |
| Pressure Out (psi): | 179 |

**Design Data:**

| Type: | Centrifugal |
| Volumetric Flow Rate: | 15,050,000 ft³/hr |
| Pressure Change: | 127.6 psi |
| Power Required: | 84000 kW |
| Efficiency | 0.75 |

**Comments:** K-102T is a centrifugal compressor that takes air and compresses it for the fuel gas turbine
### Compressor

**Identification:**
- **Item Name:** Compressor
- **Item Number:** K-103T
- **Date:** 5/8/2012

**Function:** To compress air from the atmosphere to fuel gas turbine pressure

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-110T</td>
<td>S-109T</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>3753000</td>
<td>3753000</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>N2</td>
<td>2879000</td>
<td>2879000</td>
</tr>
<tr>
<td>O2</td>
<td>874000</td>
<td>874000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>T In (F)</th>
<th>95.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>T Out (F)</td>
<td>405.6</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Pressure In (psi):</th>
<th>179</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure Out (psi):</td>
<td>644.5</td>
</tr>
</tbody>
</table>

**Design Data:**

- **Type:** Centrifugal
- **Volumetric Flow Rate:** 4,300,000 ft³/hr
- **Pressure Change:** 445.6 psi
- **Power Required:** 84000 kW
- **Efficiency:** 0.75

**Comments:** K-103T is a centrifugal compressor that takes air and compresses it for the fuel gas turbine
# Compressor

**Identification:**

| Item Name: | Compressor | Item Number: | K -104T | Date: | 5/8/2012 |

**Function:**
To compress air from the atmosphere to fuel gas turbine pressure

**Operation:** Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-106T</td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-105T</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Quantity (lb/hr)</td>
<td>3753000</td>
<td>3753000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>N2</td>
<td>2879000</td>
<td>2879000</td>
</tr>
<tr>
<td>O2</td>
<td>874000</td>
<td>874000</td>
</tr>
</tbody>
</table>

| T In (F) | 95.0 |
| T Out (F) | 405.2 |
| Pressure In (psi): | 624.5 |
| Pressure Out (psi): | 2176 |

## Design Data:

<table>
<thead>
<tr>
<th>Type:</th>
<th>Centrifugal</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>1,219,000 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>1551 psi</td>
</tr>
<tr>
<td>Power Required:</td>
<td>85000 kW</td>
</tr>
<tr>
<td>Efficiency</td>
<td>0.75</td>
</tr>
</tbody>
</table>

**Comments:** K-104T is a centrifugal compressor that takes air and compresses it for the fuel gas turbine.
# Turbine Unit Equipment Specification Sheets

## Turbine

<table>
<thead>
<tr>
<th>Identification:</th>
<th>Item Name:</th>
<th>Turbine</th>
<th>Date:</th>
<th>5/8/2012</th>
</tr>
</thead>
<tbody>
<tr>
<td>Item Number:</td>
<td>K-100FG</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To make power using syngas as the fuel source

**Operation:** Continuous

### Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
<th>Quantity (lb/hr)</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-119T</td>
<td>170000</td>
<td>170000</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-120T</td>
<td>2879000</td>
<td>2879000</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-119T</td>
<td>168000</td>
<td>168000</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-120T</td>
<td>745000</td>
<td>745000</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Composition:**

- CO2: 170000 in, 170000 out
- N2: 2879000 in, 2879000 out
- O2: 168000 in, 168000 out
- H2O: 745000 in, 745000 out

| Inlet Temp (F): | 3684 |
| Outlet Temp (F): | 1413 |
| Pressure in (psi): | 2176 |
| Pressure out (psi): | 14.7 |

### Design Data:

<table>
<thead>
<tr>
<th>Type:</th>
<th>Radial-Flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>3184000 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>-2161 psi</td>
</tr>
<tr>
<td>Power Generated:</td>
<td>991000 kW</td>
</tr>
<tr>
<td>Efficiency</td>
<td>0.75</td>
</tr>
</tbody>
</table>

**Comments:** K-100FG provides the main power for the plant. It uses high temperature and pressure syngas as a fuel source, and provides enough heat at the outlet to generate steam to run the steam turbine K-100ST.
## Turbine Unit Equipment Specification Sheets

<table>
<thead>
<tr>
<th>Turbine</th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Identification:</strong></td>
<td><strong>Item Name:</strong></td>
<td>Turbine</td>
<td><strong>Date:</strong> 5/8/2012</td>
</tr>
<tr>
<td><strong>Item Number:</strong></td>
<td>K-100ST</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**Function:** To make power using steam as the fuel source

**Operation:** Continuous

**Materials Handled:**

<table>
<thead>
<tr>
<th>Stream ID:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-122T</td>
<td></td>
<td></td>
</tr>
<tr>
<td>S-124T</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>1089000</td>
<td>1089000</td>
<td>1089000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition:</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>H2O</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1089000</td>
<td>1089000</td>
<td></td>
</tr>
</tbody>
</table>

| Inlet Temp (F): | 1112 |
| Outlet Temp (F): | 165.4 |
| Pressure in (psi): | 2176 |
| Pressure out (psi): | 2.9 |

**Design Data:**

<table>
<thead>
<tr>
<th></th>
<th>Radial-Flow</th>
</tr>
</thead>
<tbody>
<tr>
<td>Volumetric Flow Rate:</td>
<td>428000 ft³/hr</td>
</tr>
<tr>
<td>Pressure Change:</td>
<td>-2173 psi</td>
</tr>
<tr>
<td>Power Generated:</td>
<td>126000 kW</td>
</tr>
<tr>
<td>Efficiency</td>
<td>0.75</td>
</tr>
</tbody>
</table>

**Comments:** K-100ST provides the added power for the plant. It uses high temperature and pressure steam as a driving force, and generates approximately 126 MW of extra power in tandem with the fuel gas turbine K-100FG.
# Pump & Motor

## Identification:
- **Item Name:** Pump
- **Item Number:** P-100T
- **Date:** 4/18/2012

## Function:
To pump condensed water in the steam loop back to K-100ST

## Operation:
Continuous

## Materials Handled:

<table>
<thead>
<tr>
<th>Stream ID</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>S-125T</td>
<td>S-121T</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Quantity (lb/hr)</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>1089000</td>
<td>1089000</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Composition</th>
<th>Inlet</th>
<th>Out</th>
</tr>
</thead>
<tbody>
<tr>
<td>H2O</td>
<td>1089000</td>
<td>1089000</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Temp (F)</th>
<th>Inlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>122</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>P&lt;sub&gt;in&lt;/sub&gt; (psi)</th>
<th>Inlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>2.9</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>P&lt;sub&gt;out&lt;/sub&gt; (psi)</th>
<th>Inlet</th>
</tr>
</thead>
<tbody>
<tr>
<td>2176</td>
<td></td>
</tr>
</tbody>
</table>

## Design Data:
- **Type:** Centrifugal
- **Volumetric Flow Rate:** 17600 ft<sup>3</sup>/hr
- **Pressure Change:** 2173 psi
- **Power Required:** 3721 hp
- **Efficiency:** 0.75
- **Materials of Construction:** Stainless Steel

## Comments:
P-100T is a stainless steel centrifugal pump that pressurized condensed water in the steam loop and sends it back to the steam turbine K-100ST.
Appendix C: Polymath Models for Gasification Kinetics

Shrinking Core Kinetic Model

The shrinking core model was used for the design of the gasifier as it provided the best description of gasification kinetics when compared with experimental data. Inputs include Coal Dry fuel analysis, coal feed rate, oxygen feed rate, water feed rate, gasifier temperature, and gasifier pressure. Outputs are syngas composition and carbon conversion.

\[ C = 74.05 \text{ wt percent} \]
\[ H = 6.25 \text{ wt percent} \]
\[ S = 0.97 \text{ wt percent} \]
\[ N = 0.71 \text{ wt percent} \]

\#pyrolysis
\[ \frac{d(F_{\text{Coal}})}{d(V)} = r_{\text{Coal}} \]
\[ F_{\text{Coal}}(0) = 74594.9 \text{ g/s} \]
\[ r_{\text{Coal}} = 1.14E5 \times e^{-8900/T} \times (F_{\text{Coal}} \times C/100 + F_{\text{Coal}} \times H/1000) \times (1 - 0.066 \times \ln(P_{\text{total}}) - F_{\text{Coal}})/1E6 \]

\[ \frac{d(F_{\text{C}})}{d(V)} = r_{\text{C}} \]
\[ F_{\text{C}}(0) = 0.00001 \text{ mol/s} \]
\[ F_{\text{Cin}} = 3173.51 \]
\[ \frac{d(F_{\text{O2}})}{d(V)} = r_{\text{O2}} \]
\[ F_{\text{O2}}(0) = 2121 \text{ mol/s} \]
\[ \frac{d(F_{\text{CO}})}{d(V)} = r_{\text{CO}} \]
\[ F_{\text{CO}}(0) = 0 \text{ mol/s} \]
\[ \frac{d(F_{\text{CO2}})}{d(V)} = r_{\text{CO2}} \]
\[ F_{\text{CO2}}(0) = 0 \text{ mol/s} \]
\[ \frac{d(F_{\text{H2O}})}{d(V)} = r_{\text{H2O}} \]
\[ F_{\text{H2O}}(0) = 2113 \text{ mol/s} \]
\[ \frac{d(F_{\text{H2}})}{d(V)} = r_{\text{H2}} \]
\[ F_{\text{H2}}(0) = 0 \]
\[ \frac{d(F_{\text{H2S}})}{d(V)} = r_{\text{H2S}} \text{ mol/s} \]
\[ F_{\text{H2S}}(0) = 0 \]
\[ F_{\text{Ar}} = 43.3 \text{ mol/s} \]
\[ \frac{d(F_{\text{NH3}})}{d(V)} = r_{\text{NH3}} \]
\[ F_{\text{NH3}}(0) = 0 \]

\[ \frac{d(P_{\text{total}})}{d(V)} = \left(1/(V+0.001)\right) \times (P_0/P_{\text{total}}) \times (F_{\text{total}}/F_{\text{total}0}) \text{ atm/cm}^3 \]
\[ P_{\text{total}0} = 40 \text{ atm} \]
\[ P_0 = 40 \text{ atm} \]
\[ F_{\text{total}0} = 7450.81 \]
\[ \frac{d(t)}{d(V)} = 1/(((1.5\times100)^2/4)\times3.14\times v_s) \]
\[ t(0) = 0 \]

\[ T = 1673 \]
\[ x = (F_{\text{Cin}} - F_{\text{C}})/F_{\text{Cin}} \]
\[ d_p = 0.05 \text{ cm} \]
#char O2 reactions
\[
 r_{\text{C}_2\text{O}_2} = \frac{(-1)}{0.292\phi^2(4.26/T)^*(T/1800)^1.75}/(P_{\text{total}}/d_p) + 1/(8710\exp(-17967/T)^*(1-x)/(1-0.5))^{(1/3)} \]
\[
 \phi = (2*2500^\exp(-6249/T)+2)/(2500^\exp(-6249/T)+2) \quad \# \text{for} \quad d_p \leq 0.005 \text{ cm}
\]

#char steam reactions
\[
r_{\text{C}_2\text{H}_2\text{O}} = ((-1)/(1/(10E-4*(T/2000)^0.75))/(P_{\text{total}}/d_p) + 1/(247^\exp(-21060/T)^*(1-x)/(1-0.5))^{(1/3)^2}) + (1/(10E-4*(T/2000)^0.75))(P_{\text{total}}/d_p) (1^2.5)(1^((1-x)/(1-0.5))^{(1/3)+0.001}) - 1))) \]
\[
 P_{\text{H}_2} = (P_{\text{H}_2} - P_{\text{CO}})/\exp(17.644 - 30260/(1.8*T)) \quad a/12 \quad \# \text{g/cm}^3\text{s}^3
\]

#char CO2 reactions
\[
r_{\text{C}_2\text{O}_2} = ((-1)/(1/(7.45E-4*(T/2000)^0.75))/(P_{\text{total}}/d_p) + 1/(247^\exp(-21060/T)^*(1-x)/(1-0.5))^{(1/3)^2+0.001}) + (1/(7.45E-4*(T/2000)^0.75))(P_{\text{total}}/d_p) (1^2.5)(1^((1-x)/(1-0.5))^{(1/3)+0.001}) - 1))) \]
\[
 P_{\text{CO}_2} = (P_{\text{CO}_2}/d_p) /\exp(17.644 - 30260/(1.8*T)) \quad a/12 \quad \# \text{g/cm}^3\text{s}^3
\]

#partial pressures of product gas
\[
P_{\text{H}_2} = (F_{\text{H}_2}/F_{\text{total}})^*P_{\text{total}}
\]
\[
P_{\text{O}_2} = (F_{\text{O}_2}/F_{\text{total}})^*P_{\text{total}}
\]
\[
P_{\text{H}_2O} = (F_{\text{H}_2O}/F_{\text{total}})^*P_{\text{total}}
\]
\[
P_{\text{CO}_2} = (F_{\text{CO}_2}/F_{\text{total}})^*P_{\text{total}}
\]
\[
P_{\text{CO}} = (F_{\text{CO}}/F_{\text{total}})^*P_{\text{total}}
\]

#mole fractions of product
\[
y_{\text{O}_2} = F_{\text{O}_2} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{CO}} = F_{\text{CO}} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{CO}_2} = F_{\text{CO}_2} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{H}_2O} = F_{\text{H}_2O} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{H}_2} = F_{\text{H}_2} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{H}_2S} = F_{\text{H}_2S} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{Ar}} = F_{\text{Ar}} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
y_{\text{NH}_3} = F_{\text{NH}_3} / (F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3})
\]
\[
 F_{\text{total}} = F_{\text{C}} + F_{\text{O}_2} + F_{\text{CO}} + F_{\text{CO}_2} + F_{\text{H}_2O} + F_{\text{H}_2} + F_{\text{H}_2S} + F_{\text{Ar}} + F_{\text{NH}_3}
\]
**Langmuir-Hinshelwood Kinetic Model**

