Evaluation of a Process to Convert Biomass to Methanol Fuel
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Glossary

ASME American Society of Mechanical Engineers
atm atmospheres
BMS burner management system
FV Solenoid flow valve
HPR hydropyrolyzer
HX Heat exchanger
MSR methanol synthesis reactor
P&ID process and instrumentation diagram
PID proportional, integral and derivative
psia pounds per square inch, absolute
psig pounds per square inch, gauge
scf standard cubic feet
scfm standard cubic feet per minute
SPR steam pyrolysis reactor

Metric Conversions

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<th>English equivalent</th>
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<tr>
<td>degrees C</td>
<td>(degrees F x 9/5) + 32</td>
</tr>
<tr>
<td>kilogram (kg)</td>
<td>pound x 2.2</td>
</tr>
<tr>
<td>kilopascal (kPa)</td>
<td>psi x 20.88</td>
</tr>
<tr>
<td>liters per minute (L/min)</td>
<td>cubic feet per minute x 0.035</td>
</tr>
<tr>
<td>meter (m)</td>
<td>foot x 3.28</td>
</tr>
<tr>
<td>millimeter (mm)</td>
<td>inch x 0.039</td>
</tr>
<tr>
<td>megapascal (MPa)</td>
<td>psi x 20885</td>
</tr>
<tr>
<td>Normal cubic meters per hour (Nm³/hr)</td>
<td>scfm x 0.0283 x 60</td>
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Abstract

The U.S. Environmental Protection Agency and the University of California entered into Cooperative Agreement 824308-010 to develop and demonstrate the Hynol Process, a high-temperature, high-pressure method for converting biomass to methanol fuel. The period of performance was June 1995 to June 2000. At the bench scale, the Hynol Process has demonstrated about 75% carbon conversion efficiency with indications of low tar formation. A model developed during the bench scale testing predicts an increased carbon conversion efficiency from 75% to 88% for an increase in residence time from 1 to 7 hours. The Hynol reactor was designed to have a 7-hour residence time. The high efficiency and the potential for low tar formation hold promise for a cost effective technique for renewable fuel production.

The specific requirements of the Cooperative Agreement were for the UC Riverside College of Engineering-Center for Environmental Research and Technology (CE-CERT) to develop a pilot-scale (23 kg of feedstock/hr) Hynol facility and to operate it using woody biomass and natural gas as cofeedstocks. Cofunding was provided by the California Energy Commission, and the Riverside County Waste Resources Management District. CE-CERT contributed substantial additional funding.

The research focuses on producing methanol for use in a vehicle; however, the process can be modified to yield hydrogen, methane, or other fuels suitable for use in electricity generation. It also will contribute to environmental goals by reducing emissions of greenhouse gases, by providing a clean fuel, and by mitigating problems associated with disposal of carbonaceous waste. The key objective was to demonstrate the biomass gasification step of the Hynol Process and its reactions to produce a synthesis gas. The processes involved in converting this gas to fuel are demonstrated commercial technology, and those systems can be added later.

Design and construction, based on equipment specifications developed by Acurex under EPA Project No. 600/R-96-006 (Hynol Process Engineering: Process Configuration, Site Plan, and Equipment Design, February 1996), were originally expected to be completed by June 1998. Because of errors in the report and problems with the facility, the actual completion date has been pushed back. This report describes numerous design considerations that were reviewed; design modifications made; and preliminary results from operating the facility.
1. Introduction

1.1 Background and Theoretical Approach

Producing methanol from biomass offers significant environmental, energy, and economic advantages over other liquid fuel resources. Methanol is cleaner-burning than gasoline, so its widespread use can contribute to air quality improvements in urban areas. The fuel also can be produced from domestic, renewable resources, which brings advantages in emissions of greenhouse gases, energy security, and local jobs.

Process simulation studies indicate that the Hynol process should result in improved efficiencies in methanol production through increased yields over conventional processes. The advantages of the Hynol Process to the EPA are its potential to (1) produce liquid transportation fuel at a cost competitive with conventional fuels when used in fuel cell vehicles; (2) increase the quantity of biomass that could be produced as energy crops at a price acceptable for conversion to transportation fuel, thus increasing farm income; (3) displace more petroleum fuel than any other process based on biomass as a source of energy; and (4) achieve greater overall net reduction of greenhouse gas emissions from the U.S. vehicle fleet than any other biofuel option.

The Hynol Process originated at Brookhaven National Laboratory as a method for increasing the yield of fuel from conversion of biomass (Steinberg and Dong, 1994a). Originally conceived to operate with a coal feedstock, the process has been applied to co-processing biomass with fossil fuels, coal, oil, and gas at high temperature and high pressure. The process produces
methanol, a liquid fuel that can be used for transportation, industrial processes, electrical power generation, and military needs. Alternatively, the process can be modified to produce hydrogen or other chemicals for industrial uses.

The process involves three phases (Figure 1-1).

1. Reaction of biomass in a hydrogasifier, also referred to as a hydropyrolyzer (HPR).

2. Steam pyrolysis of the resulting gas, which produces a synthesis gas.

3. Methanol synthesis, which leaves a recycle gas that can be returned to the HPR and waste heat that can be returned to the steam pyrolyzer (SPR).

The basic Hynol Process consists of two reactions:

1. Hydrogenation (or hydropyrolysis) of the carbonaceous feedstock to produce methane.

2. Endothermic reaction of methane with steam to produce hydrogen and carbon monoxide (steam pyrolysis).

For methanol production, the carbon monoxide formed in the steam pyrolysis step is catalytically combined with the hydrogen in a third phase to produce methanol. Excess hydrogen is recycled as a feed gas for hydropyrolysis. Biomass is fed into a fluidized-bed HPR and reacted with recycled H₂-rich process gas at 30 atm pressure and 800 °C (Steinberg and Dong, 1994b). Steam at a rate of 0.2 kg per kg of biomass is simultaneously fed into the HPR. The independent reactions taking place in the HPR can be expressed as:

\[ C + 2H_2 \rightarrow CH_4 \]  
\[ C + H_2O \rightarrow CO + H_2 \]  
\[ CO_2 + H_2 \rightarrow CO + H_2O \]

(1)  
(2)  
(3)

The process gas produced in the HPR contains 13 mole % CO, 38 mole % H₂, and 20 mole % CH₄. Nitrogen that comes from the feedstock forms inert N₂ in the process gas and is taken into account in the calculation of equilibrium gas composition. The objective is to demonstrate conversion of the carbon in biomass feedstock in the HPR to be over 87%. The unconverted carbon is withdrawn from the reactor with ash in the form of char. The char either can be used as fuel for the SPR (if separated from the sand, limestone, and/or kaolinite) or sequestered.

Reactions (2) and (3) are endothermic and require additional energy input to the gasifier. This is why conventional gasification processes need oxygen or air to supply combustion heat by burning some carbon in the feedstock within the gasifier. In the Hynol Process, however, thermal energy from recycled gas combined with reactions in the HPR allows for an energy-neutral gasifier without the need for an internal or external heat supply. The hydrogasification reaction (1) between the carbon in feedstocks and the hydrogen in the recycled process gas is exothermic and in theory provides sufficient heat for reactions (2) and (3) when preheated by heat exchange with the SPR effluent stream.

Before entering the SPR, process gas from the HPR of the Hynol Process usually needs to be cleaned up to remove particulate matter and impurities that may contaminate catalysts in the subsequent reaction steps. Conventional hot gas cleanup methods can be used for this purpose. Natural gas feed can be added prior to the HPR filter to cool the gas stream and maintain a more filter-friendly operating environment.

The process gas is then introduced to the steam reformer (alternatively called the SPR) where HPR outlet gas and natural gas co-feedstock react with steam to form CO and H₂. The steam reforming can be described by two independent reactions:

\[ CH_4 + H_2O \rightarrow CO + 3H_2 \]  
\[ CO_2 + H_2 \rightarrow CO + H_2O \]

(4)  
(5)

The SPR is a steam reformer using a conventional nickel catalyst but operating at higher temperature (900-950 °C and higher pressure (30 atm). The mol ratio of steam to carbon entering the SPR is 2.5. A catalyst-packed tubular externally-fired furnace reactor similar to a conventional natural gas reformer furnace reactor is used for the SPR. Steam feed ratio is 1.2 kg per kg of biomass. Methane feed into the SPR is at a rate of 0.5 kg per kg of biomass. The H₂ and CO concentrations in the exit gas of the SPR are increased to 60% and 21%, respectively. The process gas is then passed through a gas heat exchanger, where it is cooled. The recovered heat is used to heat the recycled gas. The process gas is cooled for the methanol synthesis reactor (MSR) feed. The steam produced in this way is about 1.52 times biomass feed rate in weight, which makes steam self-sufficient within the system.
The cooled process gas then enters the MSR to produce methanol. The reactions taking place in the MSR are:

\[ CO + 2H_2 \rightarrow CH_3OH \]  
\[ CO_2 + 3H_2 \rightarrow CH_3OH + H_2O \]

The methanol synthesis is performed at 30 atm and 260 °C. The MSR reactions are highly exothermic, so the released process heat can be extracted from the MSR and used to dry the biomass feedstock. Methanol and water are separated from the MSR effluent gases by a condenser and fractionated by distillation to obtain a pure methanol product. To increase the conversion of CO in the MSR, the uncondensed gas from the condenser is partially returned to the MSR. Using this approach, the recycle ratio of the internal loop is 4 moles per 1 mole of input process gas from the SPR. The net result is a 90% conversion of CO to methanol in the MSR. Unlike conventional processes where CO conversion in the MSR is a most critical parameter affecting the efficiency losses of the process, the Hynol Process reprocesses the unconverted material by recyling the gas to the HPR and thus prevents losses of process gas constituents. For this reason, the Hynol Process obtains a high thermal efficiency, even though the CO conversion through the MSR may be lower than that of conventional processes.

The condenser operates at 50 °C. The gas exiting the MSR system is introduced to the gas heat exchanger, after a small amount of gas (3.7% of the recycled gas) is purged, eliminating the accumulation of inert nitrogen in the system and keeping the nitrogen concentration in the system below 2.5 mole %. We are designing the system to accommodate a range of steam and natural gas feeds. The entry points of the steam and natural gas prior to the HPR or SPR can also be adjusted as indicated by revised process modeling assessments.

### 1.2 Design and Performance Issues

The technical challenges of the Hynol gasifier are to optimize carbon conversion, minimize tar formation, control alkali agglomeration, maintain gasification bed temperatures, achieve steady-state operation, and demonstrate particulate control. In addition, reliable operation of the bed height estimation, biomass feed system, ash removal process, cyclone efficiency, alkali sampling system, and tar sampling system must be demonstrated.

The Hynol facility at Riverside was designed around optimizing carbon conversion of biomass in the gasifier. Carbon conversion is strongly dependent on the residence time inside the reactor (Dong and Cole., 1996; Dong, 1998). There is a tradeoff with too long a residence time because higher residence times reduce biomass throughput. A three-parameter kinetic model was developed and used for quantitatively investigating biomass conversion and reaction rate phenomena (Dong et al., 1996, 1998) as part of a project sponsored by the EPA. The effects of particle size, gas velocity, system pressure, reaction temperature, and gas composition on biomass hydrogasification behavior were investigated. The conclusions from this study were the basis for the operation of the Hynol reactor located at UC Riverside.

Below, the conclusions are summarized, and their significance in the design of the reactor is discussed.

1. The carbon conversion takes place in two stages. There is first a period of rapid reaction of biomass thermal decomposition (seconds), followed by a slow reaction of the residual char (hours). The twofold residence time was used to design the reactor bed height, the expanded zone, the cyclone, the fluidization velocity, and the bed media particle size. The bulk of the biomass is converted in first few seconds while the residual char is left in the fluidized bed until the particle size is small enough to pass through the cyclone. The cyclone was designed to return 95% of the particle fines to the reactor. The reactor zone (3 m height, 150 mm diam) was designed for the rapid reaction of biomass. The expanded zone (1.5 m height, 300 mm diam) was designed for the slow residual char conversion. The expected velocities in both sections are 0.3 m/s and 0.08 m/s, respectively.

2. The developed model can be used to predict biomass conversion as a function of reaction time assuming similar conditions for the tests. The model is used to estimate expected carbon conversion efficiencies for each test based on expected residence time from the operating conditions.

3. The gas film mass transfer is negligible at gas flow rates greater than 0.1 m/s. This information supports the decision to operate the fluidized bed at a velocity of 0.3 m/s.

4. Biomass particle sizes less than 3.2 mm do not have a strong impact on the rapid reaction rate. Although particle sizes below 3.2 mm are recommended, CE-CERT is using white oak from the waste stream of a hardwood door manufacturer. The particle size is
distributed with particles ranging from 6.4 mm to less than 3.2 mm. The larger 6.4 mm particles are expected to have a small effect on conversion efficiency because the designed residence time is greater for the pilot scale tests than the bench scale tests.

5. Nearly all the hydrogen and oxygen can be converted into gas products in 20 minutes using 3.2 mm wood particles. The Hynol reactor was designed to have a residence time of approximately 7 hours assuming the internal cyclone is 95% efficient.

6. At 30 atm and 800 °C, about 87% of poplar wood or 75% of its carbon content can be converted in 60 minutes. Extending the reaction time and increasing particle attrition in the reaction zone can achieve higher conversions. The model predicts that an increase in carbon conversion efficiency from 75% to 88% can be achieved with a residence time of approximately 7 hours. CE-CERT expects an 85% carbon conversion efficiency for the tests performed under this testing program.