Langmuir-Hinshelwood Kinetics provide a good estimation of the syngas exit composition. Inputs are char feed, oxygen feed, water feed, gasifier temperature and gasifier pressure.

\[
\frac{d(F_C)}{d(V)} = r_C \\
F_C(0) = 2100 \text{ mol/s} \\
\frac{d(F_O2)}{d(V)} = r_{O2} \\
F_O2(0) = 1470 \text{ mol/s} \\
F_{O2in} = 1470 \\
\frac{d(F_CO2)}{d(V)} = r_{CO2} \\
F_CO2(0) = 0 \text{ mol/s} \\
\frac{d(F_H2O)}{d(V)} = r_{H2O} \\
F_H2O(0) = 1000 \text{ mol/s} \\
F_{H2Oin} = 1000 \\
\frac{d(F_H2)}{d(V)} = r_{H2} \\
F_H2(0) = 0 \\
\frac{d(P_{total})}{d(V)} = \left(1/(Ac)\right) \left(P_0/P_{total}\right) \left(F_{total}/F_{total0}\right) \text{ Pa/cm}^3 \\
P_{total}(0) = 4E6 \text{ Pa} \\
P_0 = 4E6 \text{ Pa} \\
F_{total0} = 4570 \\
T = 1573 \text{ Kelvin} \\
\]

\[
r_{O2} = -\left(120^*k_a*y_{O2}/(1+ (k_a/(k_d+0.0001))*y_{O2})\right) \text{ mol/[cm}^3*s] \\
k_a = 8.75E-03*exp(-10040/(1.9859^*T)) \text{ mol/[cm}^2*s] \\
k_d = 163.56*exp(-32950/(1.9859^*T)) \text{ mol/[cm}^2*s] \\
\]

\[
r_{pCO} = (0.0006^*T-0.2165)^*(-r_{O2}) \\
r_{pCO2} = (-0.0006^*T+1.2165)^*(-r_{O2}) \\
r_{CO} = \text{if } F_C<0 \text{ then } 0 \text{ else } (r_{pCO} + r_{COrxn}) \\
r_{CO2} = \text{if } F_C<0 \text{ then } 0 \text{ else } (r_{pCO2} + r_{CO2rxn}) \\
r_{C} = \text{if } F_C<0 \text{ then } 0 \text{ else } (-r_{pCO} + r_{pCO2} + r_{Crxn}) \\
r_{Crxn} = -\left((K1^*P_{CO2} + K2^*P_{H2O})/(1+ K4^*P_{CO2} + K5^*P_{H2O})\right)/1000 \\
r_{H2O} = \text{if } F_C<0 \text{ then } 0 \text{ else } (-2^*(K2^*P_{H2O})/(1+ K4^*P_{CO2} + K5^*P_{H2O})/1000) \\
r_{H2} = -r_{H2O} \\
r_{CO2rxn} = \left((1.2)^*(2*K1^*P_{CO2}+K2^*P_{H2O})/(1+K4^*P_{CO2} + K5^*P_{H2O})\right)/1000 \\
r_{CO2rxn} = \text{if } F_C<0 \text{ then } 0 \text{ else } (1.2)*(K1^*P_{CO2} + (1+K4^*P_{CO2} + K5^*P_{H2O}))/1000 \\
\]

K1 = 1.445E01*exp(-2.214E05/(8.314^*T)) \\
K2 = 6.472E02*exp(-2.508E05/(8.314^*T)) \\
K4 = 1.155E-5*exp(-6.137E03/(8.314^*T)) \\
K5 = 5.592E-4*exp(-5.801E04/(8.314^*T))

\[
y_{O2} = F_{O2}/F_{total} \\
y_{CO} = F_{CO}/F_{total} \\
y_{CO2} = F_{CO2}/F_{total} \\
y_{H2O} = F_{H2O}/F_{total} \\
y_{H2} = F_{H2}/F_{total} \\
P_{O2} = (F_{O2}/F_{total})^*P_{total} \\
P_{H2O} = (F_{H2O}/F_{total})^*P_{total} \\
P_{CO2} = (F_{CO2}/F_{total})^*P_{total}
\]
\[ F_{\text{total}} = F_C + F_{O2} + F_{CO} + F_{CO2} + F_{H2O} + F_{H2} \]

\[ V_{\text{cubicm}} = V/(100^3) \]

\[ D = 1.5 \text{ m} \]
\[ Ac = ((D*100)^2/4)*3.14 \text{ cm}^2 \]
\[ A_{\text{sqrm}} = Ac/10000 \]

\[ V(0) = 0 \]
\[ V(f) = 10000000 \]

\[ L = V_{\text{cubicm}}/A_{\text{sqrm}} \text{ m} \]

\[ X_{H2O} = (F_{H2Oin} - F_{H2O})/F_{H2Oin} \]
\[ X_{pCO2} = ((F_{O2in} - F_{O2})/F_{O2in})*(-0.0006*T+1.2165) \]
\[ X_{pCO} = ((F_{O2in} - F_{O2})/F_{O2in})*(0.0006*T-0.2165) \]
\[ X_{CO2rxn} = (F_{pCO2} - F_{CO2})/(F_{pCO2}+0.00001) \]

\[ \frac{d(F_{pCO2})}{d(V)} = r_{pCO2} \]
\[ F_{pCO2}(0) = 0 \text{ #mol/s} \]
Appendix D: First Semester Work Breakdown Schedule
<table>
<thead>
<tr>
<th>ID</th>
<th>Task Name</th>
<th>Duration</th>
<th>Start</th>
<th>Finish</th>
<th>Prec1</th>
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<td>1</td>
<td>Assignments</td>
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<td>Wed 11/9/11</td>
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<td>Team at stations for Fridays at Calvin</td>
<td>1 day</td>
<td>Fri 11/11/11</td>
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<td>Mon 11/14/11</td>
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<td>10</td>
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<td>Mon 11/21/11</td>
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<td>Mon 10/31/11</td>
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<td>Mon 10/24/11</td>
<td>Tue 11/15/11</td>
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<td>Fri 10/28/11</td>
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Project: Project WBS
Date: Sat 1/14/12
Project: Project WBS
Date: Sat 1/14/12
Appendix E: Second Semester Work Breakdown Schedule
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<td>40</td>
<td>Senior Design Night</td>
<td>1 day</td>
<td>Mon 5/7/12</td>
<td>Mon 5/7/12</td>
</tr>
</tbody>
</table>
Appendix F: Equipment Cost Calculation Sheets

The following calculation sheets show the equations used to determine cost of the IGCC plant equipment. Unless specified otherwise, all equations were taken from Warren Seider’s ‘Product and Process Design Principles.’
Equipment Cost Calculations for Air Separation Unit

**PCI** := \( \frac{574}{500} \)  
Chemical Engineering Plant Cost Index - Adjusting to 2015 Values

### Main Air Compressors K-100A and K-101A - Centrifugal Compressors

- \( F_D := 1 \)  
- \( F_M := 1 \)  
- \( P_C := 3.975 \times 10^4 \)  
- \( C_B := \exp[7.5800 + 0.80(\ln(P_C))] \)  
- \( C_P := F_D F_M C_B PCI \)

\[ C_{K100} := C_P = 10748893.68 \]

Purchase Cost of K-100

\[ C_{K101} := C_P = 10748893.68 \]

Purchase Cost of K-101

### Heat Exchanger E-100A - Fixed Head Shell and Tube

- \( A := 1.827 \times 10^4 \)  
- \( P := 38.43 \)  
- \( C_{M} := \exp\left[11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2\right] \)  
- \( a := 0.00 \)  
- \( b := 0.00 \)

\[ F_M := a + \left(\frac{A}{100}\right)^b \]  

Material Factor (22.44)

\[ F_L := 1.00 \]  

Tube Length Correction

\[ F_P := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left(\frac{P}{100}\right) + 0.0017 \left(\frac{P}{100}\right)^2 & \text{otherwise} \end{cases} \]  

Pressure Factor (22.45)

\[ C_D := F_P F_M F_L C_B PCI \]

\[ C_{E100} := C_D = 112729.64 \]

Purchase Cost of E-100

### Heat Exchanger E-101A - Fixed Head Shell and Tube

- \( A := 3.600 \times 10^3 \)  
- \( P := 38.43 \)  
- \( C_B := \exp\left[11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2\right] \)  

Base Cost (22.40)
\( a = 1.75 \quad \text{Shell - Carbon Steel Materials of Construction Factors} \)
\( b = 0.13 \quad \text{Tubes - Stainless Steel} \)

\[ F_M := a + \left( \frac{A}{100} \right)^b \]
Material Factor (22.44)

\( F_L := 1.00 \quad \text{Tubes - 20 ft Tube Length Correction} \)

\[ F_P := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \]
Pressure Factor (22.45)

\[ C_E := F_P F_M F_L C_B PCI \]
Purchase Cost (22.43)

\[ C_{E101} := C_P = 94394.51 \]
Purchase Cost of E-101

**Heat Exchanger E-102A - Fixed Head Shell and Tube**

\( A := 1.379 \times 10^4 \quad \text{Heat Exchange Area (ft}^2\text{)} \)
\( P := 38.43 \quad \text{Shell Side Pressure (psia)} \)

\[ C_B := \exp(11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2) \quad \text{Base Cost (22.40)} \]

\( a := 1.75 \quad \text{Shell - Carbon Steel Materials of Construction Factors} \)
\( b := 0.13 \quad \text{Tubes - Stainless Steel} \)

\[ F_M := a + \left( \frac{A}{100} \right)^b \]
Material Factor (22.44)

\( F_L := 1.00 \quad \text{Tubes - 20 ft Tube Length Correction} \)

\[ F_P := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \]
Pressure Factor (22.45)

\[ C_E := F_P F_M F_L C_B PCI \]
Purchase Cost (22.43)

\[ C_{E101} := C_P = 311672.4 \]
Purchase Cost of E-101

**Vessel V-100A and V-101A - Air Pre-Purification Unit**

**Vessel Construction**

\( D := 6.56 \quad \text{D} := 12 \cdot D \quad \text{Vessel Diameter (ft)} \)
\( L := 39.04 \quad \text{Vessel Length (ft)} \)
\[ P_{o} := 79.57 \quad \text{Operating Pressure (psig)} \]
\[ S := 15000 \quad \text{Maximum Allowable Stress (psi)} \]
\[ E := 1 \quad \text{Weld Efficiency} \]
\[ F_{M} := 1.0 \quad \text{Carbon Steel Material Factor} \]
\[ \rho := 489.9 \quad \text{Density of Wall Material (lb/ft}^3\text{)} \]
\[ t_{P} := \frac{3}{8} \quad \text{Wall Thickness (in)} \]
\[ P_{d} := \exp\left(0.60608 + 0.91615 \ln(P_{o}) + 0.0015655 \ln(P_{o})^2\right) \quad \text{Design Pressure (psig) (22.61)} \]
\[ W := \pi(D + t_{P})(L + 0.8 \cdot D)t_{P}\rho \quad \text{Vessel Weight (22.59)} \]
\[ C_{V} := \exp\left(8.9552 - 0.2330 \ln(W) + 0.04333 \ln(W)^2\right) \quad \text{Horizontal Vessel Cost (22.53)} \]
\[ C_{PL} := 2005(D)^{0.20294} \quad \text{Horizontal Vessel Added Cost (22.55)} \]

**Adsorbant Cost**

\[ V_{\text{sieves}} := 1003 \quad \text{Volume of Sieves per Vessel (ft}^3\text{)} \]
\[ V_{\text{alumina}} := 148 \quad \text{Volume of Alumina per Vessel (ft}^3\text{)} \]
\[ C_{\text{sieves}} := 75 \cdot V_{\text{sieves}} \quad \text{Cost of Molecular Sieves (T-22.32)} \]
\[ C_{\text{alumina}} := 45 \cdot V_{\text{alumina}} \quad \text{Cost of Alumina (T-22.32)} \]

**Total Cost**

\[ C_{P} := (F_{M}C_{V} + C_{PL} + C_{\text{sieves}} + C_{\text{alumina}})PCI \quad \text{Purchase Cost (22.52)} \]
\[ C_{V100} := C_{P} = 395708.61 \quad \text{Purchase Cost of V-100} \]
\[ C_{V101} := C_{P} = 395708.61 \quad \text{Purchase Cost of V-101} \]

*Main Heat Exchanger E-103A - Brazed Aluminum Plate Fin*

\[ A := 101036 \quad \text{Heat Exchange Area (ft}^2\text{)} \]
\[ C_{P} := 12 \cdot A \cdot PCI \quad \text{Purchase Cost (bre.com)} \]
\[ C_{E103} := C_{P} = 1391871.94 \quad \text{Purchase Cost of E-103} \]

*Tower T-100A - High Pressure Distillation Column*

**Vessel Cost**

\[ D := 23.0 \quad \text{D}_{1} := 12 \cdot D \quad \text{Tower Diameter (ft)} \]
\[ L := 39.87 \quad \text{L}_{1} := 12 \cdot L \quad \text{Tower Height (ft)} \]
\[ P := 68.7 \quad \text{Operating Pressure (psig)} \]
\[ S := 15000 \quad \text{Maximum Allowable Stress (psi)} \]
\[ E := 1 \]
\[ F_M := 1.7 \quad \text{Stainless Steel 304} \]
\[ \rho := 490.0 \]
\[ P_{\text{exp}} := \exp \left( 0.60608 + 0.91615 \ln(P_o) + 0.0015655 \cdot \ln(P_o)^2 \right) \]
\[ t_w := \frac{0.22(D_1 + 18)L_1^2}{S \cdot D^2} \]
\[ W := \pi(D + t_p)(L + 0.8 \cdot D)t_w \rho \]
\[ C_V := \exp \left( 7.2756 + 0.18255 \ln(W) + 0.02297 \ln(W)^2 \right) \]
\[ C_{PL} := 300.9D^{0.63316} \cdot 0.80160 \]