7. Increases in reactor pressure from 10 atm to 60 atm only slightly increase biomass conversion. The reactor was designed for operation at 30 atm to reduce capital costs. Future designs should be based on desired throughput and reactor cost for optimized pressure rating. The 30 atm design base will be a good starting point for evaluating the process. Once the process is confirmed, the economics can be investigated.

8. Biomass conversion is greatly increased when reaction temperatures are raised from 800 °C to 950 °C. Although higher temperatures are desirable, alkali formation also increases with higher temperature. Therefore, it is desirable to operate at 800 °C ± 50 °C. Alkali formation was not investigated in the reaction rate experiments by Dong and Cole (1996).

9. The biomass conversion is proportional to the hydrogen partial pressure in the recycle feed gas. During operation one way to improve biomass conversion is to increase the hydrogen partial pressure. This will be useful when integrating phases 2 and 3.

10. The biomass conversion is proportional to steam partial pressure in the recycle feed gas. Increasing the steam partial pressure may increase the biomass conversion, but it reduces reactor temperatures. Again, this is a tradeoff between high conversion and maintaining bed temperatures.

11. When the methane co-feeding in the recycle is less than 15%, it effects are negligible to the biomass conversion. No methane co-feeding will be performed during phase 1 of the Hynol project, but it will be investigated for phases 2 and 3.

12. Carbon monoxide and carbon dioxide concentrations have no significant effect on the gasification behavior. This information was used to eliminate those variables that could affect the conversion efficiencies during operation.

13. The bench scale hydrogasification experiments with poplar have shown the potential for low tar formation. The Hynol reactor sample system was designed to evaluate tar formation during gasification tests.

Optimum gasification may not be ideal for steady-state operation. Alkali formation was not studied in the thermo balance reactor, and alkali agglomeration is very common in coal and biomass gasifiers (Miles et al., 1998; Unnasch, 1996). One solution to the problem of alkali formation is to use an adsorbent or chemical alkali getter. Kaolinite was found to be the most efficient in alkali control, but this is based on equilibrium models and has not been tested in gasification (Unnasch, 1996). Once steady-state hydrogasification is achieved, the alkali problem will be investigated.

Another problem with optimum carbon conversion is that the high bed temperatures increase the alkali formation concentration, and thus require the use of more alkali getter. Unnasch (1996) found that alkali formation starts at temperatures greater than 750 °C and peaks at 850 to 900 °C. Therefore, gasification temperatures need to be optimized for carbon conversion, but not at a high cost to alkali formation. There is a balance where too much getter is needed to offset the gains in carbon conversion. This operating point will be a function of the biomass and alkali getter used.

The design of the internal cyclone needs to be evaluated for its effectiveness and performance. This can be accomplished by operating at steady state and sampling the ash removed at the bottom of the filter compared
with the ash removed from the bottom of the gasifier. The cyclone was designed to remove 90-95% of the large particles and the filter is designed to remove 99.99% of the fines including particles as small as 0.5 µm. CE-CERT has installed a high-pressure sample system to evaluate the filter and cyclone performance. The sampling system also was designed to sample for tars and alkalis. However, due to design limitations, the sampling of tars and alkalis needs improvement to accurately follow ASTM standards.
2. Methodology

The basis for the approach to this project was provided by Unnasch (1996). The HPR system will initially be operated independently, decoupled from the SPR and MSR. For this reason, two elements of the Hynol system will not be available for decoupled HPR operation and will need to be simulated (i.e., recycle gas from the MSR and its heat source, the exit gas from the SPR). The recycle gas will be simulated by mixing gases from tube trailers with natural gas, steam, and vaporized liquid CO₂. Since the SPR will not be operating, the inter-reactor heat exchanger (preheating the recycle stream) will not operate at a high enough temperature, and approach temperatures will be too low to provide for the required HPR inlet temperature. An electric heater will provide the additional heat energy to the recycle gas that normally would be recovered from the SPR effluent stream. A flowsheet for the decoupled HPR system needs to consider the source and temperature of the simulated recycle gas since these gases will not be produced from system recycle but rather from bottled gases.

Unreacted solids and ash are removed from the reactor in two ways. The solids are removed directly from the bottom of the reactor using a lockhopper system. Lighter ash is removed from the top of the bed from an overflow passage, on the side of the vessel, which empties into a lockhopper system.

An internally mounted cyclone separates the majority of particles from the exiting gas. The outlet gas passes through a filter which is pulse-cleaned with nitrogen. The hot outlet gas is heat-exchanged with the cold inlet gas.

Some elements of the integrated Hynol system were incorporated into the design of the HPR system. The SPR uses an air compressor and natural gas compressor, and it should use a steam jacket. All of these systems are common with the HPR system and were incorporated into the HPR system design. The demand for methane and air vary with the different Hynol cases, so the feed requirements were incorporated into the HPR system.

The HPR system process flow diagram includes the following features that are of interest or differ between the theoretical integrated system and the actual decoupled system:

1. H₂, CO, CO₂, and N₂ are added from bottled gases, heated with a heat exchanger, and then heated further with an electric heater. These gases simulate some of the recycled HPR feed.

2. Steam from a heat exchanger is added upstream of the electric heater. This flow simulates both water vapor that is in the recycle stream and steam that is added to the HPR system. The electric heater raises the temperature to 1000 °C. Higher temperatures are difficult to achieve with an electric heater.

3. Provisions are also made to add natural gas downstream of the HPR. This stream represents the

The hydrogen, carbon monoxide, and nitrogen are fed from truck tube trailers and mixed to simulate the recycle gas in the fully integrated system. The inlet gases at ambient temperature are heated in the heat exchanger by the HPR outlet gas. An exhaust gas heat exchanger converts water to steam, which is injected after the heat exchanger. The mixture passes through an electrical heater before entering the burner where methane is injected.

The hydrogasifier is fed with a mixture of solids, primarily chipped wood (white oak, ~6.4 mm chips) and an alkali absorbing (gettering) agent. The green waste and getter are mixed together and fed into a day bin and lockhopper; the sand and getter can be fed separately directly into the lockhopper or mixed with the biomass feed. A screw-feeder meters the solids into the reactor vessel where they are fluidized.
methane feed to the SPR in the full (complete) system. About 10 percent of this stream is split off and used to purge the cyclone in the HPR. The balance of the natural gas is added after the HPR. For decoupled HPR operation, most of the methane need not be added to the system.

4. Methanol is present in small percentages in the recycle gas. However, the methanol would dissociate in a heat exchanger with an 888 °C outlet temperature. Therefore, for the decoupled HPR system, methanol should be added in the form of its constituent CO and H₂. The mass-flow (associated with methanol vapor) entering the HPR is held constant between the decoupled and integrated systems.

5. Some lockhopper pressurization gas carries over into the HPR, since the biomass voidage volume in the lockhopper is pressurized with nitrogen. Consequently, nitrogen gas enters the HPR. The mass and enthalpy of the nitrogen should be considered in the energy balance for the process. They are included on the flow sheet for the decoupled HPR.

6. Air and natural gas are combusted to warm up the HPR before startup. The corresponding flow rates are shown in the process flow diagram. Nitrogen that is heated with the electric heater will also be used during start-up operations. Nitrogen can flow through the electric heater, which will prevent the heater wires from overheating before simulated recycle gas is added to the system. Air may also need to be added upstream of the electric heater to allow for periodic oxidation of the heater wires.
3. Facility Construction

Construction began in February 1996 and was completed by March 1998. Modifications were made from June 1997 through December 1998. The first gasification test was completed in December 1998. Since December 1998, CE-CERT has completed three more gasification tests. Each test has shown some successes but revealed other design/operation problems. CE-CERT continues to perform gasification tests and process design shake-down evaluation.

There were a variety of delays with the construction, installation, and process evaluation that resulted in the schedule shown in Figure 3-1. The delivery delays were due to construction of a forged burner tee, repair to the refractory, and fabrication of a high-temperature inconel distributor and cyclone. The design modification delays were due to curing the refractory on site, repairing a large crack on the main burner spool piece, and installing the high-pressure feed system. The final process delays were due to process modifications that included repairing the feed system valves, biomass bridging, heater elements, ash removal valves, process controls, flow meter calibrations, and the burner management system.

CE-CERT has successfully demonstrated the automated high-pressure biomass feed system, maintained bed temperatures, achieved reliable burner operation, achieved reliable electric heater operation, and performed consistent ash removal. The main design challenges remaining are to overcome an agglomerating problem in the Hynol reactor and to perform steady state gasification.

Arcadis designed the vessels, refractory, and process flow, and CE-CERT designed the layout of the Hynol gasification test facility. CE-CERT also is responsible for operating and evaluating the Hynol process. The Arcadis design details can be found in Hynol Process Engineering: Process Configuration, Site Plan, and Equipment Design (EPA-600/R-96-006, Office of Research and Development, Washington, DC) (Unnasch, 1996). The facility layout houses a 59.4 Nm³/hr air compressor, a 24.8 MPa, 6.8 Nm³/hr natural gas compressor with 84,950 liter storage, a 1.7 Nm³/hr CO₂ booster pump, a 300kV electrical distribution panel, a biomass bulk storage area, a tube trailer bulk storage area, and a control room (Figure 3-2). As part of the project, the facility layout was designed to accommodate future phases 2 and 3 of the Hynol Process development.

The site development started in February 1996 and was finished by May 1996. Part of the site development was

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Figure 3-1. Hynol facility schedule.
a geological study and an environmental impact report; see Appendices I and II for copies of these studies. Figure 3-3a,b,c show the progress of work from earth moving to the completed foundation.

The vessel construction was contracted through Bay City Boilers, and the refractory was contracted through Dee Engineering. The vessels were constructed out of schedule 80 pipe except for the large burner spool pieces. ASME codes required the tee section to be forged. The forging delayed the delivery date and increased the vessel cost. The vessels were finished and delivered by March 1997 (Figure 3-4).

The steel structure, foundation and lighting were contracted through Martec International. Figure 3-5 shows a model layout of the structure and vessels. Once the vessels arrived at CE-CERT, the structure was erected in about two weeks (Figure 3-6). CE-CERT found a structure problem that could have been serious if it had gone
Figure 3-4. Vessel fabrication at Bay City Boilers. Main reactor piece 24 inch schedule 80 pipe with class 181.4 kg flanges (a); forged burner tee (b).

Figure 3-5. Structure design concept drawing with all the vessels.
unnoticed: One of the main four I-beam supports was twisted due to eccentric loading legs (Figure 3-7). The problem was repaired during vessel installation.

HP Construction was responsible for installing the vessels and completing the pressure leak test (Figure 3-8). The vessel installation was delayed due to problems with the inconnel pieces, refractory curing, installing heat exchanger insulation, and repairing the burner refractory. The vessel installation was completed March 1998 by successfully maintaining 5.2 mPa for 24 hours with a loss of no more than 10%.

Before the reactor could be operated, the control, instrumentation, and process equipment had to be installed. The extension of the schedule from June 1998 to December 1998 accounts for the time necessary to modify the process before operation of the facility could begin.
Figure 3-8. Vessel installation at CE-CERT’s Hynol test facility.
4. Design Modifications

The vessels, refractory, process flow, and other mechanical systems were originally designed under a separate EPA project. The design details can be found in Unnasch (1996). This section describes the modifications to the original design. There were no as-built drawings provided by Arcadis or CE-CERT; the following subsections describe the as-built system. The design modifications are in chronological order.

4.1 HPR Reactor

Refractory

The refractory arrived damaged and brittle. In some sections the refractory would flake off when touched. CE-CERT hired a local refractory installer to repair the damaged areas using VSL 50 to provide a 51mm-thick hot face to the lightweight GreenCast 19L. The VSL 50 has a maximum temperature rating of 1371 °C, while GreenCast 19L has a maximum temperature rating of 1038 °C. The new material serves two purposes. First, it allows the hard surface for the high gas flows and ash to pass without damage to the refractory surface. Second, it protects the under layer refractory from overheating. Figure 4-1 shows some of the damage to the refractory and its repair.

Curing Process

There were two options for curing the refractory: One was to have it done at Dee Engineering when the refractory was poured, and the other was to have the refractory cured at CE-CERT using the natural gas burner. When the vessels arrived at CE-CERT they were not cured, thus CE-CERT was given the task to use the burner system to cure the refractory. There were many problems with trying to cure the refractory that delayed progress by three to four weeks. The burner was unreliable, and the heat rate was too quick in the burner section and slow at the far end near the filter. Wet refractory should be heated no faster than 10 °C/hr according to the refractory manufacturer. Ideally, the process should have taken only four to five days, not three to four weeks. Also, the curing process was a 24-hour operation, thus making small problems larger due to time required to reheat cooled sections.

Burner Management System

While trying to cure the refractory, CE-CERT had to first modify the burner design to heat the vessels. The burner was designed to preheat the reactor, filter, and heat exchanger sections on start-up. The burner operation was controlled by a Honeywell Burner Management system (BMS). The purpose of the BMS was to control the flame based on a signal from a flame sensor. If the flame signal was below 1 volt, the gas valves were turned off. The BMS works as follows: The BMS starts the pilot and the igniter (similar to a spark plug on a vehicle). The BMS senses whether a flame is present by using a flame indicating rod. If no flame is sensed, the BMS turns off the pilot valves and stops the igniter after 30 seconds and waits to be restarted. If a flame signal is sensed, the BMS opens the main gas valves, closes the pilot valves and stops the igniter. The BMS continues to check the flame signal. If the signal gets weak (below 1 volt) the BMS turns off the main gas valves and waits to be restarted. The BMS function is common to a typical modern home furnace and gas laundry dryer; they operate as long as a good flame signal is present. The main difference was that our system operates in an enclosed environment at 689 kPa over ambient pressure. A high-pressure burner pilot/main system can be purchased, but they are typically custom-designed and expensive.