**Packing Cost**

\[ V_p := \pi \left( \frac{D}{2} \right)^2 L \]
\[ C_{PK} := 250 \]
\[ n := 4 \]
\[ C_{DR} := 125n \]

**Total Cost**

\[ C_T := F_M C_V + C_{PL} + V_p^r C_{PK} + C_{DR} \]
\[ C_{T100} := C_P^r PCI = 13760904.88 \]

**Subcooler E-104A - Brazed Aluminum Plate Fin**

\[ A := 103011 \quad \text{Heat Exchange Area (ft}^2) \]
\[ C_P := 12 \cdot A \cdot PCI \]
\[ C_{Main} := C_P = 1419079.54 \]

**Turbine K-102A - Gas Expander**

\[ P := 1872.6 \]
\[ F_M := 1.7 \quad \text{Stainless Steel} \]
\[ C_K := 530 \cdot F_M \cdot P^{0.81} \cdot PCI \]
\[ C_{K102} := C_P = 462750.76 \]

**Tower T-101A - Low Pressure Distillation Column**
Vessel Cost

\[ D := 26.0 \quad D := 12 \cdot D \]
\[ L := 37.63 \quad L := 12 \cdot L \]
\[ P := 7.06 \]
\[ S := 15000 \]
\[ E := 1 \]
\[ F := 1.7 \quad \text{Stainless Steel 304} \]
\[ \rho := 490.0 \]
\[ P_d := \exp\left(0.60608 + 0.91615 \ln(P) + 0.0015655 \cdot \ln(P)^2\right) \]
\[ W := \pi(D + t)\left(L + 0.8 \cdot D\right)w^2 \rho \]
\[ C_V := \exp\left(7.2756 + 0.18255 \ln(W) + 0.182297 \ln(W)^2\right) \]
\[ C_W := 300.9D^{0.63316}L^{0.80160} \]

Tower Diameter (ft)
Tower Height (ft)
Operating Pressure (psig)
Maximum Allowable Stress (psi)
Weld Efficiency
Material Factor
Density of Wall Material (lb/ft^3)
Design Pressure (psig)
Wall Thickness (in)
Vessel Weight
Vertical Vessel Cost
Tower Added Cost

Packing Cost

\[ V := \pi\left(\frac{D}{2}\right)^2 L \]
\[ CP := 250 \]
\[ n := 4 \]
\[ CDR := 125n \]

Packing Volume (ft^3)
Installed Cost of Packing (per ft^3)
Liquid Distributors Needed
Installed Cost of Liquid Distributors

Total Cost

\[ C := \left(F \cdot C_V + C_{PL} + V \cdot C_{PK} + C_{DR}\right)PCI \]
\[ C_{T101} := C = 13834255.95 \]

Purchase Cost
Purchase Cost of T-101

Pump P-100A - Centrifugal Pump

Centrifugal Pump

\[ Q := 993.1 \]
\[ H := 1993.6 \]
\[ S := Q \cdot H^{0.5} \]
\[ C_B := \exp\left(9.7171 - 0.6019 \ln(S) + 0.0519 \ln(S)^2\right) \]
\[ F := 2.00 \quad \text{Stainless Steel} \]
\[ F_T := 8.90 \]

Flow Rate Through Pump (gpm)
Pump Head (ft)
Size Factor
Base Cost
Material Factor
Type Factor
\[ C_n := F_T F_M C_B PCI \]

\[ C_{P100} := C_P = 2.06 \times 10^5 \]

**Electric Motor**

\[ \rho := 151.3 \]

\[ P_B := 755.3 \]

\[ \eta_p := -0.316 + 0.24015 \ln(Q) - 0.01199 (\ln(Q))^2 \]

\[ \eta_M := 0.80 + 0.0319 \ln(P_B) \]

\[ P_C := \frac{Q \cdot H \cdot \rho}{33000 \cdot \eta_p \eta_M} \]

\[ P_B := \exp\left(5.8259 + 0.13141 \ln(P_C) + 0.053255 \ln(P_C)^2 + 0.028628 \ln(P_C)^3 - 0.0035549 \ln(P_C)^4\right) \]

\[ F_T := 1.8 \]

**Base Cost**

\[ C_M := C_P = 5.585 \times 10^3 \]

**Purchase Cost**

\[ C_{P100} := C_P = 2.06 \times 10^5 \]

**Purchase Cost of P-100**

**Density of Pumped Liquid (lb/gal)**

**Pump Brake Horsepower**

**Efficiency of Pump**

**Efficiency of Motor**

**Pump Power Consumption**

**Base Cost**

**Purchase Cost of M-100**
**Capital Cost of Gasification Section**

Demand Variables from Polymath Model

\[ m_{\text{coal}} := 6445 \quad \text{Mass Coal rate (tons/day)} \]
\[ m_{\text{syngas}} := 112500 \quad \text{syngas mass flow rate (lb/hr)} \]

**Coal Handling**

\[ C_{\text{CH}} := 5.466 \cdot m_{\text{coal}} \quad \text{Frey_Akunuri_2001} \]
\[ C_{\text{CH}} := \left( \frac{574}{394} \right) C_{\text{CH}}^{1000} = 5.132 \times 10^7 \quad \text{Combined Purchase Cost for Coal Handling/Treatment} \]

**Gasifier and Radiant Syngas Cooler**

\[ C_G := 15.88 \cdot m_{\text{coal}}^{0.943} \quad \text{Rubin 2007 Eq 22} \]
\[ C_{\text{R100G}} := \left( \frac{574}{394} \right) C_G^{1000} = 9.044 \times 10^7 \quad \text{Purchase Cost of Gasifier R-100G and RSC E-100G} \]

**Heat Exchanger E-101G - Fixed Head Shell and Tube**

\[ A := 4.067 \cdot 10^4 \quad \text{Heat Exchange Area (ft}^2\text{)} \]
\[ P := 38.43 \quad \text{Shell Side Pressure (psia)} \]
\[ C_B := \exp \left( 11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2 \right) \quad \text{Base Cost (22.40)} \]
\[ a := 1.75 \quad \text{Shell - Carbon Steel} \]
\[ b := 0.13 \quad \text{Tubes - Inconel} \]
\[ F_M := a + \left( \frac{A}{100} \right)^b \quad \text{Material Factor (22.44)} \]
\[ F_L := 1.00 \quad \text{Tubes - 20 ft} \]
\[ F_P := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \quad \text{Pressure Factor (22.45)} \]
\[ C_P := F_P \cdot F_M \cdot F_L \cdot C_B \quad \text{Purchase Cost (22.43)} \]
\[ C_{\text{E101G}} := \left( \frac{574}{500} \cdot C_P \right) = 1062083 \quad \text{Purchase Cost of E-101G} \]
**Pump P-100G - Centrifugal Pump**

### Centrifugal Pump

- \( Q := 1373.7 \)  
- \( H := 886.81 \)  
- \( S := Q \cdot H^{0.5} \)  
- \( C_B := \exp\left(9.7171 - 0.6019 \ln(S) + 0.0519 \ln(S)^2\right) \)  
- \( F := 2 \)  
- \( F_T := 8.90 \)  
- \( C_{CP100} := C_P \cdot \frac{574}{500} = 1.978 \times 10^5 \)

### Electric Motor

- \( \rho := 10.85 \)  
- \( P_B := 505 \)  
- \( \eta_P := -0.316 + 0.24015 \ln(Q) - 0.01199 (\ln(Q))^2 \)  
- \( \eta_M := 0.80 + 0.03191 \ln(P_B) \)  
- \( P_C := \frac{Q \cdot H \cdot \rho}{33000 \cdot \eta_P \cdot \eta_M} \)  
- \( C_{CM101} := C_P \cdot \frac{574}{500} = 4.665 \times 10^4 \)

### Base Cost Formulas

- \( S := Q \cdot H^{0.5} \)  
- \( C_B := \exp\left(9.7171 - 0.6019 \ln(S) + 0.0519 \ln(S)^2\right) \)  
- \( F := 2 \)  
- \( F_T := 8.90 \)  
- \( C_{CP100} := C_P \cdot \frac{574}{500} = 1.978 \times 10^5 \)

### Motor Type Factor

- \( F_T := 1.4 \)  
- \( C_{CM101} := C_P \cdot \frac{574}{500} = 4.665 \times 10^4 \)
Water Gas Shift Equipment Costs

**Syngas Compressor K-100W - Centrifugal Compressor**

- $F_D := 1$  
  Electric Motor
- $F_M := 3.9$  
  Inconel 617
- $P_C := 1.193 \cdot 10^5$  
  Horsepower
- $C_B := \exp[7.5800 + 0.80 \ln(P_C)]$  
  Base Cost (22.36)
- $C_P := F_D F_M C_B$  
  Purchase Cost (22.35)
- $C_{K100} := C_P = 8796894.61$  
  Purchase Cost of K-100W

**Heat Exchanger E-100W - Fixed Head Shell and Tube**

- $A := 4.341 \cdot 10^3$  
  Heat Exchange Area (ft^2)
- $P := 87.02 - 14.7$  
  Shell Side Pressure (psia)
- $C_B := \exp(11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2)$  
  Base Cost (22.40)
- $a := 3.5$  
  Shell - Carbon Steel
- $b := 0.13$  
  Tubes - Inconel 617
- $F_M := a + \left(\frac{A}{100}\right)^b$  
  Material Factor (22.44)
- $F_L := 1.00$  
  Tubes - 20 ft
- $F_P := \begin{cases} 
1 & \text{if } P \leq 114.7 \\
0.9803 + 0.018 \left(\frac{P}{100}\right) + 0.0017 \left(\frac{P}{100}\right)^2 & \text{otherwise} 
\end{cases}$  
  Pressure Factor (22.45)
- $C_B := F_P F_M F_L C_B$  
  Purchase Cost (22.43)
- $C_{E100} := C_P = 144186.28$  
  Purchase Cost of E-100W

**Heat Exchanger E-101W - Fixed Head Shell and Tube**

- $A := 7.151 \cdot 10^4$  
  Heat Exchange Area (ft^2)
- $P := 87.02 - 14.7$  
  Shell Side Pressure (psia)
- $C_B := \exp(11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2)$  
  Base Cost (22.40)
- $a := 3.5$  
  Shell - Carbon Steel
- $b := 0.13$  
  Tubes - Inconel 617
- $C_{E101} := C_P$  
  Purchase Cost of E-101W
Material Factor (22.44)

\[
F_M := a + \left( \frac{A}{100} \right)^b
\]

Tube Length Correction

\[
F_L := 1.00 \quad \text{Tubes - 20 ft}
\]

Pressure Factor (22.45)

\[
F_P := \begin{cases} 
1 & \text{if } P \leq 114.7 \\
0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise}
\end{cases}
\]

Purchase Cost (22.43)

\[
C_P := F_P F_M F_L C_B
\]

Purchase Cost of E-101W

\[
C_{E101} := C_P = 2748187.13
\]

**Reactor R-100W - High Temperature Reactor**

Weight of a pressure vessel, assuming constant shell and head thickness with 2:1 elliptical heads,

\[
\rho := 521.9 \quad \text{Density of Inconel 617 (lb/ft}^3\text{)}
\]

\[
D_i := 14.34 \quad \text{Diameter (ft)}
\]

\[
L := 65.617 \quad \text{Length (ft)}
\]

\[
P_o := 1200 - 14.7 \quad \text{Pressure (psig)}
\]

\[
T := 646 \quad \text{Temperature (C)}
\]

\[
S := \begin{cases} 
15000 & \text{if } T \leq 750 \\
14750 & \text{if } 750 < T \leq 800 \\
14200 & \text{if } 800 < T \leq 850 \\
13100 & \text{if } 850 < T \leq 900
\end{cases}
\]

\[
E := 1
\]

\[
P_d := \begin{cases} 
1.1 \cdot P_o & \text{if } P_o > 1000 \\
\exp \left[ 0.60608 + 0.91615 \cdot \ln(P_o) + 0.0015655 \cdot (\ln(P_o))^2 \right] & \text{otherwise}
\end{cases}
\]

\[
\frac{(P_d \cdot D_i)}{(2 \cdot S \cdot E - 1.2 \cdot P_d)}
\]

\[
\frac{tp = 0.658}{ts := \frac{11}{16}} \quad \text{Wall Thickness (in)}
\]

\[
W := \pi (ts + D_i) \cdot (L + 0.8 \cdot D_i) \cdot ts \cdot \rho
\]

\[
W = 1305837.15 \quad \text{weight of vessel in lbs}
\]

Because the environment is corrosive and includes hydrogen,
an acceptable construction material is Inconel 617

\[ \text{Fm} := 3.9 \quad \text{See Seader Table 22.26 p. 576} \]

For horizontal vessels $9,000 < W < 920,000 \text{ lbs}$

\[ \text{Cv} := \exp\left[8.9552 - 0.2330 \cdot \ln(W) + 0.04333 \cdot \left(\ln(W)\right)^2\right] \]

For horizontal vessels $3 < D_i < 12 \text{ ft}$

\[ \text{Cpl} := 2005 \cdot D_i^{0.20294} \]

\[ \text{Cp} := \text{Fm} \cdot \text{Cv} + \text{Cpl} \]

\[ \text{Cp} = 6128372.74 \quad \text{f.o.b. purchase cost at a CE index = 500} \]

assuming an CE cost index of 574 for the year 2015:

\[ \text{Cost} := \text{Cp} \left(\frac{574}{500}\right) \]

\[ \text{Cost} = 7035371.9 \quad \text{Capital Cost for high temperature reactor} \]