To start the burner, CE-CERT’s responsibility was to set the pilot and main mixture valves for a lean burn. This was done using flow meters on the pilot and main gas valves. The air-to-fuel ratio was 9.5 to 1 (by volume), which was equivalent to a burner mixture of lambda = 1.1 (lean of stoichiometric). Once the gases were set at a proper mixture, CE-CERT had to choose an appropri-
ate flow velocity that would not blow out the flame. CE-CERT found it hard to prevent the pilot and main flames from self-extinguishing at the desired flow velocities. The high velocities were needed to preheat all the vessels. CE-CERT finally found a good pilot and main flow velocities that gave reliable flames. A good flame was confirmed using two thermocouples installed at the pilot and at the main flame areas.

Once a good pilot and main flame were achieved, the BMS was ready to be turned on. The BMS failed to work for two reasons:

1. The igniter was arcing on the flame rod, and the flame rod was not always in the flame. The high-voltage arcing between the flame rod and the igniter would send a high-voltage spike to the BMS, which would turn off the BMS instantly. To start the pilot, the BMS turns on the igniter and waits for the flame signal, but the igniter was arcing when the flame sensor sent an incorrect signal to the BMS. The flame rod could not be moved away from the igniter because it needed to be in the flame near the pilot and the pilot needed to be next to the igniter. CE-CERT then looked at home furnace pilot designs to understand the problem. Home furnace designs have a metal plate separating the igniter from the flame rod, while directing the flame to the flame rod.

2. The second problem was related to the function of the BMS when the main flame turns on. The problem is in the flame rod location. The sensing tip of the rod was in the pilot flame, not in the main flame. When the pilot flame is off, the flame rod needs to be in the main flame as well. The flame rod was too close to the base of the main flame, thus generating a low flame signal. This caused the BMS to turned off the main gas valves seconds after they are turned on. Moving the flame rod to a spot in the main flame would solve the problem when the main was on, but would not allow the pilot to sense a flame signal. CE-CERT chose to bypass the BMS system and to control the flame manually using flame temperature as the control signal.

Secondary Air

To reach the high volume (59.4 Nm³/hr air and 3.4 Nm³/hr natural gas) of heat needed to preheat all the vessels, the system had to be pressurized to 241 kPa, and secondary air was required to reduce flame temperatures and to move the heat throughout the vessels. The secondary air was designed to be at a 90-degree angle to the mid-point of the main flame. Secondary air at this location would carve a hole in the refractory wall, because the Green Cast 19L was too light to handle the high velocities. Replacement to the refractory in the burner section was not possible unless it was completely replaced. Instead, CE-CERT tried putting the secondary...
air at the base of the flame, where the high-density refractory could handle the high velocities. At this location, the secondary air blew out the flame. The secondary air was then moved to the bottom of R101 (where the Mogas valves mount). This location delivered the heat to all the vessels as required. To keep the secondary air on the bottom of R101, CE-CERT will modify the start-up program to allow the Mogas valves to be open during the preheating process. Future designs should allow for secondary air to be added in the direction of the flame near the middle or end. A higher-density burner refractory is also recommended.

**Low-Pressure Igniter**

After operating the burner it was discovered that the flame and igniter rods were leaking at the electrical connections. CE-CERT found that the flame and igniter rods were low-pressure units and needed to be replaced with sealed high-pressure units. Because the BMS is not going to be used, CE-CERT has replaced the igniter unit with a high-pressure equivalent. Figure 4-2 is a photo of the old flame sparker and the new high-pressure unit fabricated by CE-CERT. During gasification test 1a, the modified spark system turned out to be unreliable, so a third design was necessary. This is described in Section 5 of this report.

**Refractory Damage (#2)**

Burner temperatures exceeded 1093 °C in order to heat up the entire reactor as necessary according to the cure SOP and to preheat the reactor for gasification tests. The refractory in the burner section was only rated for 1037 °C and failed during operation at temperatures above 1037 °C.

When the refractory was repaired the first time, a section in the burner was not repairable because of its location. Unfortunately, this section failed while the vessels were being cured. CE-CERT hired the same engineering firm that repaired the other refractory surfaces to repair the burner casting. The repair required removing the vessel from the structure, shipping it to B&B Engineering, and removing the old casting (Figure 4-3). The new refractory used GreenKleen 60 for the hot face and GreenCast 19L for the secondary lining. The hot face was 76 mm and the secondary lining was 89 mm. Before delivery to CE-CERT, the refractory was cured in an oven.

The new burner casting was rated to 1649 °C and has the rigidity of concrete. The burner skin temperature was expected to be no higher than 165 °C with a gas temperature of 1204 °C. The new casting should be able to withstand high gas velocities, which means the secondary air can be located back to the designed location near the burner flame. During gasification tests this refractory also failed during operation, as discussed in Section 5.

**Burner Vessel Crack On Lower 18-inch Flange**

CE-CERT found a surface crack on the lower 18-inch flange weld (Figure 4-4). Secondary inspection was required to determine the depth and details of the crack. Bay City Boilers hired an ultrasonic ASME inspector to X-ray the crack. The inspector found the crack penetrated a depth of 13 mm and almost 270 degrees around the flange. Because of the depth and length of the crack, the

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*Figure 4-2. Igniter systems: (a) the old spark system; (b) modified spark system using a high-pressure electrical pass-through; and (c) pressure vessel for location of ground pilot system.*
flange was cut off and replaced. The vessel was removed and sent to Bay City Boilers (which warranted the repair for CE-CERT). After the crack was repaired and recertified, CE-CERT sent the vessel to a refractory shop to replace the refractory as mentioned earlier.

After finding one crack, CE-CERT hired an ASME inspector to test another 10% of the welds. If any serious cracks were found, CE-CERT would then test another 10% of the welds. No serious cracks were found, but two minor cracks on the heat exchanger vessel were identified (Figure 4-5). Both cracks were weld-
ed over using a certified welder, and the inspection grindings were brought up to full penetration. During the repair a pin hole in the connecting tubing was found and repaired.

4.2 Biomass Feed System

The feed system was designed to automatically feed biomass into the high-pressure reactor from a feed storage hopper. Interference problems and design operational problems had to be solved before the system was operational. In addition, the electrical work and control logic were not included with the design. CE-CERT took the responsibility to modify the mechanical interferences and to program and wire the system to the point of automation.

Feed System Overflow Chutes

The feed system overflow chute alignment was not designed properly and interfered with the feed valves and bucket elevator. The overflow chute was cut, rotated and re-welded to allow for proper feed system operation (Figure 4-6). The figures show the type of modifications and interferences that prevented the feed system from being installed. To make the feed overflow chute fit from the top of the valves to the storage bin, the angle needed to be decreased from 55° to 45°. The change in angle caused a bridging problem, which is discussed below.
Feed System Storage

The biomass storage bin required extra mounts not covered in the preliminary site development. CE-CERT designed and installed new supports for the storage bin (Figure 4-7). A second problem was getting biomass into the storage bin. A truckload of white oak biomass was delivered to CE-CERT and conveyed into the storage bin using a rotary blower. It takes about one day to load enough biomass for a two-day gasification test. In the future, CE-CERT expects to rent a skip loader, which should do the same work in less than three hours.

The biomass is stored in the hopper and on the ground under nylon tarpis. Moisture analyses were conducted throughout the year. The moisture content averages about 13% on the ground pile and 6% in the storage bin pile ±2%. The average moisture value on the ground piles at a depth of 1.5 m was 6%, and there was no moisture difference for the piles in the storage bin.

Electrical Controls

The final assembly of the feed system required installing the vibrators, electrical controls, motor variable frequency drives, and bucket elevator. Additionally, the entire feed system needed to be programmed by CE-CERT before operation could be achieved. The vibrators were mounted on the grating and connected to variable resistors to set the vibration control point (Figure 4-8). Each vibrator, one for sand and one for aluminum oxide, was calibrated for a specific mass flow and programmed into the control software.

Appendix VII shows the locations of key components and sensors. The electrical controls for the hopper screw (SC-801), meter screw (SF-805), and feed screw (SF-806) were installed on the back side of the feed valves for easy access and close proximity to the drive motors (Figure 4-9a,b). Ladder logic drawings that show the wiring detail were drawn by CE-CERT and are included in Appendix III.

The level sensor, installed in the meter bin, is a capacitive type sensor (Figure 4-9c). When biomass fills the meter bin the capacitance increases with height. During preliminary runs the level signal was too weak to measure. To fix the problem, the surface area of the capacitive sensor was increased. Also, when installing the sensor it was discovered the bend angle was incorrect. CE-CERT modified the bend angle to get it to fit into the meter bin. The capacitive level sensor has been a reliable measure of biomass level.

The bucket elevator shown in Figure 4-9d is used to convey biomass from the ground floor to the top floor, a distance of 7.3 m. Once the precharge hopper is full, biomass overflows back into the hopper bin. The bucket elevator was assembled at CE-CERT and required minimal troubleshooting. The bucket elevator is a reliable method for conveying biomass from the ground floor to the top floor.

CE-CERT tried feeding biomass into the reactor once the feed system was mechanically and electrically functional. Feeding biomass into the reactor was possible because the burner spool piece (B-037) was removed for repair. Biomass could exit the reactor through the 152 mm opening between the burner and the reactor. The plan was to operate the feed system at different speeds and calibrate the mass flow. Unfortunately, the calibration was delayed because bridging problems were discovered in eight locations (Figure 4-10). The problems were located at:

1. Hopper bin.
2. Bucket elevator neck.
3. Return overflow chute connection.

Figure 4-7. Support legs for large storage bin (a). Biomass delivery system (b) and biomass transport into the hopper (c) also are shown.
4. Precharge hopper.
5. Feed valve.
7. Feed valve.
8. Meter bin.

Each problem required some type of mechanical or electrical design change to get continuous flow through the feed system.

Location 1 was a bridging problem in the hopper storage bin (Figure 4-11). The biomass easily supported an edge and prevented flow into the screw below. If the problem went unnoticed for about one hour, the hopper screw would not have any biomass and the system would run dry. This problem was solved by manually leveling the pile hourly. For commercial autonomous-level operation, large eccentric rotating vibrators would be recommended.

Location 2 was a bridging problem caused by operating the hopper screw at too high a speed. Too much mass was flowing, and it would jam in the small neck at the connection from the bucket elevator and the top of the precharge hopper. To fix the problem the hopper screw rate was reduced as needed.

Location 3 was a bridging problem in the overflow neck on top of the precharge hopper. This problem was fixed by a cleanout air pulse five times at the beginning of a feed cycle. For commercial applications, this modification may not be necessary because of the larger size of the opening.

Location 4 was a bridging problem in the overflow chute connecting the precharge hopper to the hopper bin. The solution was to provide a steady air stream down the chute controlled with the operation of the bucket elevator (Figure 4-12). The air jet was on whenever the bucket elevator was operating. For commercial installations, the bin could be located such that the angle
Figure 4-10. Bridging problem locations.

Figure 4-11. Bridging problems in the (a) hopper screw; (b) bridging with feed screw exposed; (c) bridging with feed screw empty thus not feeding biomass.
from the overfill point to the bin is >55 degrees. A steep angle tends to prevent bridging problems.

Locations 5 and 7 were similar bridging problems that occurred in feed valves FV-839 and FV-840. The large Everlasting valves require a purge gas during closing to prevent any materials from getting caught behind the valve as it swings closed (Figure 4-13). The solution was to incorporate a high-pressure pulse prior to closing both the top and lower feed valves. Location 6 was a problem with a mechanical liner designed to help biomass flow through the valves. The relief angle on the liner made the final hole too small for biomass to pass through. CE-CERT replaced the liner with a straight section, which solved the problem. For commercial applications, problems 5 and 7 will still exist, but problem 6 most likely will not.

The last bridging problem was in the meter bin, Location 8. Location 8 was the hardest to solve because
the problem was not consistent. The problem occurred when the level in the meter bin got too high. The biomass would compact and bridge over the meter screw. The solution was to limit the biomass level to a safe height. In commercial applications, this may or may not be an issue because of the increased cross-sectional area.

The feed screw mass flow rate was calibrated after the bridging problems were solved and the control program was finalized. The mass flow rate was too high at the lowest motor controller setting. There were two solutions: One was to replace the gear box, and the other was to replace the chain and gear drive. The gear drive was the easier and less costly approach. CE-CERT replaced the original 22-tooth gear with a larger 72-tooth gear (Figure 4-14a). The feed was reduced from 40.8 kg/hr. to 19.5 kg./hr. at a command signal of 20 Hz.

The feed calibration was performed after solving all of these feed system problems. CE-CERT calibrated the feed system using a 2-minute sample time because the theoretical residence time (Dong and Cole, 1996) is two minutes (Figure 4-15). The average mass flow over the sample population was 21.8 kg/hr., with a standard deviation of 1.7 kg/hr. The two-minute mass flow uncertainty was estimated to be 3.3 kg/hr., or 15%. The mass flow is dependent on the bulk density and the packing in the screw flutes, both of which could change without warning. In addition, calibrating the screw feed before and after a test requires labor to remove the bottom reactor piece from the tower. To satisfy the measurement plan for biomass feed rate, load cells were installed under each leg of the hopper bin (Figure 4-14). The change in weight before and after a test will be used as the measure of carbon fed into the system. Unfortunately, the load cells are accurate only for tests greater than 24 hrs. Appendix IV contains details of the sample plan.