**Catalyst Costs**

\[ r\text{Vol} := 300 \]

\[ \text{voidfrac} := 0.7 \quad \text{m}^3 \]

\[ \text{catdens} := 5193 \quad \text{kg/m}^3 \]

\[ \text{catweight} := r\text{Vol} \cdot (1 - \text{voidfrac}) \cdot \text{catdens} \]

\[ \text{catweight} = 4.674 \times 10^5 \]

\[ \text{catcost} := 17.31 \cdot \left(\frac{574}{402}\right) \cdot \text{catweight} \]

\[ \text{catcost} = 1.155 \times 10^7 \]

\[ \text{Totalcost} := \text{Cost} + \text{catcost} \]

\[ \text{Totalcost} = 18587014.38 \]

**Reactor R-101W - Low Temperature Reactor**

Weight of a pressure vessel, assuming constant shell and head thickness with 2:1 elliptical heads,

\[ \rho := 521.9 \quad \text{density of Inconel 617 (lb/ft}^3\text{)} \]

\[ D_i := 19.12 \quad \text{Diameter (ft)} \]
\[ L := 196.85 \text{ Length (ft)} \]
\[ P_0 := 1193 - 14.7 \text{ Operating Pressure (psig)} \]
\[ T := 511 \text{ Operating Temp (C)} \]
\[ S := \begin{cases} 15000 & \text{if } T \leq 750 \\ 14750 & \text{if } 750 < T \leq 800 \\ 14200 & \text{if } 800 < T \leq 850 \\ 13100 & \text{if } 850 < T \leq 900 \end{cases} \]
\[ E := 1 \]
\[ P_d := \begin{cases} 1.1 Po & \text{if } Po > 1000 \\ \exp \left[ 0.60608 + 0.91615 \ln(Po) + 0.0015655 \ln(Po)^2 \right] & \text{otherwise} \end{cases} \]
\[ t_p := \frac{(P_d \cdot Di)}{(2 \cdot S \cdot E - 1.2 \cdot P_d)} \]
\[ t_p = 0.871 \text{ increase thickness to 7/8" due to availability} \]
\[ t_s := \frac{7}{8} \text{ wall thickness in inches} \]
\[ W := \pi \cdot (ts + Di) \cdot (L + 0.8 \cdot Di) \cdot ts \cdot \rho \]
\[ W = 6085573.03 \text{ weight of vessel in lbs} \]

Because the environment is corrosive and includes hydrogen, an acceptable construction material is Inconel 617

\[ F_m = 3.9 \text{ See Seader Table 22.26 p. 576} \]

For horizontal vessels, \( 9,000 < W < 920,000 \text{ lbs} \)
\[ C_v := \exp \left[ 8.9552 - 0.2330 \ln(W) + 0.04333 \cdot (\ln(W))^2 \right] \]

For horizontal vessels, \( 3 < Di < 12 \text{ ft} \)
\[ C_{pl} := 2005 \cdot Di^{0.20294} \]
\[ C_p := F_m \cdot C_v + C_{pl} \]
\[ C_p = 31024334.72 \text{ f.o.b. purchase cost at a CE index = 500} \]

assuming an CE cost index of 574 for the year 2015:
\[ \text{Cost} := C_p \cdot \left( \frac{574}{500} \right) \]
\[ \text{Cost} = 35615936.26 \text{ Capital Cost for low temperature reactor} \]
Catalyst Costs

\[ r_{\text{Vol}} := 1600 \quad \text{m}^3 \]

\[ \text{voidfrac} := 0.7 \]

\[ \text{catdens} := 5168 \quad \text{kg m}^{-3} \]

\[ \text{catweight} := r_{\text{Vol}} \cdot (1 - \text{voidfrac}) \cdot \text{catdens} \]

\[ \text{catweight} = 2.481 \times 10^6 \]

\[ \text{catcost} := 17.31 \left( \frac{574}{402} \right) \cdot \text{catweight} \]

\[ \text{catcost} = 6.131 \times 10^7 \]

\[ \text{Totalcost} := \text{Cost} + \text{catcost} \]

\[ \text{Totalcost} = 96928100.94 \]

**Vessel V-100W**

Weight of a pressure vessel, assuming constant shell and head thickness with 2:1 elliptical heads,

\[ \rho := 521.9 \quad \text{density of Inconel 617} \quad \text{(lb/ft}^3) \]

\[ D_i := 11.50 \quad \text{diameter if V-100W} \quad \text{(ft)} \]

\[ L := 63.25 \quad \text{length of V-100W} \quad \text{(ft)} \]

\[ P_o := 1189 \quad \text{operating pressure of V-100W} \quad \text{(psig)} \]

\[ T := 35 \quad \text{operating temperature of V-100W} \quad \text{(Celsius)} \]

\[ S := \begin{cases} 15000 & \text{if } T \leq 750 \\ 14750 & \text{if } 750 < T \leq 800 \\ 14200 & \text{if } 800 < T \leq 850 \\ 13100 & \text{if } 850 < T \leq 900 \end{cases} \]

\[ E := 1 \]

\[ P_d := \begin{cases} 1.1 \cdot P_o & \text{if } P_o > 1000 \\ \exp[0.60608 + 0.91615 \cdot \ln(P_o) + 0.0015655 \cdot (\ln(P_o))^2] & \text{otherwise} \end{cases} \]

\[ t_p := \frac{(P_d \cdot D_i)}{(2 \cdot S \cdot E - 1.2 \cdot P_d)} \]

\[ t_p = 0.522 \quad \text{increase thickness to 9/16" due to availability} \]
\[ ts := \frac{9}{16} \quad \text{wall thickness in inches} \]

\[ W := \pi \cdot (ts + Di) \cdot (L + 0.8 \cdot Di) \cdot ts \cdot \rho \]

\[ W = 806000.69 \quad \text{weight of vessel in lbs} \]

Because the environment is corrosive and includes hydrogen, an acceptable construction material is Inconel 617

\[ Fm := 3.9 \quad \text{See Seader Table 22.26 p. 576} \]

For vertical vessels \(4,200 < W < 1,000,000\) lbs

\[ C_v := \exp \left[ 7.0132 - 0.18255 \cdot \ln(W) + 0.02297 \cdot (\ln(W))^2 \right] \]

For vertical vessels \(3 < Di < 12\) ft

\[ C_{pl} := 361.8 \cdot Di^{0.73960} \cdot L^{0.70684} \]

\[ C_p := Fm \cdot C_v + C_{pl} \]

\[ C_p = 66642.25 \quad \text{f.o.b. purchase cost at a CE index} = 500 \]

assuming an CE cost index of 574 for the year 2015:

\[ \text{Cost} := C_p \left( \frac{574}{500} \right) \]

\[ \text{Cost} = 76505.3 \quad \text{Capital Cost for V-100W} \]
Capital Cost of Turbine Unit Unit

\[ PCI := \frac{574}{500} \]  
Chemical Engineering Plant Cost Index - Adjusting to 2015 Values

**Air Compressor K-100T - Centrifugal Compressor**

\[ F_D := 1 \]  
Electric Motor

\[ F_M := 3.90 \]  
Inconel

\[ P_C := 1.882 \times 10^4 \]  
Horsepower

\[ C_B := \exp\left[7.5800 + 0.80\ln(P_C)\right] \]  
Base Cost (22.36)

\[ C_P := F_D F_M C_B PCI \]  
Purchase Cost (22.35)

\[ C_{K100} := C_P = 23049120.75 \]  
Purchase Cost of K-100T

**Air Compressor K-101T - Centrifugal Compressor**

\[ F_D := 1 \]  
Electric Motor

\[ F_M := 1 \]  
Carbon Steel

\[ P_C := 1.090 \times 10^5 \]  
Horsepower

\[ C_B := \exp\left[7.5800 + 0.80\ln(P_C)\right] \]  
Base Cost (22.36)

\[ C_P := F_D F_M C_B PCI \]  
Purchase Cost (22.35)

\[ C_{K101} := C_P = 24089914.54 \]  
Purchase Cost of K-101T

**Heat Exchanger E-100T - Fixed Head Shell and Tube**

\[ A := 4.147 \times 10^3 \]  
Heat Exchange Area (ft^2)

\[ P := 624.5 - 14.7 \]  
Shell Side Pressure (psia)

\[ C_A := \exp\left[11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2\right] \]  
Base Cost (22.40)

\[ a := 0 \]  
Shell - Carbon Steel

\[ b := 0 \]  
Tubes - Carbon Steel

\[ F_M := a + \left(\frac{A}{100}\right)^b \]  
Materials of Construction Factors

\[ F_L := 1.00 \]  
Tubes - 20 ft

\[ F_L := 1.00 \]  
Tube Length Correction
\[ F_p := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \] Pressure Factor (22.45)

\[ C_{\text{E100}} := F_p F_{\text{M}} F_{\text{L}} C_B P \text{C} \] Purchase Cost (22.43)

\[ C_{\text{E100}} := C_p = 35981.04 \] Purchase Cost of E-100T

**Air Compressor K-102T - Centrifugal Compressor**

\[ F_D := 1 \]

\[ F_{\text{M}} := 1 \]

\[ P_C := 1.125 \times 10^5 \]

\[ C_B := \exp \left[ 7.5800 + 0.80 \ln(P_C) \right] \] Base Cost (22.36)

\[ C_P := F_D F_{\text{M}} C_B P \text{C} \] Purchase Cost (22.35)

\[ C_{\text{K102}} := C_p = 24706776.06 \] Purchase Cost of K-102T

**Heat Exchanger E-101T - Fixed Head Shell and Tube**

\[ A := 1.066 \times 10^3 \] Heat Exchange Area (ft^2)

\[ P := 51.34 - 14.7 \] Shell Side Pressure (psia)

\[ C_B := \exp \left[ 11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2 \right] \] Base Cost (22.40)

\[ a := 0.00 \] Shell - Carbon Steel

\[ b := 0.00 \] Tubes - Carbon Steel

\[ F_{\text{M}} := a + \left( \frac{A}{100} \right)^b \] Material Factor (22.44)

\[ F_{\text{L}} := 1.00 \] Tubes - 20 ft Tube Length Correction

\[ F_{\text{D}} := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \] Pressure Factor (22.45)

\[ C_{\text{E101}} := F_p F_{\text{M}} F_{\text{L}} C_B P \text{C} \] Purchase Cost (22.43)

\[ C_{\text{E101}} := C_p = 67229.12 \] Purchase Cost of E-101T

**Air Compressor K-103T - Centrifugal Compressor**
\[ F_D := 1 \]
\[ F_M := 1 \]
\[ P_C := 1.128 \times 10^5 \]
\[ C_B := \exp\left[7.5800 + 0.80\ln(P_C)\right] \]
\[ C_D := F_D F_M C_B \cdot PCI \]
\[ C_{K103} := C_P = 24759469.81 \]

**Electric Motor**

**Carbon Steel**

**Horsepower**

**Base Cost (22.36)**

**Purchase Cost (22.35)**

**Heat Exchanger E-102T - Fixed Head Shell and Tube**

\[ A := 1.081 \times 10^4 \]
\[ P := 179 - 14.7 \]
\[ C_B := \exp\left[11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2\right] \]

**Heat Exchange Area (ft^2)**

**Shell Side Pressure (psia)**

**Base Cost (22.40)**

\[ a := 0 \]
\[ b := 0 \]

**Shell - Carbon Steel**

**Tubes - Carbon Steel**

**Materials of Construction Factors**

\[ F_M := a + \left(\frac{A}{100}\right)^b \]

**Material Factor (22.44)**

\[ F_L := 1.00 \]

**Tubes - 20 ft**

**Tube Length Correction**

\[ F_P := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left(\frac{P}{100}\right) + 0.0017 \left(\frac{P}{100}\right)^2 & \text{otherwise} \end{cases} \]

**Pressure Factor (22.45)**

\[ C_D := F_P F_M F_L C_B \cdot PCI \]
\[ C_{E102} := C_P = 69071.95 \]

**Purchase Cost (22.43)**

**Purchase Cost of E-102T**

**Air Compressor K-104T - Centrifugal Compressor**

\[ F_D := 1 \]
\[ F_M := 1 \]
\[ P_C := 1.140 \times 10^5 \]
\[ C_B := \exp\left[7.5800 + 0.80\ln(P_C)\right] \]
\[ C_D := F_D F_M C_B \cdot PCI \]
\[ C_{K104} := C_P = 24969965.48 \]