The screw feed also was calibrated at different sample time intervals and at different meter bin levels. During 1-minute and 60-minute samples, the uncertainty was estimated to be 30% and 10% respectively. Sample times greater than 60 minutes did not reduce the uncer-
tainty any further. One-minute flow samples were taken at high, medium, and low meter bin levels (LS-849). The effect due to meter bin level was not significant enough to affect the mass flow into the reactor.

CE-CERT noticed the bottom feed valve was binding on the drive arm after operating the feed system for over 100 cycles. The valve was removed and broken down to investigate the problem. Rust damage was found on the journal bearing, which caused the drive to seize (Figure 4-16). CE-CERT purchased a new drive system and installed it. There have been no more problems with the feed valve since. The manufacturer recommended operating the valves on a routine basis to prevent rust buildup. CE-CERT implemented a preventive maintenance cycle to operate all large pneumatic valves during down times.

4.3 Process Gas Supply

Gas Flow Rate Measurement

Six gas flow rates must be measured accurately as described in the sample plan (Appendix IV). The original design was to use the manufacturer’s specification sheet for flow calculations. CE-CERT determined that the manufacturer’s estimation had an uncertainty of 50% minimally. CE-CERT performed a 5-point calibration of the orifice plate and flange tap systems over the range of expected flow conditions using a dry gas meter suitable for each flow following ASME standards as listed in the Sample Plan (see Appendix V for calibration curves and tables). Each calibration was completed with nitrogen at pressure and temperature. A summary of the measured uncertainties is listed in Table 4-1.

According to the ASME standard, the gas flow calculations assume temperature, pressure, and differential pressure are being recorded at each orifice plate. The original design did not specify individual temperatures or pressures, but assumed a common temperature and pressure for all the gases. CE-CERT made the modification by adding 5 temperatures and 5 pressure measurements for CH₄, H₂, CO₂, CO, and the effluent.

The orifice flanges were made of plain carbon steel without any protective coatings. Steel rusts quickly and could damage the orifice plates in a short period. To prevent the rust contamination, the flanges were plated with a 0.0005 in (12.7μm) gold erudite plating.

Steam Flow Metering

An orifice type flowmeter was specified by the design for measurement of steam flow. This is not a recommended sensor type for measuring steam flow because the density can significantly change across an orifice when water condenses due to pressure drop. The solution was to use a 500 mL burette at the inlet to the steam pump and to spot-check the flow of the steam pump during testing. The steam pump is a constant flow pump and should not vary from check to check. Actual system uncertainties will be estimated once more experience is gained with reactor operation. Preliminary flow tests show this method is reliable. During preliminary testing CE-CERT noticed a problem maintaining flow with a back pressure. More investigation is required.

4.4 Hot Gas Filter

Nitrogen Pulse Heater

Preheated nitrogen is used to pulse-clean the high-temperature ceramic filter. The filter element normally operates at 538 °C under steady-state gasification. During gasification the filter elements need to be pulse-cleaned hourly with a high-pressure pulse of nitrogen. The preheated nitrogen is necessary to prevent filter element damage due to thermal shock when pulse-cleaned. Originally, the nitrogen was to be preheated using waste gas from the effluent heat exchanger (HX-107). This

![Figure 4-16. Everlasting feed valve drive and assembly: (a) damaged drive torque arm; (b) valve assembly showing seals and valve seat; (c) valve mechanism after reassembly.](image)
was modified when the steam heat exchangers were found to be undersized. The steam is preheated in both HX-107 and HX-108. The solution was a 2 kW band heater controlled by the Cyrano program. The vessel was designed to be 4 feet long by 4 inches in diameter and holds a total of 20 scf when pressurized to 800 psi (Figure 4-17). The pulse heater does operate as expected, but there are problems with the pulsing valves as explained in the subsection below on Solenoid Valves.

4.5 Controls

The controls were started by Arcadis and finalized by CE-CERT. This section covers the changes made to the controlling program by CE-CERT.

Bed Height Calculation

Knowing the bed height is necessary to help estimate optimum residence time and understand how the gasifier is performing (Dong and Cole, 1996). Arcadis ran a cold flow model simulating the Hynol facility to determine the expected bed height at different flows and bed material. Analysis by Arcadis concluded that the bed height does not follow this simulation, but rather follows a more theoretical approach (Dong, 1998, personal communication). CE-CERT implemented both methods in the control program, allowing the operator to choose based on operational experience.

Flow Calculation

Flow calculations for CH₄, CO₂, CO, N₂, and H₂ are programmed into Cyrano following the methods outlined in ASME P.C. 19.5 (see Sample Plan, Appendix IV). The controller samples gauge pressure, gas temperature and differential pressure at the orifice plates for each of the gases. After each signal is measured the controller converts the units to absolute for flow calculation. The controller then calculates a flow rate based on the square root of the pressure, temperature, differential pressure and molecular weight. The mass based flow rate (normal cubic meters per hour) is then calculated using the orifice calibration constant at a standard condition of 15.5 °C and 1 atm. The gas flow rate is logged once a minute and updated every five seconds on the computer screen for start up, steady state and shut down operations.

Equations were programmed into Cyrano for the calculation of gas flow for CH₄, CO₂, CO, N₂, and H₂. The program samples the differential pressure, temperature, and pressure at the orifice and calculates the normal differential pressure based on actual temperatures and pressures. The following equations show the calculations for the hydrogen orifice:

\[
Q_e = 12.892 \Delta P_a^{0.4291}
\]

where

\[
Q_e = \text{Normalized calibration flow at 3.2 MPa and 519.67 °R)}
\]

\[
\Delta P_a = \text{Actual pressure differential across the orifice plate.}
\]

And the final flow is calculated by correcting for the local temperature, pressure, and specific gravity:

<table>
<thead>
<tr>
<th>Entering</th>
<th>Designed Flow (kmol/hr)</th>
<th>Flow SI (Nm³/hr)</th>
<th>Flow ENGL (scfm)</th>
<th>Standard Error (scfm)</th>
<th>Uncertainty Measured %</th>
<th>Expected DelP* in H₂O</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
<td>0.1612</td>
<td>3.61</td>
<td>2.12</td>
<td>0.00622</td>
<td>1.00%</td>
<td>1.042</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.0983</td>
<td>2.2</td>
<td>1.29</td>
<td>0.0115</td>
<td>1.00%</td>
<td>1.533</td>
</tr>
<tr>
<td>CH₄</td>
<td>0.0449</td>
<td>1.01</td>
<td>0.592</td>
<td>0.0066</td>
<td>1.50%</td>
<td>0.45</td>
</tr>
<tr>
<td>H₂</td>
<td>1.771</td>
<td>39.7</td>
<td>23.4</td>
<td>0.462</td>
<td>2.00%</td>
<td>4.00</td>
</tr>
<tr>
<td>N₂</td>
<td>0.1711</td>
<td>3.83</td>
<td>2.25</td>
<td>0.0057</td>
<td>1.00%</td>
<td>4.67</td>
</tr>
<tr>
<td>H₂O</td>
<td>0.2859</td>
<td>11.5</td>
<td>6.77</td>
<td>n/a</td>
<td>in progress</td>
<td>n/a</td>
</tr>
</tbody>
</table>

Exiting effluent 3.056 68.45 40.3 0.240 1.00% n/a

* Flow calibrations are based on a STP pressure =14.696 psi (1 atm) and temperature =60 °F (15.55°C).
where

\[ Q_a = \frac{T_c P_a SG_a}{\sqrt{P_c T_a SG_c}} \]

(9)

\[ Q_a \] = Actual flow rate in (scfm) corrected for pressure, temperature, and specific gravity.

\[ T_c \] = Normalized calibration temperature (519.67 °R).

\[ P_a \] = Actual pressure at the orifice flange tap in psi units.

\[ SG_a \] = Actual specific gravity of the gas flowing through the orifice.

\[ P_c \] = Normalized calibration pressure (3.2 MPa).

\[ T_a \] = Actual temperature of the gas upstream of the flange tap °F units.

\[ SG_c \] = Specific gravity of calibration gas (Hydrogen was used for \( H_2 \) flow and nitrogen was used for \( CO, CO_2, N_2, \) and \( CH_4 \) flows).

**Burner System**

Since the BMS was not functional, all the controls for this portion were removed from the control program and the program was modified to reflect the new safety interlock and process requirements to start, monitor, and stop the burner. The program prevents operation of the burner if an alarm is activated. Also, the controller calculates the air-to-fuel ratio of the gas going through the burner.

**Water and Steam Pump**

No remote starting systems were designed for the steam pump and water pump. They are necessary to start the steam pump at a desired process temperature. Easy starting of the steam pump and cooling water pumps is necessary for the safe operation of the facility. The starting of these pumps is remotely controlled in the control room by either manual or automatic modes.

**Nitrogen Pulse Heater**

The nitrogen pulse preheater controls include manual and automatic controls. The start, stop and desired set point are controlled by the computer or manually at the heating element. A proportional, integral and derivative (PID) controller was programmed in Cyrano, which ramps up the temperature to the desired set point.

**Feed System**

The feed system automation was programmed in OPTO 22. There were many revisions to control logic due to the problems mentioned earlier. The final flow block diagram is shown in Figure 4-18. First, the feed system is manually started and the feed system checks the level meter (LS-498). If the level is greater than 6 mA, the feed cycle is initiated; otherwise, the program waits for the level to drop below 6 mA. Once the cycle starts, the
nitrogen purge flow is checked. If the flow is too low (less than 2 inH₂O) the operator is instructed to open FCV-404 before the cycle will continue. The nitrogen flow is used to keep positive pressure in the meter bin, which prevents heat from flowing into the bin. When the lower feed valves open, there is a brief inrush of gas into lockhopper LH-801. Without nitrogen flow, hot gases from R101 would be drawn into the meter bin, eventually causing mechanical and process damage.

Once the N₂ flow is set >2 inH₂O, the additives are combined with the biomass in the precharge hopper. The program then checks to make sure the bottom valve FV-840 is closed before depressurizing. If the valve is not closed, the controller closes the valve. Each time FV-840 or FV-839 is opened or closed, the time out and stop and alarm blocks are activated. If the valve does not do what it was supposed to, an alarm is sounded, the operator sees an error message, and the plant is put in standby mode.

After successful indication that FV-840 is closed, the lockhopper LH-801 is depressurized. Once the pressure in LH-801 is less than 34.5 kPa, the top feed valve (FV-839) is opened. Biomass drops into LH-801 while a pulse air blast cleans out the neck area #2 (see Figure 4-10). The valve (FV-839) is closed and pressurized through FV-839 cleanout port connection. Once the pressure is ≥ reactor pressure (PT-804), the bottom valve is opened. Biomass drops into the meter bin. The feed level should increase by at least 0.5 mA. If the level remains unchanged, the operator receives an error message, and the feed system is terminated until reset.

After biomass drops into the meter bin, a second nitrogen blast cleans out the bottom valve (FV-840). The valve closes and the cycle checks the level. If the level is greater than 13 mA (high level), the cycle is ended. Otherwise the level is still less than the high level mark, and the cycle repeats itself after timer KS-840 expires (90 seconds).

Each feed cycle takes approximately 90 seconds from start to stop. Between each feed cycle there is a 90-second wait time to fill the precharge hopper. During steady state testing, it was estimated the feed cycle takes 15 minutes to fill the meter bin with a 5- to 10-minute wait between ended cycles.

### 4.6 Cooling System

The cooling system includes a cooling tower, cooling pump, and an effluent gas heat exchanger. CE-CERT provided a drip type naturally aspirated water cooling tower. It is well overdesigned for the process and is expected to be adequate during steady-state gasification testing. The water pump provided for the project was only ½ hp (186 W) and undersized for the flow needs of the process. CE-CERT installed an available 4 hp (3 kW) cooling pump that has been working error-free.

The cooling system also has a 24.4 m heat exchanger that extends from the top of the tower to the flare stack. It was designed to bring the final effluent temperature down to 93.3 °C and to preheat the steam before it enters the main heat exchanger vessel. (Figure 4-20).

#### 4.7 Solenoid Valves

The original design specified solenoid operated gate valves designed and manufactured by Atkomatic. During shakedown, five of the fifteen valves have been troublesome. Two have been replaced and currently the other three are being bypassed. CE-CERT recommends replacing the valves with pneumatic valves because of reliability and longevity.

#### 4.8 Exit Flare Stack

The flare stack came with a 2.1 m extension and no mounting hardware. According to local code, the flare had to be 4.6 m above the working surface. CE-CERT modified the system with a 2.5 m extension and guide-wire support anchored to the concrete pad (Figure 4-21). The electrical controls were installed by CE-CERT. The flare starts manually, but has an automatic shutdown and external alarm to the operator for safety. The flare stack has been successfully operated.

### Main Heat Exchanger HX-205

The purpose of the main heat exchanger is to preheat inlet gases before they enter R101. The heat exchange takes place between the hot effluent exit gas and the cool simulated recycle inlet gas. The original heat exchanger was made by Arcadis from porous ceramic block. Ceramic was chosen due to expected 760 °C effluent gas temperatures.

The problem was that the ceramic heat exchanger would allow gas to transfer from the cool side (higher pressure) to the hot side (lower pressure). The pressure difference is the drop in pressure in the system. The pressure drop is determined from the resistance to flow through the heater, burner, distributor plate, three meters of biomass, a cyclone, and the high pressure filter before it becomes
the hot effluent gas. The pressure drop is expected to be around 34.5 kPa. A test rig was set up (Figure 4-21) to evaluate the heat exchanger performance. The goal was to measure how quickly the gas would leak into the process side with one end of the heat exchanger plugged and the other side pressurized to 34.5 kPa. The leak rate was less than ½ second thus requiring HX-205 to be redesigned and installed. The new design is 55 ft of 12.7 mm x 0.889 mm Haynes Alloy 630 tubing coiled with a radius 127 mm and a gap of 1.59 mm between each pair of coils. The heat exchanger will be evaluated once steady state gasification is achieved.