**Electric Motor**

**Carbon Steel**

**Horsepower**

**Base Cost (22.36)**

**Purchase Cost (22.35)**

**Purchase Cost of K-104T**
**Heat Exchanger E-103T - Fixed Head Shell and Tube**

\[ A = 1.095 \times 10^4 \]  
Heat Exchange Area (ft^2)

\[ P = 624.5 - 14.7 \]  
Shell Side Pressure (psia)

\[ C_B = \exp \left( 11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2 \right) \]  
Base Cost (22.40)

\[ a = 0 \]  
Shell - Carbon Steel

\[ b = 0 \]  
Tubes - Carbon Steel

\[ F_M = a + \left( \frac{A}{100} \right)^b \]  
Material Factor (22.44)

\[ F_L = 1.00 \]  
Tubes - 20 ft

\[ F_P = \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \]  
Pressure Factor (22.45)

\[ C_P = F_P F_M F_L C_B PCI \]  
Purchase Cost (22.43)

\[ C_{E-103} = C_P = 79448.77 \]  
Purchase Cost of E-103T

**Heat Exchanger E-104T - Fixed Head Shell and Tube**

\[ A = 1.475 \times 10^5 \]  
Heat Exchange Area (ft^2)

\[ P = 624.5 - 14.7 \]  
Shell Side Pressure (psia)

\[ C_B = \exp \left( 11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2 \right) \]  
Base Cost (22.40)

\[ a = 0 \]  
Shell - Carbon Steel

\[ b = 0 \]  
Tubes - Carbon Steel

\[ F_M = a + \left( \frac{A}{100} \right)^b \]  
Material Factor (22.44)

\[ F_L = 1.00 \]  
Tubes - 20 ft

\[ F_P = \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left( \frac{P}{100} \right) + 0.0017 \left( \frac{P}{100} \right)^2 & \text{otherwise} \end{cases} \]  
Pressure Factor (22.45)

\[ C_P = F_P F_M F_L C_B PCI \]  
Purchase Cost (22.43)
**Heat Exchanger E-105T - Fixed Head Shell and Tube**

\[ A := 1.247 \times 10^5 \]
\[ P := 624.5 - 14.7 \]

\[ C_B := \exp\left(11.0545 - 0.9228 \ln(A) + 0.09861 \ln(A)^2\right) \]

\[ a := 0 \quad \text{Shell - Carbon Steel} \]
\[ b := 0 \quad \text{Tubes - Carbon Steel} \]

\[ F_M := a + \left(\frac{A}{100}\right)^b \]

\[ F_T := 1.00 \quad \text{Tubes - 20 ft} \]

\[ F_P := \begin{cases} 1 & \text{if } P \leq 114.7 \\ 0.9803 + 0.018 \left(\frac{P}{100}\right) + 0.0017 \left(\frac{P}{100}\right)^2 & \text{otherwise} \end{cases} \]

\[ C_B := F_P F_M F_L C_B PCI \]

\[ C_{E105} := C_P = 1307754.66 \]

**Pump P-100T - Centrifugal Pump**

Centrifugal Pump

\[ Q := 2193.1 \]
\[ H := 5072 \]

\[ S := Q - H^{0.5} \]

\[ C_B := \exp\left(9.7171 - 0.6019 \ln(S) + 0.0519 \ln(S)^2\right) \]

\[ F_M := 2.00 \quad \text{Stainless Steel} \]

\[ F_T := 8.90 \]

\[ C_B := F_T F_M C_B PCI \]

\[ C_{P100} := C_P = 4.245 \times 10^5 \]

**Electric Motor**

\[ \rho := 8.25 \]
\[ P_B := 3721 \]

\[ \eta_P := -0.316 + 0.24015 \ln(Q) - 0.01199 (\ln(Q))^2 \]

---

*CE104 := CP = 1656517.92*

*CE105 := CP = 1307754.66*
\[ \eta_M := 0.80 + 0.0319 \ln(P_B) \]  

Efficiency of Motor (22.18)

\[ P_C := \frac{Q \cdot H \cdot \rho}{33000 \cdot \eta_P \cdot \eta_M} \]  
Pump Power Consumption (22.16)

\[ C_B := \exp \left(5.8259 + 0.13141 \cdot \ln(P_C) + 0.053255 \ln(P_C)^2 + 0.028628 \cdot \ln(P_C)^3 - 0.0035549 \ln(P_C)^4\right) \]  
Base Cost (22.19)

\[ F_T := 1.8 \quad \text{Explosion Proof Motor} \]  
Motor Type Factor

\[ C_M := F_T \cdot C_B \cdot P_C \]  
Purchase Cost (22.20)

\[ C_{M101} := C_P = 6.308 \times 10^4 \]  
Purchase Cost of M-100
Appendix G: Break Even and Selling Price Calculation Sheets

EES Code for determining break even price:

\[ m_.c = 2 \times 6445 \text{[ton/day]} \times \text{convert(ton/day,kg/s)} \]
\[ c\_coal = 55\text{[$/ton]} \times \text{convert(1/ton,1/kg)} \]
\[ W\_\text{out} = 1000\text{[MW]} \]

"Cost Calculations"

\[ i=0.04 \]
\[ i\_\text{eff} = ((1+(i/m))^m)-1 \]
\[ r\_n=0.04 \]
\[ r\_n\_\text{coal} = 0.039 \]
\[ UF = 0.91 \]
\[ n = 22 \]
\[ m = 12 \]
\[ k = (1+r\_n)/(1+i\_\text{eff}) \]
\[ \text{CELF} = ((k^\star(1-k^\star n))/(1-k))^\star \text{CRF} \]
\[ \text{CRF} = ((i\_\text{eff}^\star(1+i\_\text{eff})^\star n)/(1+i\_\text{eff})^\star n-1) \]
\[ C\_\text{coal} = c\_\text{coal} \times m\_.c \times \text{convert($/s,$/hr)} \]
\[ C\_\text{air} = 0 \text{[$/hr]} \]

\[ \text{OM\_L} = 909077012 \]

\[ \text{TRR\_L} = (\text{OM\_L})/(8776\text{[hr]}^\star UF) + C\_\text{coal} \]

\[ \text{Cost\_elec} = ((\text{TRR\_L}) / W\_\text{out})^\star \text{convert($/MW-hr, cents/kW-hr)} \]

"Purchased Equipment Costs"

\[ \text{PEC\_total} = 2962E6 \text{[$]} \]

EES Solution:

\[ \text{CELF} = 1.521 \]
\[ \text{CRF} = 0.06969 \]
\[ \dot{C}\_\text{coal} = 29540 \text{[$/hr]} \]
\[ i = 0.04 \]
\[ k = 0.9993 \]
\[ \dot{m}\_c = 135.3 \text{[kg/s]} \]
\[ \text{OM\_L} = 9.091E+03 \text{[$]} \]
\[ r\_n = 0.04 \]
\[ \text{TRP\_L} = 143371 \text{[$/hr]} \]
\[ W\_\text{out} = 1000 \text{[MW]} \]

\[ \text{Cost\_elec} = 14.34 \text{[cents/kW-hr]} \]
\[ C\_\text{air} = 0 \text{[$/hr]} \]
\[ c\_\text{coal} = 0.05063 \text{[$/kg]} \]
\[ i\_\text{eff} = 0.04074 \]
\[ m = 12 \]
\[ n = 22 \]

\[ \text{PEC\_total} = 2.962E+09 \text{[$]} \]
\[ r\_n\_\text{coal} = 0.039 \]

\[ UF = 0.91 \]
### IGCC Coal Power Plant

#### FINANCIAL SPREADSHEET:

**PRODUCT:** Electricity  
**HURDLE RATE:** 12.0%  
**LIFE, YRS.:** 22  
**CAPACITY:** 8000 MM PY  
**INFLATION, PRICE:** 4%  
**TAX RATE:** 39%  
**LOCATION:** Michigan  
**INFLTN, VAR.COST:** 4%  
**BOOK/TAX DEPRECIATION:** 9.09% [11 YR. SL]  
**CAPITAL ($000):** 68,000  
**INFL,FIXD.&CPTL.CST:** 4%  
**LAND ($000):** 200  
**BEGINNING YEAR:** 2015  
**FIXED COST, (CURRENT YEAR) ($000):** 15,129  

#### P & L

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#### BALANCE SHEET

**FIXED INVESTMENT** | 165,260 | 66,160 |
| WORKING CAPITAL (18%) | --- | --- |
| **TOTAL ASSETS** | 165,260.00 | 66,160.00 |
| **LESS DEPR RESERVE** | --- | --- |
| **CAPITAL EMPLOYED** | 192,980.71 | 251,680.16 |

**Minnesota**

**CAPITAL EMPLOYED**

**INFLATION, PRICE**

**TAX RATE**

**BOOK/TAX DEPRECIATION**

**LOCATION:** Michigan

**INFLTN, VAR.COST:** 4%

**START-UP COST**

**LAND ($000):** 200

**BEGINNING YEAR:** 2015

**FIXED COST, (CURRENT YEAR) ($000):** 15,129
### CASH FLOW SCHEDULE

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</tr>
<tr>
<td>WORK’G CAPTL (CHANGES)</td>
<td>---</td>
<td>---</td>
<td>130,562.5</td>
<td>65,281.26</td>
<td>75,726.27</td>
<td>10,862.8</td>
<td>11,297.31</td>
<td>11,749.21</td>
<td>12,219.18</td>
<td>12,707.94</td>
<td>13,216.26</td>
<td>13,744.91</td>
<td>14,294.71</td>
<td>14,866.5</td>
<td>15,461.15</td>
<td>16,079.6</td>
</tr>
<tr>
<td>TOTAL CASH OUT</td>
<td>16,520</td>
<td>49,640</td>
<td>132,603</td>
<td>65,281</td>
<td>75,726</td>
<td>10,863</td>
<td>11,297</td>
<td>11,749</td>
<td>12,219</td>
<td>12,708</td>
<td>13,216</td>
<td>13,745</td>
<td>14,295</td>
<td>14,866</td>
<td>15,461</td>
<td>16,080</td>
</tr>
</tbody>
</table>

| CASH IN | | | | | | | | | | | | | | | | | |
| PROFIT AFTER TAX | (732) | (3,155) | 21,989 | 19,097 | 31,633 | 33,058 | 34,541 | 36,084 | 37,688 | 39,356 | 41,060 | 42,895 | 44,771 | 50,493 | 52,523 | 54,634 | 56,629 |
| DEPRECIATION (11YR. SL) | --- | --- | 6181.82 | 6181.82 | 6181.82 | 6181.82 | 6181.82 | 6181.82 | 6181.82 | 6181.82 | 6181.82 | 0.00 | 0.00 | 0.00 | 0.00 |
| RETURN OF WORKG CAPITAL (1) | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | --- | 434,792 |
| TOTAL CASH IN | (732) | (3,155) | 28,171 | 25,279 | 37,814 | 39,240 | 40,723 | 42,265 | 43,869 | 45,537 | 47,272 | 49,076 | 50,953 | 50,493 | 52,523 | 54,634 | 491,621 |
| CUMULATIVE CASH IN | -732.09 | -3887.26 | 24283.59 | 40952.47 | 87376.80 | 126617.03 | 167340.18 | 209605.59 | 253474.94 | 299012.38 | 346284.65 | 395361.13 | 446313.99 | 496807.38 | 549330.27 | 603963.83 | 3333333.33 |
Appendix H: Heat Exchanger Design Sheets

The following sheets show the equations and method used to calculation heat transfer coefficients and heat transfer areas for the heat exchangers in the IGCC power plant. The method used was adopted from Warren Seider’s ‘Product and Process Design Principles.’
Heat Exchanger E-100A Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[
Q := 3.819 \times 10^7 \quad \text{Heat Exchanger Duty (W)}
\]

**Tube Side Properties**

- \(D_i := 0.0135\) Tube Inside Diameter (m)
- \(k := 0.6279\) Thermal Conductivity of Tube Side Fluid (W/m-K)
- \(C_p := 4316\) Mass Heat Capacity of Tube Side Fluid (J/kg-K)
- \(F := \frac{590.0}{1074}\) Mass Flow Rate of 1 Tube (kg/s)
- \(A := \pi \left( \frac{D_i}{2} \right)^2\) Cross Sectional Area for Flow (m^2)
- \(G := \frac{F}{A}\) Fluid Mass Velocity (kg/m^2-s)
- \(\mu := 6.933 \times 10^{-4}\) Viscosity (Pa-s)
- \(n := 0.3\) Constant: \(n = 0.4\) if heating, 0.3 if cooling

\[
h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i \cdot G}{\mu} \right)^{0.8} \left( \frac{C_p \cdot h}{k} \right)^n \quad \text{Sieder-Tate Equation (18.13)}
\]

\[
h_i = 1.354 \times 10^4 \quad \text{Convective Heat Transfer Coefficient (W/m^2-K)}
\]

**Shell Side Properties**

- \(D_o := 0.0191\) Tube Outside Diameter (m)
- \(D_s := 0.9398\) Shell Diameter (m)
- \(k := 0.03076\) Thermal Conductivity of Shell Side Fluid (W/m-K)
- \(C_p := 1022\) Mass Heat Capacity of Fluid (J/kg-K)
- \(F := 308.9\) Mass Flow Rate of Fluid (kg/s)
- \(A := \pi \left( \frac{D_s}{2} \right)^2 - \left( \frac{D_o}{2} \right)^2 \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \) Cross Sectional Area for Flow (m^2)
- \(G := \frac{F}{A}\) Fluid Mass Velocity (kg/m^2-s)
- \(\mu := 2.203 \times 10^{-5}\) Viscosity (Pa-s)
- \(C := 0.36\) \(n_s := 0.55\) Kern Correlation Constants
\[ h_o := C \cdot \frac{k}{D_o} \left( \frac{D_o \cdot G}{\mu} \right)^n \left( \frac{C_p \cdot \mu}{k} \right)^{0.333} \]

Kern Correlation (18.21)

Convective Heat Transfer Coefficient (W/m^2-K)

**Inlet and Outlet Temperatures**

<table>
<thead>
<tr>
<th>Side</th>
<th>Temperature</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hot Side</td>
<td>T(_{hi}) := 156.0</td>
</tr>
<tr>
<td></td>
<td>T(_{ho}) := 35.0</td>
</tr>
<tr>
<td>Cold Side</td>
<td>T(_{ci}) := 30.0</td>
</tr>
<tr>
<td></td>
<td>T(_{co}) := 45.0</td>
</tr>
</tbody>
</table>

\[ T_{lm} := \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right) \]

\[ \ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right) \]

**Materials of Construction**

Shell Side - Carbon Steel
Tube Side - Carbon Steel

k := 46.6

Thermal Conductivity of Tubes (W/m-K)

L := 6.096

Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi \cdot D_i \cdot L \quad A_o := \pi \cdot D_o \cdot L \]

\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ A_m := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \]

\[ t_w := D_o - D_i \]

\[ U := \frac{1}{h_o + \frac{t_w \cdot A_o}{k \cdot A_m} + \frac{A_o}{h_i \cdot A_i}} \]

\[ U = 692.496 \]