4.9 Sample System

The sample system is shown in Figure 4-22 and described in detail in Appendix IV. The goal of the sample system is to measure in real time CO, CO₂, CH₄, and H₂ while collecting deposits on four filter assemblies and removing water through a 689.5 kPa, 1.67 °C cooling system. The expected moisture of the analyses is 2000 ppm. From the composition and flow data, it will be possible to characterize when the plant reaches steady state and to calculate approximate carbon conversion efficiency, thermal efficiency and mass balance.
Figure 4-19. Effluent gas heat exchangers (HX-109, 110a, and 110b), a tube within a tube.

Figure 4-20. (a) Flare stack, extension, and guide-wire support; (b) electrical controls and valves.
Figure 4-21. (a) Heat exchanger HX-205 leak rate test setup; (b) HX-205 redesign.

Figure 4-22. High-pressure sample system: (a) high-pressure impingers submersed in ice/salt mixture (-20.6 °C) during testing; (b) array of pressure regulators prior to going into dry gas meters and after impingers; and (c) calibration gases for continuous analyzer.
5. Preliminary Results

5.1 Test 1: Air Gasification

Test Goals: The goals of tests 1a and 1b were to demonstrate reliable operation of the burner, feed system, heater, and bed height pressure drop. Secondly CE-CERT also hoped to achieve 1472 °F (800 °C) bed temperatures to demonstrate optimum biomass gasification.

The locations of thermocouples and valves listed below are shown in the Diagrams provided in Appendix VII of this report.

Result Summary 1a

The burner was unreliable at first, but once burner temperature TE-020 was > 1500 °F (816 °C) the burner was easy to stop and start. The feed system ran reliably except for a bridging problem in FV-840. The bridging problem caused a gas leak through FV-840 and FV-839. The leak caused heat to flow into the feed system meter bin, thus overheating feed screw TE-808. At the same time the process filter (F-104) clogged; this was indicated by a large (>50 psi, 344 kPa) pressure differential between inlet and exit pressures (PT-030 - PT-823). As a result of these problems, the ideal bed temperatures and bed height pressure drops were not achieved and gasification was not performed.

Test Setup and Operation

12/7 Primed R101 bed with 26 liters of sand (static bed height of 1.3 m). Tried to preheat the reactor with the electric heater (H-036) to 204 °C as necessary to prevent water condensing during burner operation. The heater capacity was not sufficient to preheat the reactor to 204 °C; thus, the burner was started even though water condensation would occur inside R101. The burner would not start because of a problem with the spark rod igniter. Shutting down until burner operational.

12/10 Same static bed height. Tried starting burner, but modified spark rod igniter failed again. The method being used is not reliable. A new design is necessary. While fixing the spark system, CE-CERT modified the air and natural gas plumbing to make start-up safer and quicker.

12/16 Designed new spark system that worked reliably on the bench and in the reactor. Operated burner until TE-809 was 427 °C.

12/17 Cycled the feed system valves as a safety check prior to starting the feed system while waiting for the temperature at TE-809 to increase.

12/18 Operated feed system after TE-809 reached 427 °C. The feed system ran reliably except for a bridge problem in tank T-805. The bridging problem caused a gas leak through FV-840 and FV-839. The leak caused heat to flow into the feed bin, thus overheating the feed screw TE-808. At the same time the process filter (F-104) clogged, as indicated by a large (>50 psi) pressure differential between inlet and exit pressures (PT-030 - PT-823). As a result of these problems, the ideal bed temperatures and bed height pressure drops were not achieved. Shut down system and started purging with nitrogen.

12/19 Purged reactor with nitrogen.

End of Gasification Test 1a

The automation of the feed system was the main success from test 1a. Figure 5-1 shows a typical automation profile for the high-pressure feed system, as per the design modification. Once the level signal (LSL-849) goes below 6 mA, the feed cycle is started. First the lockhopper LH-801 is pressurized from ambient to reactor pres-
sure (PT-804) + 34.5 kPa. The excess pressure is used to help push the biomass into the meter bin (T-805). If the pressure is in excess of 68.9 kPa, the extra pressure will prematurely open the bottom valve (FV-840) and prevent the cycle from completing. Once the pressure (PT-847) in the lock hopper (LH-801) is 34.5 kPa greater than the reactor pressure (PT-840), the bottom valve opens after a determined delay and the biomass drops into the meter bin. The added biomass to the meter bin can be seen by the rise in LSL-849. The completion of a cycle is noted by PT-847 going from reactor pressure (PT-804) down to ambient pressures. Also note the constant drop in LSL-849 between cycles as biomass is constantly fed into the reactor (R101). The delay between cycles is necessary to fill LH-801 with biomass, kaolinite and sand.

Unfortunately, a bridging problem occurred at the biomass feed valves. The bridging was due to a human error, not a process control error. FV-427 was not opened as listed in the preliminary startup procedures. As a result of not opening FV-427, biomass collected on the back side of the lower feed valve (FV-840), preventing the valve from fully closing. Because the valve was not fully closed, gas leaked past the valve seat during each feed cycle. The gas leak through the valves caused a rapid overheating of the feed screw TE-808. Once the feed screw temperature exceeded 204 °C, the plant was shut down and the problem was investigated.

The filter also clogged during test 1a. According to the filter designer, after gasifying/combusting biomass the filters need to be purged with hot air to burn off any residual carbon. Purging with hot air would be possible by operating the burner for 2 to 3 hours after gasification. Because of the rapid increase in TE-808, the reactor was shut off and not brought down slowly with the burner. CE-CERT believes shutting off the reactor without purging the filters with hot air is what caused the filters to clog. To fix the problem, the manufacturer suggested burning off the residual carbon with air at 315 °C. CE-CERT rented an industrial propane heater and burned off the residual carbon from the filters. The heater was installed at the base of the filter with access through tank T-104. CE-CERT confirmed the filters were cleared by successfully running a simple pressure test. A filter cleaning SOP has been established for the future.

One of the more time-consuming tasks was to get a reliable spark at the center of an 457 mm pipe through a 6.4 mm hole inside a vessel at 7 atm. The original design as described in the Burner Management section solved the problems initially, but after extended use that modification became unreliable. During operation the spark rod and ground moved from thermal expansion and gas velocities. If the gap were more than 3.2 mm, the spark-er would not work. In addition, adjustments to the gap were made 61 cm into a dark 25.4 mm hole, which made it difficult to set the correct gap size. The modified
design had the spark rod and ground in the same housing. This design would have worked, but the high-voltage spark found an easier path through the insulating seal of the high-pressure pass-through. The next design took advantage of the same concept of providing the spark and the ground in the same unit, but with a better pressure seal. The high-pressure pass-through manufacturer had a special seal material available, but the cost was in excess of $5,000 each with a minimum order of 10. Instead, CE-CERT modified an off-the-shelf spark plug to serve as the burner spark system. It proved to be very reliable and simple to fabricate.

**Result Summary 1b**

The burner operated reliably every time. Ideal bed temperatures were achieved and air gasification was noticed, but with low CO and CH$_4$ concentrations. The electric heater was damaged again and the feed system was jamming consistently. After removing the burner spool piece, large agglomerations were found surrounding the feed system inlet to R101.

**Test Setup and Operation**

1/5/99 The bed is still primed with 26 liters of sand minus any losses. Turned on electric heater to 20%, 40% then 60%. Started burner after TE-020 was >400 °F (200 °C).

1/6/99 Turned feed system on once TE-809 reached 427 °C. Feed system set at 22/7 kg/hr., 1.13 kg/hr. kaolinite, 1.13 kg/hr. sand and an air flow set at 22.1 Nm$^3$/hr (sub-stoichiometric combustion by 1/3). The burner and nitrogen flows were also turned off. The reactor temperatures were very unstable. The feed system was jamming every 10 minutes. Tried adjusting the air flow to keep temperature stable. The feed system was toggled on and off every 5 to 15 minutes to clear jamming problem.

1/7/99 Because the natural gas storage capacity was below the needed supply pressure for the natural gas burner heating, the reactor was temporarily shut down to recharge the compressed cylinders to 24.8 MPa. Turned burner back on to attempt gasification again. Waiting for TE-809 to reach 427 °C.

1/8/99 Temperature of 427 °C was reached. Before starting up feed system, added sand to increase the bed height. Added an additional 20 liters of sand at 06:00. Now the static bed height is 2.5 m minus any losses.

1/9/99 All attempts to gasify were well below ideal. Shutting down reactor to investigate problems with the feed system and the reactor gasification area.

![Figure 5-2. Gasification product concentrations for test 1b, January 8, 1999.](image-url)
End of Gasification Test 1b

Even though ideal gasification was not achieved, there were some signs of gasification. Ideal gasification should yield CO and CH₄ percentages around 10 to 15%. The best results from test #1b were 2% CO, 0.5% CH₄, and 9% CO₂ (Figure 5-2). The low CO and CH₄ concentrations and the high CO₂ concentration indicate more combustion than gasification was occurring. One reason for the poor gasification could be the problem with agglomeration. The agglomeration plugged the entire reaction area, thus preventing proper fluidization.

The feed system operated with no bridging problems, but there was an operational problem to consider for future designs. The biomass that remains in the precharge has a tendency to get wet during idle months. Future designs need to accommodate weather protection, and operational designs need to purge out the biomass prior to shutdown. Our SOPs (Appendix VI) include five feed system cycles during shutdown and a pretest check for all chutes to be clean and free of any obstructions.

A second problem was noticed with the feed system. After about an hour of gasification, the main feed screw SF-806 motor would overload and stop. CE-CERT initially thought the problem was due to a mechanical interference with the feed screw and shaft housing. After removing the bottom vessel to inspect the feed area, large stonelike agglomerations were found (Figure 5-3). In Figure 5-3a, notice the blockage filling the entire reaction area of R101 just above the feed system. Figure 5-3b shows the agglomeration pieces that dropped out of R101, coming to rest on the burner B-037. Figure 5-3c and d show R101 and B-037 after being cleaned and ready for reassembly.

Operating bed temperatures were achieved during test 1b (Figure 5-4). Unfortunately, the problem in achieving steady-state gasification prevented stable uniform bed temperatures. Notice the reactor temperature (TE-809) on January 6 between 1:00 and 9:00 varied by approximately 315 °C. The sudden increases in temperature were noticed after toggling the feed system back on after it jammed. It is believed the agglomeration problem was a result of the fluctuating temperatures and/or a poor choice of bed materials. Local hot spots could have gotten hot enough to fuse the sand.

During maintenance checks CE-CERT found all the heating elements damaged. The damage was due to poor support for the heater elements. According to the manufacturer, the elements need to be loosely supported from horizontal movement every 304 mm. The type of support installed was rigid and in contact with the elements. The modified design (Figure 5-5) used advanced powder metal (maximum temperature 1482 °C) for a support rod and 10 ceramic disks with a 6.4 mm gap for the loose support. This system also failed, as discussed in test #2 results below.

Figure 5-3. Agglomeration results after air gasification tests at 5 atm and 13 scfm air with 50lb./hr. biomass. Bed temperatures reached 982 °C, but were not consistent.
Figure 5-4. Bed temperature profile from the gasification area for test 1 on January 6, 1999.

Figure 5-5. Installation of new elements and supports: (a) heater element assembly; (b,c) element installation into heater vessel (H-036); and (d) final assembly with loose packing to prevent shorting.
The agglomeration problem is thought to be due to either melting the bed material or an alkali formation with the sand fluidizing media and potassium and sodium in the biomass. Alkali formation is a common problem with coal and biomass gasifiers. If the problem is from alkali formation, one solution is to use an alkali getter such as aluminum oxide. Unnasch (1996) recommended using kaolinite for the Hynol process at 5% by weight with the feed system as designed into the controls. If the problem is due to a poor choice of bed materials, other sands could be investigated.

5.2 Test 2: Air Gasification

Test Goals: The goals of this test were to demonstrate reliable burner operation, heater operation, and to achieve operating bed temperatures. Once bed temperatures were achieved, CE-CERT would attempt to fluidize the bed and demonstrate hot ash removal cycles with a sand/kaolinite mix. After successful ash removal cycles were completed, CE-CERT would then attempt biomass gasification, with the bed primed with 2.5L sand and 0.5L kaolinite.

Results Summary

Reliable burner and heater operation were demonstrated. There was no damage to the heater elements. Bed temperatures of 800 °C were achieved. The ash removal cycles were successful at operating temperatures. The agglomeration problem still exists, but the severity was much less than in test 1a and 1b.

Test Setup and Operation

3/30/99
08:30 Started electric heater.
10:25 Started burner easily and operated until the reactor bed TE-809 reached 427 °C.

3/31/99
02:50 TE-809 at 427 °C. Completed 20 successful ash removal cycles. For each test, 2 liters of sand and 0.5 liters of kaolinite were added, but only 2 liters of mixture was removed from the ash cycles. Each ash cycle removed an average of 0.3 liters of mixture.
08:00 Shut down system to prime bed with 2.5 liters of sand and 0.5 liters of kaolinite, which represents a 152 mm static bed height. Slowly pressurized the system to 25 psig (172 kPa) and set the air flow to 0.3 Nm³/hr. Successfully demonstrated an ash removal cycle before starting the feed system. Ran the feed system for 30 seconds to add approximately 0.23 kg of biomass (Figure 5-6).