Heat Transfer Coefficient (W/m^2-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 8.067 \quad S = 0.119 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.95 \]

Ft Correction Factor for 2-4 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]

\[ A = 1.698 \times 10^3 \]

Heat Exchange Area Needed (m^2)
Heat Exchanger E-101A Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q = 1.573 \times 10^7 \]

**Tube Side Properties**

- \( D_i := 0.0135 \)
- \( k := 0.030 \)
- \( C_p := 1037 \)
- \( F := \frac{70.89}{1074} \)
- \( A := \pi \left( \frac{D_i}{2} \right)^2 \)
- \( G := \frac{F}{A} \)
- \( \mu := 2.223 \times 10^{-5} \)
- \( n := 0.3 \)
- \( h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \)
- \( h_i = 1.076 \times 10^3 \)

**Shell Side Properties**

- \( D_o := 0.0191 \)
- \( D_s := 0.9398 \)
- \( k := 0.03371 \)
- \( C_p := 1033.5 \)
- \( F := 314.4 \)
- \( A := \pi \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \)
- \( G := \frac{F}{A} \)
- \( \mu := 2.389 \times 10^{-5} \)
- \( C := 0.36 \quad n_s := 0.55 \)

Heat Exchanger Duty (W)
Tube Inside Diameter (m)
Thermal Conductivity of Tube Side Fluid (W/m-K)
Mass Heat Capacity of Tube Side Fluid (J/kg-K)
Mass Flow Rate of 1 Tube (kg/s)
Cross Sectional Area for Flow (m^2)
Fluid Mass Velocity (kg/m^2-s)
Viscosity (Pa-s)
Constant: \( n = 0.4 \) if heating, 0.3 if cooling
Sieder-Tate Equation (18.13)
Convective Heat Transfer Coefficient (W/m^2-K)
Tube Outside Diameter (m)
Shell Diameter (m)
Thermal Conductivity of Shell Side Fluid (W/m-K)
Mass Heat Capacity of Fluid (J/kg-K)
Mass Flow Rate of Fluid (kg/s)
Cross Sectional Area for Flow (m^2)
Fluid Mass Velocity (kg/m^2-s)
Viscosity (Pa-s)
Kern Correlation Constants
\[ h_o := C \frac{k}{D_o} \left( \frac{D_o G}{\mu} \right)^n \left( \frac{C_p \mu}{k} \right)^{0.333} \]

Kern Correlation (18.21)

Convective Heat Transfer Coefficient (W/m^2-K)

**Inlet and Outlet Temperatures**

Hot Side
\[ T_{hi} := 162.6 \]
\[ T_{ho} := 114.2 \]

Cold Side
\[ T_{ci} := -94.83 \]
\[ T_{co} := 100.0 \]

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**
- Shell Side - Carbon Steel
- Tube Side - Stainless Steel

\[ k := 16.3 \]

Thermal Conductivity of Tubes (W/m-K)

\[ L := 6.096 \]

Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi D_i L \quad A_o := \pi D_o L \quad A_m := \frac{\pi L (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \]

\[ t_w := D_o - D_i \]

\[ U := \frac{1}{\frac{1}{h_o} + \frac{t_w A_o}{k A_m} + \frac{A_o}{h_i A_i}} \]

\[ U = 350.622 \]

Heat Transfer Coefficient (W/m^2-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 0.248 \]

\[ S = 0.757 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.95 \]

Ft Correction Factor for 2-4 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U F_T T_{lm}} \]

\[ A = 388.848 \]

Heat Exchange Area Needed (m^2)
Heat Exchanger E-102A Design

Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 1.257 \times 10^7 \]  

**Tube Side Properties**

- Tube Inside Diameter (m)
- Thermal Conductivity of Tube Side Fluid (W/m-K)
- Mass Heat Capacity of Tube Side Fluid (J/kg-K)
- Mass Flow Rate of 1 Tube (kg/s)
- Cross Sectional Area for Flow (m^2)
- Fluid Mass Velocity (kg/m^2-s)
- Viscosity (Pa-s)
- Constant: \( n = 0.4 \) if heating, \( 0.3 \) if cooling
- Sieder-Tate Equation (18.13)

\[ h_i := 0.023 \cdot \frac{k}{D_i} \left( \frac{D_i \cdot G}{\mu} \right)^{0.8} \cdot \left( \frac{C_p \cdot \mu}{k} \right)^n \]

\[ h_i = 2.848 \times 10^3 \]  

**Shell Side Properties**

- Tube Outside Diameter (m)
- Shell Diameter (m)
- Thermal Conductivity of Shell Side Fluid (W/m-K)
- Mass Heat Capacity of Fluid (J/kg-K)
- Mass Flow Rate of Fluid (kg/s)
- Cross Sectional Area for Flow (m^2)
- Fluid Mass Velocity (kg/m^2-s)
- Viscosity (Pa-s)
- Kern Correlation Constants

\[ A := \pi \left( \frac{D_s}{2} \right)^2 - \left( \frac{D_s}{2} \right)^2 \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \]

\[ \frac{F}{A} := \frac{D_s}{2} \]

\[ \mu := 2.389 \times 10^{-5} \]

\[ C := 0.36 \quad n_a := 0.55 \]
\[ h_o := C \cdot \frac{k}{D_o} \left( \frac{D_o \cdot G}{\mu} \right)^n \left( \frac{C_p \cdot \mu}{k} \right)^{0.333} \quad \text{Kern Correlation (18.21)} \]

Convective Heat Transfer Coefficient (W/m^2-K)

\[ h_o = 875.589 \]

**Inlet and Outlet Temperatures**

<table>
<thead>
<tr>
<th>Hot Side</th>
<th>Cold Side</th>
</tr>
</thead>
<tbody>
<tr>
<td>( T_{hi} := 122.0 )</td>
<td>( T_{ci} := 0.0 )</td>
</tr>
<tr>
<td>( T_{ho} := 35.0 )</td>
<td>( T_{co} := 110.5 )</td>
</tr>
</tbody>
</table>

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln\left(\frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}}\right)} \]

**Materials of Construction**

- Shell Side - Carbon Steel
- Tube Side - Stainless Steel

\[ k := 16.3 \quad \text{Thermal Conductivity of Tubes (W/m-K)} \]

\[ L := 6.096 \quad \text{Tube length (m)} \]

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi \cdot D_i \cdot L \quad A_o := \pi \cdot D_o \cdot L \quad A_m := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln\left(\frac{D_o}{D_i}\right)} \quad t_w := D_o - D_i \]

\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ U := \frac{1}{\frac{1}{h_o} + \frac{t_w \cdot A_o}{k \cdot A_m} + \frac{A_o}{h_i \cdot A_i}} \]

\[ U = 488.877 \quad \text{Heat Transfer Coefficient (W/m^2-K)} \]

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 0.787 \quad S = 0.906 \]

From the \( R \) and \( S \) Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.95 \quad \text{Ft Correction Factor for 2-4 Shell Pass Heat Exchanger} \]

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]

\[ A = 1.282 \times 10^3 \quad \text{Heat Exchange Area Needed (m^2)} \]
Heat Exchanger E-100G Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 1.542 \times 10^8 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

- \( D_i := 0.0135 \)  
  Tube Inside Diameter (m)
- \( k := 0.1261 \)  
  Thermal Conductivity of Tube Side Fluid (W/m-K)
- \( C_p := 1622 \)  
  Mass Heat Capacity of Tube Side Fluid (J/kg-K)
- \( F := \frac{141.8}{1074} \)  
  Mass Flow Rate of 1 Tube (kg/s)
- \( A := \pi \left( \frac{D_i}{2} \right)^2 \)  
  Cross Sectional Area for Flow (m^2)
- \( G := F / A \)  
  Fluid Mass Velocity (kg/m^2-s)
- \( \mu := 3.865 \times 10^{-5} \)  
  Viscosity (Pa-s)
- \( n := 0.3 \)  
  Constant: \( n = 0.4 \) if heating, 0.3 if cooling
- \( h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \)  
  Sieder-Tate Equation (18.13)
- \( h_i = 4.442 \times 10^3 \)  
  Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

- \( D_o := 0.0191 \)  
  Tube Outside Diameter (m)
- \( D_s := 0.9398 \)  
  Shell Diameter (m)
- \( k := 0.6279 \)  
  Thermal Conductivity of Shell Side Fluid (W/m-K)
- \( C_{ps} := 4224 \)  
  Mass Heat Capacity of Fluid (J/kg-K)
- \( F := 2432 \)  
  Mass Flow Rate of Fluid (kg/s)
- \( A := \pi \left( \frac{D_s}{2} \right)^2 - \left( 1022 \pi \left( \frac{D_o}{2} \right)^2 \right) \)  
  Cross Sectional Area for Flow (m^2)
- \( G := F / A \)  
  Fluid Mass Velocity (kg/m^2-s)
- \( \mu := 6.596 \times 10^{-4} \)  
  Viscosity (Pa-s)
- \( C := 0.36 \)  
  Kern Correlation Constants
\[ h_0 = 1.49 \times 10^4 \]  

Convective Heat Transfer Coefficient \((W/m^2-K)\)

**Inlet and Outlet Temperatures**

<table>
<thead>
<tr>
<th>Hot Side</th>
<th>Cold Side</th>
</tr>
</thead>
<tbody>
<tr>
<td>( T_{hi} := 700 )</td>
<td>( T_{ci} := 30.0 )</td>
</tr>
<tr>
<td>( T_{ho} := 35.0 )</td>
<td>( T_{co} := 45.0 )</td>
</tr>
</tbody>
</table>

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**

- Shell Side: Carbon Steel
- Tube Side: Inconel

\[ k_\text{tubes} := 19.76 \quad \text{Thermal Conductivity of Tubes (W/m-K)} \]

\[ L := 6.096 \quad \text{Tube length (m)} \]

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi \cdot D_i \cdot L \quad A_o := \pi \cdot D_o \cdot L \quad A_m := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \quad t_w := D_o - D_i \]

\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ U := \frac{1}{\frac{1}{h_0} + \frac{t_w \cdot A_o}{k \cdot A_m} + \frac{A_o}{h_i \cdot A_i}} \]

\[ U = 1.387 \times 10^3 \quad \text{Heat Transfer Coefficient (W/m}^2\text{-K)} \]

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 44.333 \quad S = 0.022 \]

From the \( R \) and \( S \) Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 1.0 \quad \text{Ft Correction Factor for 2-4 Shell Pass Heat Exchanger} \]

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]

\[ A = 833.933 \quad \text{Heat Exchange Area Needed (m}^2\text{)} \]

\[ A_{ft} := A \cdot 10.76 = 8.973 \times 10^3 \quad \text{Heat Exchange Area Needed (ft}^2\text{)} \]
Heat Exchanger E-100W Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\( Q := 1.096 \times 10^8 \)  
Heat Exchanger Duty (W)

**Tube Side Properties**

- \( D_i := 0.0135 \)  
  Tube Inside Diameter (m)
- \( k := 0.1150 \)  
  Thermal Conductivity of Tube Side Fluid (W/m-K)
- \( C_p := 2040 \)  
  Mass Heat Capacity of Tube Side Fluid (J/kg-K)
- \( F := \frac{219.1}{1074} \)  
  Mass Flow Rate of 1 Tube (kg/s)
- \( A := \pi \left( \frac{D_i}{2} \right)^2 \)  
  Cross Sectional Area for Flow (m²)
- \( G := \frac{F}{A} \)  
  Fluid Mass Velocity (kg/m²-s)
- \( \mu := 2.645 \times 10^{-5} \)  
  Viscosity (Pa-s)
- \( n := 0.3 \)  
  Constant: \( n = 0.4 \) if heating, \( 0.3 \) if cooling

\[ h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\( h_i = 7.637 \times 10^3 \)  
Convective Heat Transfer Coefficient (W/m²-K)

**Shell Side Properties**

- \( D_o := 0.0191 \)  
  Tube Outside Diameter (m)
- \( D_s := 0.9398 \)  
  Shell Diameter (m)
- \( k := 0.3264 \)  
  Thermal Conductivity of Shell Side Fluid (W/m-K)
- \( C_p := 3121 \)  
  Mass Heat Capacity of Fluid (J/kg-K)
- \( F := 40.89 \)  
  Mass Flow Rate of Fluid (kg/s)
- \( A := \pi \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \)  
  Cross Sectional Area for Flow (m²)
- \( G := \frac{F}{A} \)  
  Fluid Mass Velocity (kg/m²-s)
- \( \mu := 4.043 \times 10^{-4} \)  
  Viscosity (Pa-s)
- \( C := 0.36 \)  
  \( n := 0.55 \)  
  Kern Correlation Constants
\[
    h_o := C \frac{k}{D_o} \left( \frac{D_o G}{\mu} \right)^n \left( \frac{C_p \mu}{k} \right)^{0.333}
\]

Kern Correlation (18.21)

Convective Heat Transfer Coefficient (W/m^2-K)

Inlet and Outlet Temperatures

<table>
<thead>
<tr>
<th>Side</th>
<th>Temperature</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hot Side</td>
<td>646.0</td>
</tr>
<tr>
<td>Cold Side</td>
<td>30</td>
</tr>
</tbody>
</table>

\[
    T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)}
\]

Materials of Construction

- Shell Side: Carbon Steel
- Tube Side: Inconel 617

\[
    k := 19.76 \quad \text{Thermal Conductivity of Tubes (W/m-K)}
\]

\[
    L := 6.096 \quad \text{Tube length (m)}
\]

Overall Heat Transfer Coefficient and Ft Correction Factor

\[
    A_i := \pi D_i L \quad A_o := \pi D_o L
\]

\[
    A_i = 0.259 \quad A_o = 0.366
\]

\[
    U := \frac{1}{h_o + \frac{t_w A_o}{k A_m} + \frac{A_o}{h_i A_i}}
\]

\[
    U = 667.8 \quad \text{Heat Transfer Coefficient (W/m^2-K)}
\]

\[
    R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}}
\]

\[
    R = 1.447 \quad S = 0.276
\]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[
    F_T := 1.0 \quad \text{Ft Correction Factor for 2-4 Shell Pass Heat Exchanger}
\]