Figure 5-6. Gasification attempts, March-April 1999.
08:15 One aspect critical for gasification facilities is ash removal. The facility was designed with an upper and lower ash removal system. The system was designed to operate once per hour, removing 1.4 kg/hr char and 0.3 kg/hr sand. Completed 5 successful ash removal cycles.

12:20 Primed bed with 2 liters sand and 0.5 liters kaolinite. Slowly pressurized the system to 25 psig (172 kPa) and set the air flow to 0.3 Nm³/hr. Successfully demonstrated an ash removal cycle before starting the feed system. Ran the feed system for 1 minute to add approximately 0.45 kg of biomass (Figure 5-6).

12:45 Completed 5 successful ash removal cycles.

13:45 Purged process filter and primed natural gas and nitrogen storage tanks for a longer gasification run.

20:30 Primed bed with 2 liters sand and 0.5 liters kaolinite. Slowly pressurized the system to 25 psig (172 kPa) and set air flow to 0.3 Nm³/hr. Successfully demonstrated an ash removal cycle before starting the feed system. Ran the feed system for 20 minutes to add approximately 17 lb (7.6 kg) of biomass (Figure 5-6).

21:00 Noticed large pressure differential (>50 psig; 345 kPa) across the process filter. Pulled the filters and turned burner back on to provide hot air to clean out process filters.

22:30 Process filters pressure drop cleared.

4/1/99

02:13 Started burner to bring TE-809 back up to 427 °C.

02:35 TE-809 >427 °C. Slowly pressurized the system to 25 psig (172 kPa) and set the air flow to 13 scfm (22.1 Nm³/hr). Successfully demonstrated an ash removal cycle before starting the feed system. Ran the feed system for 10 minutes to add approximately 4.5 kg of biomass (Figure 5-6).

03:00 Temperature TE-809 flared up to 699 °C. Tried to perform an ash removal cycle, but it was not successful. The process filter also seemed to be clogged because of the large pressure differential across the filter.

05:00 The process filter was cleaned, but the ash removal system is still clogged. Shutting down to fix the problem.

14:00 Removed bottom valves to inspect ash removal passage. Small agglomerations were found inside the ash removal passage. Cleaned out passage and visually confirmed reaction chamber was free of large agglomerations.

End of Gasification Test 2

Test #2 confirmed the stable operation of the feed system, burner, and the electric heater. In addition, two new cycles were successfully demonstrated: the ash removal cycle and filter pulse-clean cycle. The burner was easily started after preheating with the electric heater. Once TE-809 was > 427 °C, over 20 successful ash removal cycles were performed. An ash removal cycle removes ash, sand and kaolinite from the bottom of the reaction zone through a 25.4 mm tube. The reactor was primed with 2.5 liters of sand and 0.5 liters kaolinite to give a static bed height of 152 mm. It would usually take eight cycles to remove all the sand and kaolinite, which is an average of 0.3 liters removed per cycle. One thing noticed about the ash removal cycles was 3 liters of mix was added, but only 2.5 liters of mix was removed. Because the kaolinite is so light in comparison with the biomass, it may be elutriating out of the biomass and coating the reactor walls. This doesn’t cause any operation problems, but it does prevent the kaolinite from absorbing alkali from the biomass during gasification. The next test, test #3, will be performed with limestone as the alkali getter.

Gasification was attempted at 08:00, 12:20, 20:30 and 02:20 on March 31 and April 1, 1999, as shown in Figure 5-6. The corresponding bed temperature and pressure drop profiles are shown in Figures 5-7 and 5-8. During the gasification runs, the feed system was set at 22.7 kg/hr. and air was added at a flow rate of 22.1 Nm³/hr to attempt sub-stoichiometric combustion. The 30-second and 1-minute tests at 08:00 and 12:30 showed no sign of temperature increase, but there was a significant increase in CO and CO₂. The 10-minute and 20-minute runs at 20:30 March 31 and 02:20 April 1 showed an increase in bed temperatures, and CO, CH₄ and CO₂ concentration. The CO, CH₄ and CO₂ concentrations were 15%, 4.8%, and 15% respectively at the peak performance. These results are better than previous tests, but still not high enough for achieving the expected results.
The ash removal process was successful before gasification, but there were problems after gasification. During the 20-minute and 10-minute gasification test, the ash removal passage was clogged and needed manual cleaning to break up the blockage. The blockage was removed and looked like small agglomerations similar to those in test #1. In addition, there was a sharp rise in pressure drop across the process filter that indicated clogging during these tests. The agglomeration problem may be solved by using a different alkali getter as performed in test #3, or using a different bed material.

During maintenance checks, CE-CERT found all the heating elements damaged except for one set. The center rod support failed due to excessive temperatures (Figure 5-9). The support rod was designed to operate in temperatures as high as 1204 °C, but to melt like it did the temperatures would have been in excess of 1482 °C (1480 °C). According to the manufacturer, the low density gases must not carry away the heat very well, leaving the support rod in a high-temperature zone not measured by the thermocouples. A third design (Figure 5-10) uses silicon carbide for the main support and similar ceramic spacers. The new silicon carbide support rod is designed to operate in temperatures as high as 1649 °C. To prevent future damage to heater elements, more conservative start-up and shut-down procedures are put in place. Direct element temperature is being monitored.

### 5.3 Test 3: Biomass Gasification

Test Goals: The goal of this test was to gasify biomass with hydrogen (hydrogasification) in a bed primed with 3 liters of limestone used as an alkali getter. The expected results were to eliminate the agglomeration problem. The agglomeration problem has been identified by an excess pressure differential across the process filter, causing solid blockage in the reaction zone. An additional goal was to maintain bed temperatures of 760 to 871 °C while hydrogasifying biomass.
Results Summary

The agglomeration problem still exists, but there is evidence that it may not be due to hydrogasification. There is a problem with maintaining the 760 to 871 °C bed temperatures, but this may be a result of the possible fluidization problems when the refractory failed in the burner spool piece.

Test Setup and Operation

May 4

12:00 Started burner and electric heater. The electric heater alone could not heat the bed up to 760 °C, so the burner was used in parallel.

May 5

3:00 Operated burner with excess air for 5 hours until TE-814 and TE-809 were >760 °C (Figure 5-11).

8:30 Turned burner off (burner temp TE-020, Figure 5-11) and started adding N₂ to purge air. Purging air is a necessary precaution before adding hydrogen to the reactor.

9:00 Primed the bed with 3 liters of limestone.

9:30 Pressured with nitrogen until a pressure of 30 psi was achieved.

10:20 Started adding hydrogen into the reactor. TE-809 has dropped to ~649 °C.

10:28 Set hydrogen flow to 23 scfm (1.8 kmol/hr); see PDIT-003 in Figure 5-11. Figure 5-12 is a detail from 10:20 to 11:05.

10:35 Turned on feed system after the hydrogen flow stabilized and set biomass flow to 22.7 kg/hr. Notice in Figure 5-12 the steady decline in TE 809.

10:28 Hydrogasified for 30 minutes until TE 809 dropped below 480 °C. Once the temperature dropped below 480 °C, the feed system was turned off. One positive result from this test is there didn't seem to be a increase in pressure differential across the process filter. Figure 5-8 shows the pressure profile. Notice the pressure differential (PD_R101) stayed below 34 kPa during the hydrogasification testing. During air gasification PD_R101 can rise as high as 345 kPa.

11:05 Started adding air (Air_Flow) to burn off the remaining biomass in the reactor. Minutes after adding air an increase in pressure differential

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Figure 5-11. Hynol temperature profile before, during, and after the hydrogasification test on May 5, 1999.
across the filter was indicated by a rise in PD_R101 (Figure 5-13). This is believed to be when agglomeration started. More testing is necessary to confirm this theory. The increase in pressure differential could also be explained by the increase in air flow. The pressure rise also may be due to the increased gas flow from air gasification.

11:28 Shut off the hydrogen flow (PDIT_003) and increased air flow (Air_Flow) from 11.9 Nm³/hr to 25.5 Nm³/hr to prevent TE-809 from dropping any further (high temperatures are necessary to completely burn off the biomass). Notice in Figure 5-11 the rise in TE-809 with the addition of more air. After about 15 minutes TE-809 starts to drop off, indicating the biomass has been consumed.

12:45 Turned burner on to cool down reactor slowly and to prevent tar buildup on process filters (recommendation by filter manufacturer). Startup of the burner can be seen by the rise in TE-020 (Figure 5-11) and the increase in the Air_Flow (Figure 5-13).

12:50 Pulse-cleaned high-pressure filter (no pressure differential change).

13:25 Pulse-cleaned high-pressure filter (small pressure differential change).

13:30 Shut down burner and continued purging the reactor with air and nitrogen.

14:25 Pulse-cleaned high-pressure filter (small pressure differential change).

15:05 Pulse-cleaned high-pressure filter (small pressure differential change).

15:45 Shut off air flow and nitrogen.

End of Hydrogasification Test

On May 13, 1999, the burner spool piece was removed to clear some hard blockage material found in the ash removal tube. After removing the burner spool piece, the refractory was discovered to be cracked and damaged. There were many agglomeration pieces on top of the distributor (Figure 5-14). Some pieces looked like damaged refractory and others looked like sand agglomeration. The limestone did not agglomerate, but leftover sand from previous tests could have accounted for the majority of the agglomeration.

A crack developed in the spool lining opposite the burner inlet area near TE-020. The steel shell around the port
was around 371 °C. After removing the piece, it was noted that the backup insulation was damaged and in some places missing from the top of the spool piece. The inner hot face refractory lining was not badly damaged but did have some minor cracking. The secondary lining (AP Green 19L) melted during operation thus intermediate temperatures must have exceeded 1038 °C (max rating for the 19L).

One of the reasons for the excessive heat damage to the secondary lining is a result of the locations of sensor ports. The sensor ports are holes drilled through the light weight secondary layer and hot face layer. The holes are necessary to measure internal temperatures and pressures. The ports are located in line with the air flow from the heater and from the secondary air. The problem with this location is heat follows the air flow path. The air flow is directly onto TE-020, PT-030 and TE-814. To achieve the 760 °C bed temperatures (TE-809), high gas flows were necessary. The high flow created a large pressure differential around the distributor plate causing a back pressure in the burner. The high back pressure could have allowed the heat flow to find an easier route around the distributor. Since air was directed at two sensor ports, the flow of heat melted the light secondary lining around the sensor. As the refractory melted more air started to flow through the new passage, until the flow made a direct passage around the distributor. Assuming the gas flow made a secondary route around the distributor this would reduce bed fluidization a key factor in gasification. The possible lack of gasification would explain the lack of gasification with hydrogen. More tests are necessary once the burner is repaired to confirm hydrogasification bed temperatures problems.

To fix the problem CE-CERT was more conservative with the choice of burner refractory materials. The hot face was changed from Green Clean 60 to Ultra GreenSR. The Ultra GreenSR has better thermal shock resistance and a maximum hot face temperature of 1871 °C. The secondary lining, Greencast19L (max temp 1038 °C) was replaced with Greencast45L (maximum temperature 1371 °C). These changes increased the safety margin for the insulating material, but it also doubles the heat loss through the burner spool piece. The increased heat loss will still allow the burner to preheat the reactor to operating temperatures as designed. Also, a secondary precaution was incorporated to prevent heat from seeping around sensors by using the same high temperature hot face material around all burner port fittings. This will locally raise the wall temperatures and heat losses, but the areas are small, making the overall heat loss negligible.

In addition to the damaged spool piece, a second crack was noticed on the bottom section of R101 near the feed system (Figure 5-14c). To prevent this crack from propagating, CE-CERT patched the refractory with high-temperature patch materials (maximum temperature of 1371 °C). This repair was completed on-site with out dismantling the vessels.
The inconel distributor was also damaged during the last gasification test. The center rod deformed under the excessive temperatures and blocked ash from passing through the exit tube. To improve the design, CE-CERT added extra stiffness to the tubular straight section and introduced secondary air through the Mogas valves (FV-858) to prevent overheating.

Figure 5-14. Agglomeration pieces (a) after removal of the bottom burner spool section; (b) refractory damage in the burner spool piece and top view of gas distributor; (c) refractory crack in the main reactor.
6. Discussion

6.1 Hynol Gasification Tests

One of the main problems of operating the Hynol reactor has been sand agglomerations plugging the reactor bed. Table 6-1 lists the elemental analysis from test #1a, #1b, and #3 deposits. The deposits were collected at the entry point of the feed system into the reactor (R101), and the filter ash was sampled at the base of the filter clean out (F104). Also included is an elemental analysis of the biomass ash and sand used for bed material (note that biomass is 1.2% ash by dry weight). The significant alkali metals are potassium oxide (K₂O) and sodium oxide (Na₂O). From the analysis of the original sand material, a combined 6% of the sand mass is alkali. Another interesting fact is 16.8% of the sand mass was unaccounted for. The analysis of the two deposits is similar to the analysis of the sand.

Alkali problems typically are a result of six months of continuous operation, not a few hours. From simple mass balance calculations it is obvious the agglomeration mass was too large to have come from the biomass. During test #3 only 9.1 kg of biomass was added. This amounts to 100 g of ash, and only 6% of the ash is alkali. The agglomeration in test 3 was about 680 g.

The fact that the agglomeration is similar to the composition of the sand and the fact that the formation occurred within minutes of beginning operation suggest the sand is being fused. The glass making industry uses high alkali sand because the melting point is reduced, thus lowering operating costs. This suggests the agglomerations are a result of the high temperatures in the reactor and use of a sand with a low melting temperature. It could be possible that there are hot spots in the reactor that are not measured by the temperature probes. Additional temperature sensors would be useful in determining whether this is the case.