Heat Transfer Area Needed

\[
    A := \frac{Q}{U F_T T_{lm}}
\]

\[
    A = 403.426 \quad \text{Heat Exchange Area Needed (m^2)}
\]

\[
    A_{ft} := A \cdot 10.76 = 4.341 \times 10^3 \quad \text{Heat Exchange Area Needed (ft^2)}
\]
**Heat Exchanger E-101W Design**

Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q = 3.058 \times 10^8 \]

**Heat Exchanger Duty (W)**

**Tube Side Properties**

\[ D_i = 0.0135 \]

Tube Inside Diameter (m)

\[ k = 0.1356 \]

Thermal Conductivity of Tube Side Fluid (W/m-K)

\[ C_p = 2119 \]

Mass Heat Capacity of Tube Side Fluid (J/kg-K)

\[ F = \frac{219.1}{1074} \]

Mass Flow Rate of 1 Tube (kg/s)

\[ A = \pi \left( \frac{D_i}{2} \right)^2 \]

Cross Sectional Area for Flow (m^2)

\[ G = \frac{F}{A} \]

Fluid Mass Velocity (kg/m^2-s)

\[ \mu = 2.574 \times 10^{-5} \]

Viscosity (Pa-s)

\[ n = 0.3 \]

Constant: \( n = 0.4 \) if heating, 0.3 if cooling

\[ h_i = 0.023 \frac{k}{D_i} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \]

Sieder-Tate Equation (18.13)

\[ h_i = 8.788 \times 10^3 \]

Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

\[ D_o = 0.0191 \]

Tube Outside Diameter (m)

\[ D_s = 0.9398 \]

Shell Diameter (m)

\[ k = 0.3264 \]

Thermal Conductivity of Shell Side Fluid (W/m-K)

\[ C_p = 3121 \]

Mass Heat Capacity of Fluid (J/kg-K)

\[ F = 107.0 \]

Mass Flow Rate of Fluid (kg/s)

\[ A = \pi \left( \frac{D_s}{2} \right)^2 - \left( 1022 \pi \left( \frac{D_o}{2} \right)^2 \right) \]

Cross Sectional Area for Flow (m^2)

\[ G = \frac{F}{A} \]

Fluid Mass Velocity (kg/m^2-s)

\[ \mu = 4.043 \times 10^{-4} \]

Viscosity (Pa-s)

\[ C = 0.36 \quad n_o = 0.55 \]

Kern Correlation Constants
\[ h_o := C \cdot \frac{k}{D_o} \left( \frac{D_o G}{\mu} \right)^n \left( \frac{C_p \mu}{k} \right)^{0.333} \]  
Kern Correlation (18.21)

\[ h_o = 1.738 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Inlet and Outlet Temperatures**

Hot Side
- \( T_{hi} := 511.2 \)
- \( T_{ho} := 35 \)

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

Cold Side
- \( T_{ci} := 30 \)
- \( T_{co} := 285 \)

**Materials of Construction**

Shell Side - Carbon Steel  
Tube Side - Carbon Steel

\( k_s = 19.76 \)  
Thermal Conductivity of Tubes (W/m-K)

\( L = 6.096 \)  
Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi D_i L \quad A_o := \pi D_o L \quad A_m := \frac{\pi L (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \quad t_w := D_o - D_i \]

\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ U := \frac{1}{h_o + \frac{t_w A_o}{h_i A_i} + \frac{A_o}{h_i A_i}} \]

\[ U = 932.936 \]  
Heat Transfer Coefficient (W/m^2-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 1.867 \quad S = 0.53 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.85 \]  
Ft Correction Factor for 3-6 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]

\[ A = 6.646 \times 10^3 \]  
Heat Exchange Area Needed (m^2)

\[ A_{ft} := A \cdot 10.76 = 7.151 \times 10^4 \]  
Heat Exchange Area Needed (ft^2)
Heat Exchanger E-100T Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 7.025 \times 10^6 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

\[ D_i := 0.0135 \]  
Tube Inside Diameter (m)

\[ k := 0.1927 \]  
Thermal Conductivity of Tube Side Fluid (W/m-K)

\[ C_p := 6476 \]  
Mass Heat Capacity of Tube Side Fluid (J/kg-K)

\[ F := \frac{26.27}{1074} \]  
Mass Flow Rate of 1 Tube (kg/s)

\[ A := \pi \left( \frac{D_i}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ g := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 1.232 \times 10^{-5} \]  
Viscosity (Pa-s)

\[ n := 0.3 \]  
Constant: n = 0.4 if heating, 0.3 if cooling

\[ h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\[ h_i = 4.162 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

\[ D_o := 0.0191 \]  
Tube Outside Diameter (m)

\[ D_s := 0.9398 \]  
Shell Diameter (m)

\[ k := 0.0345 \]  
Thermal Conductivity of Shell Side Fluid (W/m-K)

\[ C_p := 4223 \]  
Mass Heat Capacity of Fluid (J/kg-K)

\[ F := 40.89 \]  
Mass Flow Rate of Fluid (kg/s)

\[ A := \pi \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \]  
Cross Sectional Area for Flow (m^2)

\[ g := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 1.592 \times 10^{-5} \]  
Viscosity (Pa-s)

\[ C := 0.36 \]  
Kern Correlation Constants

\[ n := 0.55 \]  
Kern Correlation Constants
Convective Heat Transfer Coefficient (W/m²-K)

\[ h_0 = \frac{C \cdot k}{D_o \cdot \left( \frac{D_o \cdot G}{\mu} \right)^n \left( \frac{C_p \cdot \mu}{k} \right)^{0.333}} \]

Kern Correlation (18.21)

**Inlet and Outlet Temperatures**

**Hot Side**
- \( T_{hi} := 200.0 \)
- \( T_{ho} := 170.0 \)

**Cold Side**
- \( T_{ci} := 118.2 \)
- \( T_{co} := 160.0 \)

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**
- Shell Side - Carbon Steel
- Tube Side - Carbon Steel

- \( k := 46.6 \) Thermal Conductivity of Tubes (W/m-K)
- \( L := 6.096 \) Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi \cdot D_i \cdot L \quad A_o := \pi D_o \cdot L \quad A_m := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \quad t_w := D_o - D_i \]

\[ U := \frac{1}{h_0 + \frac{t_w \cdot A_o}{k \cdot A_m} + \frac{A_o}{h_i A_i}} \]

\[ U = 409.532 \] Heat Transfer Coefficient (W/m²-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 0.718 \quad S = 0.511 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\( F_T := 0.975 \) Ft Correction Factor for 2-4 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]
Heat Exchanger E-101T Design

Heat Exchanger Specifications - Based on Table 18.6

One Pass - Triangular Pitch

Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 7.686 \times 10^7 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

\[ D_i := 0.0135 \]  
Tube Inside Diameter (m)

\[ k := 0.0320 \]  
Thermal Conductivity of Tube Side Fluid (W/m-K)

\[ C_p := 1034 \]  
Mass Heat Capacity of Tube Side Fluid (J/kg-K)

\[ F := \frac{472.9}{1074} \]  
Mass Flow Rate of 1 Tube (kg/s)

\[ A := \pi \left( \frac{D_i}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ \mu := 2.267 \times 10^{-5} \]  
Viscosity (Pa-s)

\[ n := 0.4 \]  
Constant: \( n = 0.4 \) if heating, 0.3 if cooling

\[ h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i \cdot G}{\mu} \right)^{0.8} \left( \frac{C_p \cdot \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\[ h_i = 4.929 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

\[ D_o := 0.0191 \]  
Tube Outside Diameter (m)

\[ D_s := 0.9398 \]  
Shell Diameter (m)

\[ k := 0.6279 \]  
Thermal Conductivity of Shell Side Fluid (W/m-K)

\[ C_p := 4223 \]  
Mass Heat Capacity of Fluid (J/kg-K)

\[ F := 1213 \]  
Mass Flow Rate of Fluid (kg/s)

\[ A := \pi \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \]  
Cross Sectional Area for Flow (m^2)

\[ \mu := 6.956 \times 10^{-4} \]  
Viscosity (Pa-s)

\[ C := 0.36 \quad n_a := 0.55 \]  
Kern Correlation Constants
\[ h_o := C \cdot \frac{k}{D_o} \left( \frac{D_o \cdot G}{\mu} \right)^n \left( \frac{C_p \cdot \mu}{k} \right)^{0.333} \]

Kern Correlation (18.21)

Convective Heat Transfer Coefficient (W/m^2-K)

**Inlet and Outlet Temperatures**

**Hot Side**
\[ T_{hi} := 192.3 \]
\[ T_{ho} := 35.0 \]

**Cold Side**
\[ T_{ci} := 30.0 \]
\[ T_{co} := 45.0 \]

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**

Shell Side - Carbon Steel
Tube Side - Carbon Steel

\[ k := 46.6 \quad \text{Thermal Conductivity of Tubes (W/m-K)} \]
\[ L := 6.096 \quad \text{Tube length (m)} \]

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi \cdot D_i \cdot L \quad A_o := \pi \cdot D_o \cdot L \quad A_m := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \]
\[ t_w := D_o - D_i \]
\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ U := \frac{1}{h_o + \frac{k \cdot A_m}{h_i \cdot A_i}} \]

\[ U = 1.891 \times 10^3 \quad \text{Heat Transfer Coefficient (W/m^2-K)} \]

\[ \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]
\[ R := 10.487 \quad S = 0.092 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.975 \quad \text{Ft Correction Factor for 2-4 Shell Pass Heat Exchanger} \]

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]
\[ A = 990.989 \quad \text{Heat Exchange Area Needed (m^2)} \]
\[ A_{ft} := A \cdot 10.76 = 1.066 \times 10^4 \quad \text{Heat Exchange Area Needed (ft^2)} \]
Heat Exchanger E-102T Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 8.506 \times 10^7 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

\[ D_i := 0.0135 \]  
Tube Inside Diameter (m)

\[ k := 0.03276 \]  
Thermal Conductivity of Tube Side Fluid (W/m-K)

\[ C_p := 1046.5 \]  
Mass Heat Capacity of Tube Side Fluid (J/kg-K)

\[ F := \frac{472.9}{1074} \]  
Mass Flow Rate of 1 Tube (kg/s)

\[ A := \pi \left( \frac{D_i}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ G := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 2.307 \times 10^{-5} \]  
Viscosity (Pa-s)

\[ n := 0.4 \]  
Constant: \( n = 0.4 \) if heating, 0.3 if cooling

\[ h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\[ h_i = 4.988 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

\[ D_o := 0.0191 \]  
Tube Outside Diameter (m)

\[ D_s := 0.9398 \]  
Shell Diameter (m)

\[ k := 0.6279 \]  
Thermal Conductivity of Shell Side Fluid (W/m-K)

\[ C_p := 4223 \]  
Mass Heat Capacity of Fluid (J/kg-K)

\[ F := 1342 \]  
Mass Flow Rate of Fluid (kg/s)

\[ A := \pi \left[ \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \right] \]  
Cross Sectional Area for Flow (m^2)

\[ G := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 6.956 \times 10^{-4} \]  
Viscosity (Pa-s)

\[ C := 0.36 \quad n := 0.55 \]  
Kern Correlation Constants
\[ h_0 := C \frac{k}{D_0} \left( \frac{D_0 G}{\mu} \right)^n \left( \frac{C_p h}{k} \right)^{0.333} \]

Kern Correlation (18.21)

\[ h_0 = 1.062 \times 10^4 \]

Convective Heat Transfer Coefficient (W/m^2-K)

**Inlet and Outlet Temperatures**

<table>
<thead>
<tr>
<th>Side</th>
<th>Temperature Unit</th>
<th>Temperature Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hot Side</td>
<td>°C</td>
<td>207.1</td>
</tr>
<tr>
<td>Cold Side</td>
<td>°C</td>
<td>30.0</td>
</tr>
</tbody>
</table>

\[ T_{in} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**

<table>
<thead>
<tr>
<th>Side</th>
<th>Material</th>
</tr>
</thead>
<tbody>
<tr>
<td>Shell Side</td>
<td>Carbon Steel</td>
</tr>
<tr>
<td>Tube Side</td>
<td>Carbon Steel</td>
</tr>
</tbody>
</table>

\[ k := 46.6 \]

Thermal Conductivity of Tubes (W/m-K)

\[ L := 6.096 \]

Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi D_i L \quad A_o := \pi D_o L \quad A_m := \frac{\pi L (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \]

\[ t_w := D_o - D_i \]

\[ U := \frac{1}{h_o + \frac{t_w \cdot A_o}{k \cdot A_m} + \frac{A_o}{h_i \cdot A_i}} \]

\[ U = 1.923 \times 10^3 \]

Heat Transfer Coefficient (W/m^2-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \]

\[ S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 11.473 \]

\[ S = 0.085 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.975 \]

Ft Correction Factor for 2-4 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{in}} \]

\[ A = 1.005 \times 10^3 \]

Heat Exchange Area Needed (m^2)

\[ A_{ft} := A \cdot 10.76 = 1.081 \times 10^4 \]