In test #1 the bed material was sand, but in test #3 the bed material was limestone. Figure 5-14a shows the analyzed agglomeration sitting on a pile of limestone. The explanation for the same deposit analysis in test #1 as in test 3 is that there was some sand in the reactor prior to adding the limestone. The sand was probably left over from test 2. The reactor was cleaned of all sand while the burner spool piece was being repaired.

One solution to prevent fusing the sand in the reactor is to change to a high silica (SiO₂), high alumina (Al₂O₃)

Table 6-1 Elemental analysis for biomass ash, sand, kaolin, and the formed deposits from tests 1a, 1b, and 3.

<table>
<thead>
<tr>
<th>Sample ID</th>
<th>SiO₂</th>
<th>Al₂O₃</th>
<th>TiO₂</th>
<th>Fe₂O₃</th>
<th>CaO</th>
<th>MgO</th>
<th>Na₂O</th>
<th>K₂O</th>
<th>SO₃</th>
<th>P₂O₅</th>
<th>Cl</th>
<th>CO₂</th>
<th>Total</th>
<th>Other</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fuel Wht Oak</td>
<td>36.6</td>
<td>9.98</td>
<td>1.15</td>
<td>4.32</td>
<td>20.3</td>
<td>5.81</td>
<td>4.43</td>
<td>8.07</td>
<td>3.61</td>
<td>1.43</td>
<td>0.07</td>
<td>0.019</td>
<td>95.8</td>
<td>4.2</td>
</tr>
<tr>
<td>Kaolin</td>
<td>41.5</td>
<td>37.8</td>
<td>2.17</td>
<td>1.02</td>
<td>0.03</td>
<td>0.06</td>
<td>0.21</td>
<td>0.17</td>
<td>0.06</td>
<td>0.18</td>
<td>0.02</td>
<td>0.02</td>
<td>83.2</td>
<td>16.8</td>
</tr>
<tr>
<td>Sand</td>
<td>69.9</td>
<td>10.1</td>
<td>0.38</td>
<td>0.67</td>
<td>1.56</td>
<td>0.09</td>
<td>2.31</td>
<td>3.48</td>
<td>0.12</td>
<td>0.05</td>
<td>0.01</td>
<td>0.03</td>
<td>88.7</td>
<td>11.3</td>
</tr>
<tr>
<td>Deposit 1/99</td>
<td>71.3</td>
<td>18.3</td>
<td>0.77</td>
<td>1.35</td>
<td>2.67</td>
<td>0.29</td>
<td>2.73</td>
<td>3.61</td>
<td>0.07</td>
<td>0.18</td>
<td>0.01</td>
<td>0.14</td>
<td>101.4</td>
<td>-1.4</td>
</tr>
<tr>
<td>Deposit 5/99</td>
<td>70.0</td>
<td>17.7</td>
<td>0.73</td>
<td>4.88</td>
<td>3.02</td>
<td>0.29</td>
<td>2.36</td>
<td>3.64</td>
<td>0.08</td>
<td>0.15</td>
<td>0</td>
<td>0.02</td>
<td>102.9</td>
<td>-2.9</td>
</tr>
<tr>
<td>Filter 1/99</td>
<td>29.2</td>
<td>21.0</td>
<td>0.87</td>
<td>2.06</td>
<td>1.63</td>
<td>0.82</td>
<td>0.46</td>
<td>0.87</td>
<td>0.41</td>
<td>2.4</td>
<td>0.06</td>
<td>2.98</td>
<td>62.8</td>
<td>37.2</td>
</tr>
<tr>
<td>Filter 5/99</td>
<td>38.4</td>
<td>24.1</td>
<td>1.17</td>
<td>4.07</td>
<td>1.16</td>
<td>0.35</td>
<td>0.6</td>
<td>0.86</td>
<td>0.59</td>
<td>1.48</td>
<td>0.05</td>
<td>0.05</td>
<td>72.8</td>
<td>27.2</td>
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<tr>
<td>Investo Cast**</td>
<td>52.1</td>
<td>42.2</td>
<td>2.09</td>
<td>0.35</td>
<td>0.02</td>
<td>0.02</td>
<td>0.10</td>
<td>0.10</td>
<td>n/d</td>
<td>n/d</td>
<td>n/d</td>
<td>n/d</td>
<td>97.0</td>
<td>3.0</td>
</tr>
</tbody>
</table>

* Performed by Hazen Research, Inc. using ASTM D2795.
** Investo Cast 50 specification from Ione Minerals MSDS.
and low alkali concentration sand. Investo Cast 50 is an excellent fluidizing bed material (Table 6-1). Investo Cast 50 is made up of 94% silica and alumina. Only 0.2% is from the alkali family. Investo Cast comes in five sizes ranging from 0.2 mm to 1.5 mm. InvestoCast 50 has a mean particle diameter of 0.34 mm, which is small enough to fluidize at the designed gasification velocities (0.3 m/s) and large enough to prevent elutria-tion (Table 6-2). This new bed material will be used in test 4, planned for December 1999.

Table 6-2. Investo Cast 50 size distribution from Lone Minerals.

<table>
<thead>
<tr>
<th>Size (mm)</th>
<th>Weighted %</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.59</td>
<td>0.0265</td>
</tr>
<tr>
<td>0.42</td>
<td>0.1638</td>
</tr>
<tr>
<td>0.3</td>
<td>0.1065</td>
</tr>
<tr>
<td>0.22</td>
<td>0.0385</td>
</tr>
<tr>
<td>0.15</td>
<td>0.0053</td>
</tr>
<tr>
<td>Mean</td>
<td>0.3406 mm</td>
</tr>
</tbody>
</table>

* Investo Cast 50 specification from Lone Minerals MSDS.

Although InvestoCast 50 would add to the gasification operating cost, it is expected that fluidization can be maintained without the sand bed material. CE-CERT is using the bed material to help start the gasification process. Once gasification has been achieved, the ash and char would become the fluidized media. Tests will be performed to show that this theory is correct.

The analysis on the white oak biomass agrees with other analyses. Table 6-3 lists several biomass Ultimate analyses. Notice the carbon, oxygen, and hydrogen percentages all agree within a few percent. The low ash content of the white oak fuel is an advantage for our process. The white oak fuel should perform well once operation is successful. The ash elemental analysis of the biomass is also similar to other fuel types (Table 6-4).

The hydrogasification test did show signs of hydrogasification during the 20-minute test before bed temperatures started to drop below 482 °C (Figure 5-12). Based on the peak value of effluent concentration, the carbon conversion efficiency can be estimated (Table 6-5). The calculation requires a knowledge of the total mass flow at the effluent. Unfortunately, the effluent flow is based on volume, not mass. Volumetric flows depend on gas density, which is a function of composition. In addition, the condensate trap before the flow meter was not operating correctly, so water condensate was flowing through the flow meter. This would affect the overall effluent flow result. Taking all these errors into consideration, it is possible to bound the mass flow in the effluent and calculate the carbon conversion efficiency.

Table 6-3. Fuel ultimate analysis comparison*.

<table>
<thead>
<tr>
<th>Sample ID</th>
<th>Water</th>
<th>Carbon</th>
<th>Hydrogen</th>
<th>Oxygen</th>
<th>Nitrogen</th>
<th>Sulfur</th>
<th>Ash</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fuel Wht Oak</td>
<td>7.2</td>
<td>49.7</td>
<td>5.46</td>
<td>43.32</td>
<td>0.37</td>
<td>0.03</td>
<td>1.12</td>
</tr>
<tr>
<td>Furniture Wst</td>
<td>12</td>
<td>49.87</td>
<td>5.91</td>
<td>40.29</td>
<td>0.29</td>
<td>0.03</td>
<td>3.61</td>
</tr>
<tr>
<td>Urban Wood</td>
<td>37</td>
<td>51.44</td>
<td>5.67</td>
<td>38013</td>
<td>0.41</td>
<td>0.03</td>
<td>4.32</td>
</tr>
<tr>
<td>Alder Fir</td>
<td>52</td>
<td>51.02</td>
<td>5.8</td>
<td>38.54</td>
<td>0.46</td>
<td>0.05</td>
<td>4.13</td>
</tr>
</tbody>
</table>

* White Oak is reported by Hazen Labs following ASTM D3172 the other fuels are from Miles et al. (1998) alkali deposit survey.

Table 6-4. Fuel ash elemental analysis comparison.

<table>
<thead>
<tr>
<th>Sample ID</th>
<th>SiO₂</th>
<th>Al₂O₃</th>
<th>TiO₂</th>
<th>Fe₂O₃</th>
<th>CaO</th>
<th>MgO</th>
<th>Na₂O</th>
<th>K₂O</th>
<th>SO₃</th>
<th>P₂O₅</th>
<th>Cl</th>
<th>CO₂</th>
<th>Total</th>
<th>Other</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fuel Wht Oak</td>
<td>36.64</td>
<td>9.98</td>
<td>1.15</td>
<td>4.32</td>
<td>20.3</td>
<td>5.81</td>
<td>4.43</td>
<td>8.07</td>
<td>3.61</td>
<td>1.43</td>
<td>0.07</td>
<td>0.019</td>
<td>95.8</td>
<td>4.2</td>
</tr>
<tr>
<td>Furniture Wst</td>
<td>57.62</td>
<td>12.23</td>
<td>0.5</td>
<td>5.63</td>
<td>13.89</td>
<td>3.28</td>
<td>2.36</td>
<td>3.77</td>
<td>1</td>
<td>0.5</td>
<td>n/a</td>
<td>0</td>
<td>100.8</td>
<td>-0.8</td>
</tr>
<tr>
<td>Urban Wood</td>
<td>39.96</td>
<td>12.03</td>
<td>0.87</td>
<td>7.43</td>
<td>19.23</td>
<td>4.3</td>
<td>1.53</td>
<td>5.36</td>
<td>1.74</td>
<td>1.5</td>
<td>n/a</td>
<td>6.05</td>
<td>100.0</td>
<td>0.0</td>
</tr>
<tr>
<td>Alder Fir</td>
<td>35.36</td>
<td>11.54</td>
<td>0.92</td>
<td>7.62</td>
<td>24.9</td>
<td>3.81</td>
<td>1.17</td>
<td>5.75</td>
<td>0.78</td>
<td>1.9</td>
<td>n/a</td>
<td>1.85</td>
<td>95.6</td>
<td>4.4</td>
</tr>
</tbody>
</table>

* White Oak is reported by Hazen Labs following ASTM D3172 the other fuels are from Miles et al. (1998) alkali deposit survey.
The range of expected effluent flows is indicated in Table 6-6. Because the input flow is only hydrogen, the majority of the gas minus that identified by the NDIR analyzers should be hydrogen. Assuming all the unknown gas is hydrogen, a carbon conversion efficiency of 33% is the result. If there was more nitrogen than hydrogen (highly unlikely), the carbon conversion efficiency would only be 13%. In either case only the process conversion efficiency is lower than the expected 87%. One reason for the poor conversion is a result of low gasification temperatures. Dong and Cole (1996) recommend temperatures >800 °C, but CE-CERT operated at 600 °C. Future tests will be performed after the bed reaches 800 °C. CE-CERT also recommends replacing the effluent volumetric flow sensor with a true mass flow meter.

A thermal evaluation of the refractory material in R101 was performed based on the results of test 3. The evaluation was necessary to understand whether there were any unexpected heat losses during operation (Figure 6-1).

An unexpected heat loss would be evident by large temperature drops from one sensor to the next. The measure used is temperature per length between sensors. There are three physical sections to the reactor (Figure 6-2). The refractory is only insulated on the inside of the structural piping. External insulation was not possible due to temperature limitations of 204 °C on the steel surface. It is expected that section 1 should have highest value, section 2 should be low and section 3 should be medium. Section 1 should be much higher than sections 2 and 3 because of its high-density refractory.

Figure 6-2 identifies the seven locations of temperature sensors within the three sections. The heat loss between locations 2 and 3 is 3.4 times the heat loss between locations 3 and 4. The low 0.26 °C/in. (0.01 °C/mm) heat

<table>
<thead>
<tr>
<th>Component</th>
<th>Input Flow* (kmol/hr)</th>
<th>Product* (mole %)</th>
<th>Input Flow** (kmol/hr)</th>
<th>Product** (mole %)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biomass</td>
<td>22.7kg/hr</td>
<td>0</td>
<td>22.7kg/hr</td>
<td>0</td>
</tr>
<tr>
<td>CO</td>
<td>0.161</td>
<td>11.05</td>
<td>0</td>
<td>4.7</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.098</td>
<td>05.86</td>
<td>0</td>
<td>3.1</td>
</tr>
<tr>
<td>CH₄</td>
<td>0.044</td>
<td>17.87</td>
<td>0</td>
<td>2.8</td>
</tr>
<tr>
<td>H₂O</td>
<td>0.285</td>
<td>19.69</td>
<td>0</td>
<td>Not measured</td>
</tr>
<tr>
<td>H₂</td>
<td>1.771</td>
<td>40.08</td>
<td>1.8</td>
<td>Not measured</td>
</tr>
<tr>
<td>N₂</td>
<td>0.171</td>
<td>05.31</td>
<td>0</td>
<td>Not measured</td>
</tr>
<tr>
<td>Effluent</td>
<td>N/A</td>
<td>40.3 scfm</td>
<td>N/A</td>
<td>45/14 scfm</td>
</tr>
</tbody>
</table>

*Values from laboratory and theoretical analysis by EPA and Arcadis (30 atm).
**Preliminary test conducted at UC Riverside Hynol Facility (2 atm).
Figure 6-1. Temperature profile for test 3, from start up to shut down of the Hynol reactor.