Heat Exchange Area Needed (ft^2)
Heat Exchanger E-103T Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 8.797 \times 10^7 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

\[ D_i := 0.0135 \]  
Tube Inside Diameter (m)

\[ k := 0.0340 \]  
Thermal Conductivity of Tube Side Fluid (W/m-K)

\[ C_p := 1084 \]  
Mass Heat Capacity of Tube Side Fluid (J/kg-K)

\[ F := \frac{472.9}{1074} \]  
Mass Flow Rate of 1 Tube (kg/s)

\[ A := \pi \left( \frac{D_i}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ G := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 2.351 \times 10^{-5} \]  
Viscosity (Pa-s)

\[ n := 0.4 \]  
Constant: n = 0.4 if heating, 0.3 if cooling

\[ h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\[ h_i := 5.134 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

\[ D_o := 0.0191 \]  
Tube Outside Diameter (m)

\[ D_s := 0.9398 \]  
Shell Diameter (m)

\[ k := 0.6279 \]  
Thermal Conductivity of Shell Side Fluid (W/m-K)

\[ C_p := 4223 \]  
Mass Heat Capacity of Fluid (J/kg-K)

\[ F := 1388 \]  
Mass Flow Rate of Fluid (kg/s)

\[ A := \pi \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \]  
Cross Sectional Area for Flow (m^2)

\[ G := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 6.956 \times 10^{-4} \]  
Viscosity (Pa-s)

\[ C := 0.36 \quad n := 0.55 \]  
Kern Correlation Constants
\[ h_o := C \cdot \frac{k}{D_o} \left( \frac{D_o G}{\mu} \right)^n \left( \frac{C_p \mu}{k} \right)^{0.333} \]

Kern Correlation (18.21)

\[ h_o = 1.082 \times 10^4 \]

Convective Heat Transfer Coefficient (W/m²-K)

**Inlet and Outlet Temperatures**

<table>
<thead>
<tr>
<th>Hot Side</th>
<th>Cold Side</th>
</tr>
</thead>
<tbody>
<tr>
<td>( T_{hi} := 207.5 )</td>
<td>( T_{ci} := 30.0 )</td>
</tr>
<tr>
<td>( T_{ho} := 35.0 )</td>
<td>( T_{co} := 45.0 )</td>
</tr>
</tbody>
</table>

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**

- Shell Side: Carbon Steel
- Tube Side: Carbon Steel

\( k := 46.6 \)

Thermal Conductivity of Tubes (W/m-K)

\( L := 6.096 \)

Tube length (m)

**Overall Heat Transfer Coefficient and \( F_t \) Correction Factor**

\[ A_i := \pi D_i L \quad A_o := \pi D_o L \quad A_m := \frac{\pi L (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \quad t_w := D_o - D_i \]

\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ U := \frac{1}{h_o \cdot k \cdot A_m + A_o \cdot A_i} \]

\[ U = 1.96 \times 10^3 \]

Heat Transfer Coefficient (W/m²-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 11.5 \quad S = 0.085 \]

From the \( R \) and \( S \) Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\( F_t := 0.975 \)

\( F_t \) Correction Factor for 2-4 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ \frac{Q}{U \cdot F_t \cdot T_{lm}} \]

\[ A = 1.018 \times 10^3 \]

Heat Exchange Area Needed (m²)

\[ A_{ft} := A \cdot 10.76 = 1.095 \times 10^4 \]

Heat Exchange Area Needed (ft²)
Heat Exchanger E-104T Design
Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 4.494 \times 10^8 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

- Tube Inside Diameter \( D_i := 0.0135 \) (m)
- Thermal Conductivity of Tube Side Fluid \( k := 0.3701 \) (W/m-K)
- Mass Heat Capacity of Tube Side Fluid \( C_p := 3397 \) (J/kg-K)
- Mass Flow Rate of 1 Tube \( F := \frac{137.3}{1074} \) (kg/s)

\[ \Delta := \pi \left( \frac{D_i}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ G := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 2.283 \times 10^{-4} \]  
Viscosity (Pa-s)

\[ n := 0.4 \]  
Constant: \( n = 0.4 \) if heating, 0.3 if cooling

\[ h_i := 0.023 \frac{k}{D_i} \left( \frac{D_i \cdot G}{\mu} \right)^{0.8} \left( \frac{C_p \cdot \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\[ h_i = 5.086 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

- Tube Outside Diameter \( D_o := 0.0191 \) (m)
- Shell Diameter \( D_s := 0.9398 \) (m)
- Thermal Conductivity of Shell Side Fluid \( k := 0.0499 \) (W/m-K)
- Mass Heat Capacity of Fluid \( C_p := 1291 \) (J/kg-K)
- Mass Flow Rate of Fluid \( F := 499.2 \) (kg/s)

\[ \Delta := \pi \left( \frac{D_s}{2} \right)^2 - (1022) \left[ \pi \left( \frac{D_o}{2} \right)^2 \right] \]  
Cross Sectional Area for Flow (m^2)

\[ G := \frac{F}{A} \]  
Fluid Mass Velocity (kg/m^2-s)

\[ \mu := 2.800 \times 10^{-5} \]  
Viscosity (Pa-s)

\[ C := 0.36 \quad n_c := 0.55 \]  
Kern Correlation Constants
\[ h_o := C \frac{k}{D_o} \left( \frac{D_o G}{\mu} \right)^{n} \left( \frac{C_p \mu}{k} \right)^{0.333} \]

Kern Correlation (18.21)

Convective Heat Transfer Coefficient (W/m²-K)

**Inlet and Outlet Temperatures**

**Hot Side**

\[ T_{hi} := 767.2 \]

\[ T_{ho} := 70 \]

**Cold Side**

\[ T_{ci} := 51.73 \]

\[ T_{co} := 600.0 \]

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln\left(\frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}}\right)} \]

**Materials of Construction**

Shell Side - Carbon Steel

Tube Side - Carbon Steel

\[ k := 16.3 \]

Thermal Conductivity of Tubes (W/m-K)

\[ L := 6.096 \]

Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi D_i L \]

\[ A_o := \pi D_o L \]

\[ A_m := \frac{\pi L (D_o - D_i)}{\ln\left(\frac{D_o}{D_i}\right)} \]

\[ t_w := D_o - D_i \]

\[ U := \frac{1}{h_o + \frac{t_w A_o}{k A_m} + \frac{A_o}{h_i A_i}} \]

\[ U = 749.783 \]

Heat Transfer Coefficient (W/m²-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \]

\[ S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 1.272 \]

\[ S = 0.766 \]

From the R and S Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\[ F_T := 0.65 \]

Ft Correction Factor for 4-8 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U F_T T_{lm}} \]

\[ A = 1.371 \times 10^4 \]

Heat Exchange Area Needed (m²)

\[ A_{ft} := A \cdot 10.76 = 1.475 \times 10^5 \]

Heat Exchange Area Needed (ft²)
Heat Exchanger E-105T Design

Heat Exchanger Specifications - Based on Table 18.6
One Pass - Triangular Pitch
Shell Diameter - 37 inches (0.9398 m)
Tube Diameter - 3/4 inches (0.019 m)
Triangle Pitch - 1 inch (0.0254 m)
Number of Tubes - 1074

\[ Q := 3.264 \times 10^8 \]  
Heat Exchanger Duty (W)

**Tube Side Properties**

\[ D_1 := 0.0135 \]  
Tube Inside Diameter (m)

\[ k := 0.3328 \]  
Thermal Conductivity of Tube Side Fluid (W/m-K)

\[ C_p := 3053 \]  
Mass Heat Capacity of Tube Side Fluid (J/kg-K)

\[ F := \frac{137.3}{1074} \]  
Mass Flow Rate of 1 Tube (kg/s)

\[ A := \pi \left( \frac{D_1}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ \mu := 2.776 \times 10^{-4} \]  
Viscosity (Pa-s)

\[ n := 0.3 \]  
Constant: n = 0.4 if heating, 0.3 if cooling

\[ h_i := 0.023 \frac{k}{D_1} \left( \frac{D_i G}{\mu} \right)^{0.8} \left( \frac{C_p \mu}{k} \right)^n \]  
Sieder-Tate Equation (18.13)

\[ h_i = 3.851 \times 10^3 \]  
Convective Heat Transfer Coefficient (W/m^2-K)

**Shell Side Properties**

\[ D_o := 0.0191 \]  
Tube Outside Diameter (m)

\[ D_s := 0.9398 \]  
Shell Diameter (m)

\[ k := 0.6279 \]  
Thermal Conductivity of Shell Side Fluid (W/m-K)

\[ C_p := 4225 \]  
Mass Heat Capacity of Fluid (J/kg-K)

\[ F := 5148 \]  
Mass Flow Rate of Fluid (kg/s)

\[ A := \pi \left( \frac{D_s}{2} \right)^2 - \left( \frac{D_o}{2} \right)^2 \]  
Cross Sectional Area for Flow (m^2)

\[ \mu := 6.959 \times 10^{-4} \]  
Viscosity (Pa-s)

\[ C := 0.36 \quad \mu_c := 0.55 \]  
Kern Correlation Constants
\[ h_o := C \cdot \frac{k}{D_0} \left( \frac{D_o G}{\mu} \right)^n \left( \frac{C_p \mu}{k} \right)^{0.333} \]

Kern Correlation (18.21)

\[ h_o = 2.225 \times 10^4 \]

Convective Heat Transfer Coefficient (W/m^2-K)

**Inlet and Outlet Temperatures**

**Hot Side**
- \( T_{hi} := 74.13 \)
- \( T_{ho} := 50.0 \)

**Cold Side**
- \( T_{ci} = 30.0 \)
- \( T_{co} := 45.0 \)

\[ T_{lm} := \frac{(T_{hi} - T_{co}) - (T_{ho} - T_{ci})}{\ln \left( \frac{T_{hi} - T_{co}}{T_{ho} - T_{ci}} \right)} \]

**Materials of Construction**

- **Shell Side** - Carbon Steel
- **Tube Side** - Carbon Steel

\( k := 16.3 \)

Thermal Conductivity of Tubes (W/m-K)

\( L := 6.096 \)

Tube length (m)

**Overall Heat Transfer Coefficient and Ft Correction Factor**

\[ A_i := \pi \cdot D_i \cdot L \quad A_o := \pi D_o \cdot L \]

\[ A_i = 0.259 \quad A_o = 0.366 \]

\[ A := \frac{\pi \cdot L \cdot (D_o - D_i)}{\ln \left( \frac{D_o}{D_i} \right)} \]

\[ t_w := D_o - D_i \]

\[ U := \frac{1}{h_o + k \cdot A_m + h_i \cdot A_i} \]

\[ U = 1.221 \times 10^3 \]

Heat Transfer Coefficient (W/m^2-K)

\[ R := \frac{T_{hi} - T_{ho}}{T_{co} - T_{ci}} \quad S := \frac{T_{co} - T_{ci}}{T_{hi} - T_{ci}} \]

\[ R = 1.609 \quad S = 0.34 \]

From the \( R \) and \( S \) Values, determine an appropriate shell pass configuration using graphs in Seider on page 485

\( F_T := 0.95 \)

Ft Correction Factor for 4-8 Shell Pass Heat Exchanger

**Heat Transfer Area Needed**

\[ A := \frac{Q}{U \cdot F_T \cdot T_{lm}} \]
Appendix I: Radiant Syngas Cooler Design Sheets
Radiant Syngas Cooler Design

Estimation for heat exchange area

\[ m_{\text{syngas}} := 1.42 \times 10^2 \]  
Mass flow of syngas through RSC \( \text{kg/s} \)

\[ C_{p\text{syngas}} := 1.803 \]  
Heat capacity of syngas (from UNISIM) \( \text{kJ/kg\cdotK} \)

\[ U := 0.142 \]  
Heat transfer Coefficient From Wen 1979 \( \text{W/m}^2\cdot\text{K} \)

\[ T_{\text{syngasin}} := 1673 \]  
Inlet and Outlet Temperatures in Kelvin

\[ T_{\text{syngasout}} := 1023 \]

\[ T_{\text{waterin}} := 303 \]

\[ T_{\text{waterout}} := 773 \]

\[ R_h := \frac{T_{\text{syngasin}} - T_{\text{syngasout}}}{T_{\text{waterout}} - T_{\text{waterin}}} \]

\[ S_h := \frac{T_{\text{waterout}} - T_{\text{waterin}}}{T_{\text{syngasin}} - T_{\text{waterin}}} \]

\[ F_T := \sqrt{R_h^2 + 1 - \ln \left( \frac{1 - S_h}{1 - R_h S_h} \right)} \]

\[ \Delta T_1 := T_{\text{syngasout}} - T_{\text{waterin}} \]

\[ \Delta T_2 := T_{\text{syngasin}} - T_{\text{waterout}} \]

\[ \Delta T_{LM} := \frac{(\Delta T_1 - \Delta T_2)}{\ln \left( \frac{\Delta T_1}{\Delta T_2} \right)} \]

\[ \text{Area} := \frac{m_{\text{syngas}} C_{p\text{syngas}} (T_{\text{syngasin}} - T_{\text{syngasout}})}{U F_T \Delta T_{LM}} \]

\[ \text{Area} = 1.586 \times 10^3 \text{ square meters} \]

\[ \text{Area}_{ft} := \text{Area} \times 10.7639104 = 1.707 \times 10^4 \text{ square feet} \]
Outer Shell design specifications from RSC Polk IGCC in Florida

\[
\begin{align*}
H_{250\text{MW}} & := 138.75 \\
D_{250\text{MW}} & := 25.92 \\
W_{250\text{MW}} & := 1052800
\end{align*}
\]

Core design specifications from RSC Polk IGCC in Florida

\[
\begin{align*}
H_{\text{core}} & := 120 \\
D_{\text{core}} & := 21.67 \\
W_{\text{core}} & := 575000
\end{align*}
\]

Weight := \( W_{250\text{MW}} + W_{\text{core}} \)

Design Specifications for Radiant Syngas Cooler

Height := 138.75 ft

Diameter := 25.92 ft

\[
\begin{align*}
\text{Weight} & = 1.628 \times 10^6 \text{ lb} \\
\text{Area}_{\text{ft}} & = 1.707 \times 10^4 \text{ ft}^2
\end{align*}
\]