Figure 6-2. Reactor section 1 is the burner; section 2 is the main gasification reaction zone; and section 3 is the reduced velocity free board zone. The circled numbers are temperature sensor locations.
loss in locations 5 and 6 (Table 6-7) is a result of not inserting the temperature sensor into the gas stream. Overall, the heat into the reactor equates with the heat loss by the reactor and flowing gases. Approximately 43kW of heat (burner + electric heater) was going in and an accounted 40 kW was going into heating the reactor gas volume, refractory losses, and the burner gases.

**Table 6-7. Heat loss as a function of length based on test 3 with the Hynol reactor.**

<table>
<thead>
<tr>
<th>Section</th>
<th>Length (in)</th>
<th>Temp/Len (°C/in)</th>
</tr>
</thead>
<tbody>
<tr>
<td>6-7</td>
<td>34</td>
<td>1.97</td>
</tr>
<tr>
<td>5-6</td>
<td>35</td>
<td>0.259</td>
</tr>
<tr>
<td>4-5</td>
<td>28</td>
<td>1.57</td>
</tr>
<tr>
<td>3-4</td>
<td>37</td>
<td>1.61</td>
</tr>
<tr>
<td>2-3</td>
<td>58</td>
<td>5.51</td>
</tr>
<tr>
<td>1-2</td>
<td>12</td>
<td>18.1</td>
</tr>
</tbody>
</table>

### 6.2 Facility Design and Construction

The major difficulties confronting this project were a result of the Hynol pilot-scale facility design. Improved design and improved peer-review of the design could have prevented many flaws that caused significant setbacks and delays later. Additionally, CE-CERT should have insisted on demonstration of successful operation before accepting certain components and subsystems.

There have been many success in this project. The feed system has been made to operate automatically without bridging or overheating. The modified electric heater has operated without too much attention, and the burner is able to preheat the reactor area up to operating temperatures as needed.

### 6.3 Data Quality

Since steady state hydrogasification was not achieved, overall material balance cannot be used as a general quality indicator of the test data. The quality indicator in this study relies on the precision of each of the individual measurements involved in the testing.

The measurements in the Hynol reactor tests include system pressure, reaction temperature, the flow rates of hydrogen, methane, air, and nitrogen, the biomass feed rate, and the composition of the biomass samples. All of these measurements were conducted in accordance with the data quality goals listed in the sample plan and QA plan (Appendix IV).

The ASTM standard methods were used for composition analysis of the biomass, fluidized bed material, char, and deposit formations from the Hynol facility. Multiple samples were analyzed. The results of the analysis met the data quality indicators listed in the QA plan.

Each of the orifice plate flow systems was calibrated with a DGM before and after testing. The performance of the orifice plate was stable over time and met the quality indicators of the QA plan.

The desired system pressure was maintained with a back-pressure regulator. The system pressure variation exceeded the target of the QA plan, due to the inability to reach steady state gasification and a flow capacity problem with the back pressure regulator. The flow capacity problem has been resolved. Future tests should meet the system pressure target of the QA plan assuming steady state gasification is achieved.

A graduated glass tube was installed for the steam flow metering. Unfortunately, due to unstable gasification, steam addition was not performed and therefore no indication of the quality of the data was recorded or measured.

Type K thermocouples were used to measure heater, burner, and reaction zone gas temperatures. All thermocouples were calibrated before installation into the reactor following ASTM methods listed in the sample plan. All reaction zone temperature probes were installed into the process by 25.4 mm. All other TCs were installed ~10 mm into the process. Depths of all TCs were confirmed during vessel assembly and repair by visual inspection.

The desired quality indicators for the Hynol facility are from steady state operation. Because steady state operation was not achieved, meaningful carbon conversions efficiency and mass balances were not determined and further testing is needed to evaluate hydrogasification of biomass materials.
7. Conclusions and Recommendations

7.1 Conclusions

1. Biomass feeding is automated and delivers 22.7 kg/hr to the high-pressure hydrogasification reactor at 10 atm and has been operated without material automatically at 30 atm.

2. Electric heater support version 3 is reliable and capable of preheating ambient inlet gases to 649 °C. Design version 3 uses silicon carbide as the support rod for 5 ceramic disk supports.

3. The reactor is capable of withstanding 5.2 MPa with a leak rate less than 10% over 24 hours.

4. The flow measuring orifice plates are calibrated to within ±2% of the actual flow conditions for each gas. Calibrations were completed at operating pressures using nitrogen and a calibrated dry gas meter. Because hydrogen is such a light gas its calibration was done at operating pressures using hydrogen gas and a dry gas meter.

5. During operation designed skin temperatures are below the maximum rated skin temperature of the pressure vessels.

6. Manual and automatic control of all valves and motors is possible using OPTO 22 software and hardware.

7. Data logging is successful using the OPTO 22 process control software and hardware package.

8. Burner spark system is reliable and has survived more than 100 hours of service. The final igniter design is a modified automotive spark plug as shown in Figure 4-2.

9. The burner is controlled by monitoring a thermocouple located in the flame area.

10. Sample system temperatures are below that necessary to capture high temperature alkali gases and tars. These compounds will be analyzed from the sample tube. Future designs should allow proper sampling to collect for alkali and tar formations.

11. Sample conditioning is accomplished through a series of high-pressure, high-temperature impingers. Successful operation of the sample system has been demonstrated.

12. Analyzer delay times from the reactor sampling to the analyzer are dependent on the sample pressure, sample length, and analyzer response. The delay due to pressure is the most significant. At 10 atm, the delay time from sample probe to sample path is about 10 minutes. The delay from the sample path to the analyzer is 15 seconds. The analyzer response time is about 0.5 seconds. The overall sampling delay is 10 minutes at 10 atm.

13. Air gasification has been performed with success, but have not been optimized.

14. Final refractory materials for B-037 are a two piece lining with a hot face Ultra GreenSR and a secondary lining using GreenCast 47. These materials are designed for maximum peak temperatures of 1871 °C and maximum operating temperatures of 1427 °C.

15. Agglomerations occurred due to high alkali concentrations in the fluidized bed material with bed temperatures of 700 °C or greater. Care needs to be used when selecting fluidized reactor bed materials.
16. Ash removal cycles are successful at ambient and high temperature. Agglomeration formation will block the ash removal passage.

17. Bed temperatures of 800 °C were easy to obtain with air gasification.

18. Bed expansion and pressure drop across the bed were indicated, but not optimized for gasification.

19. Optimum carbon conversion efficiency has not been realizable due to agglomeration problems during gasification and low bed temperature.

20. No efficiency data are available for the internal cyclone.

7.2 Recommendations

The hydrogasification reactor has the potential to be a viable way to convert biomass into a liquid fuel. The high carbon conversion efficiency with hydrogasification shows the Hynol process could be an economical and low-CO₂ producing method for methanol production. Future testing can succeed if problems with agglomeration and the feed system are solved. Table 7-1 presents recommended test parameters for future tests. Also the process should be analyzed for the actual exit gas quality to predict other design problems for phase 2 and 3 of the methanol production process.
Table 7-1. Test conditions to be used for CE-CERT Hynol test facility.

<table>
<thead>
<tr>
<th></th>
<th>Hynol UC Riverside Test 4 Nov 10,1999</th>
<th>Hynol Arcadis/EPA Simulations</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Temperature</strong></td>
<td>800°C (1472°F)</td>
<td>800°C (1472°F)</td>
</tr>
<tr>
<td><strong>Pressure</strong></td>
<td>8 atm (103 psig)</td>
<td>30 atm (442 psia)</td>
</tr>
<tr>
<td><strong>Solids Fast Residence Time (average)</strong></td>
<td>n/a</td>
<td>15 sec</td>
</tr>
<tr>
<td><strong>Solids Slow Residence Time (average)</strong></td>
<td>n/a</td>
<td>7.86 hr</td>
</tr>
<tr>
<td><strong>Gas Residence Time</strong></td>
<td>9 sec **</td>
<td>8 sec</td>
</tr>
<tr>
<td><strong>Superficial Velocity</strong></td>
<td>0.27 m/s **</td>
<td>n/a</td>
</tr>
<tr>
<td><strong>Gaseous Input Flow Rate</strong></td>
<td>1.26 kmol/hr (4.2kg/hr)**</td>
<td>2.53 kmol/hr **</td>
</tr>
<tr>
<td></td>
<td><strong>H2</strong></td>
<td>95% **</td>
</tr>
<tr>
<td></td>
<td>CH4</td>
<td>0% **</td>
</tr>
<tr>
<td></td>
<td><strong>CO</strong></td>
<td>0% **</td>
</tr>
<tr>
<td></td>
<td><strong>CO2</strong></td>
<td>0% **</td>
</tr>
<tr>
<td></td>
<td><strong>H20</strong></td>
<td>0% **</td>
</tr>
<tr>
<td></td>
<td><strong>N2</strong></td>
<td>5% **</td>
</tr>
<tr>
<td><strong>Biomass Input Flow Rate</strong></td>
<td>18.5 kg/hr (41 lb/hr) ****</td>
<td>22.7 kg/hr (50 lb/hr) ***</td>
</tr>
<tr>
<td></td>
<td><strong>C</strong></td>
<td>49.7 %wt</td>
</tr>
<tr>
<td></td>
<td><strong>H</strong></td>
<td>5.5 %wt</td>
</tr>
<tr>
<td></td>
<td><strong>O</strong></td>
<td>43.3 %wt</td>
</tr>
<tr>
<td></td>
<td><strong>N</strong></td>
<td>0.4 %wt</td>
</tr>
<tr>
<td></td>
<td><strong>H2O</strong></td>
<td>18.7 %wt</td>
</tr>
<tr>
<td></td>
<td><strong>Ash</strong></td>
<td>1.12 %wt</td>
</tr>
<tr>
<td></td>
<td><strong>Ash Exit Flow Rate</strong></td>
<td>1.14 kg/hr</td>
</tr>
<tr>
<td></td>
<td><strong>Sand Exit Flow Rate</strong></td>
<td>0.3 L/hr</td>
</tr>
<tr>
<td></td>
<td><strong>Kaolinite Exit Flow Rate</strong></td>
<td>n/a</td>
</tr>
<tr>
<td></td>
<td><strong>H/C biomass Ratio (by mass)</strong></td>
<td>0.432</td>
</tr>
<tr>
<td><strong>Hydrogasification Products</strong></td>
<td>1.99 kmol/hr (21kg/hr)**</td>
<td>3.056 kmol/hr (45 kg/hr)**</td>
</tr>
<tr>
<td></td>
<td><strong>CH4</strong></td>
<td>10.5% **</td>
</tr>
<tr>
<td></td>
<td><strong>CO2</strong></td>
<td>2.76% **</td>
</tr>
<tr>
<td></td>
<td><strong>CO</strong></td>
<td>14.0% **</td>
</tr>
<tr>
<td></td>
<td><strong>H2</strong></td>
<td>85.8% **</td>
</tr>
<tr>
<td></td>
<td><strong>N2</strong></td>
<td>3.27% **</td>
</tr>
<tr>
<td></td>
<td><strong>H2O</strong></td>
<td>10.6% **</td>
</tr>
<tr>
<td><strong>Temperature Profile/Distribution</strong></td>
<td>Reaction Dependent</td>
<td>Electric Heater Control</td>
</tr>
<tr>
<td></td>
<td>TE-809b</td>
<td>800</td>
</tr>
<tr>
<td></td>
<td>TE-809</td>
<td>770</td>
</tr>
<tr>
<td></td>
<td>TE-810</td>
<td>730</td>
</tr>
<tr>
<td></td>
<td>TE-811</td>
<td>700</td>
</tr>
<tr>
<td><strong>Pressure Differential Expected (Distrib only)</strong></td>
<td>−15 inH20</td>
<td>n/a</td>
</tr>
<tr>
<td><strong>Pressure Differential Expected (w/out Distrib)</strong></td>
<td>−15 inH20</td>
<td>n/a</td>
</tr>
<tr>
<td><strong>Bed Material</strong></td>
<td>InvestoCast 50</td>
<td>Sand *</td>
</tr>
<tr>
<td></td>
<td><strong>Particle Diameter</strong></td>
<td>0.34 mm (0.013in)</td>
</tr>
<tr>
<td></td>
<td><strong>Volume Added</strong></td>
<td>15 L</td>
</tr>
<tr>
<td></td>
<td><strong>Composition</strong></td>
<td>52% SiO2, 42%Al2O3</td>
</tr>
<tr>
<td></td>
<td><strong>Static Bed Height @ Start-up</strong></td>
<td>45.3 cm (18 in)</td>
</tr>
</tbody>
</table>

* Evaluation of Biomass Reactivity in Hydrogasification for the Hynol Process by Yuanji Dong and Edward Cole, EPA-600/R-96-071
** Calculated with Stanjen at operating pressure and temperature using the above inputs to the reactor (Constant T and P).
*** Hynol Process Evaluation by Borgwardt EPA-600/R-97-153
**** Laboratory Analysis of White Oak Biomass Hazen Labs dry basis except for 11.35 kg/hr biomass and 18.7% moisture.
References and Bibliography


Steinberg, M., and Dong, Y. (1994b) An Economical Process for Methanol Production from Biomass and


Appendices

I. Environmental Impact Report

II. Soils Investigation

III. Hynol Facility Control Panel

IV. Quality Assurance Project Plan

V. Calibration Curves and Tables

VI. Standard Operating Procedures

VII. Diagrams and Locations of Sensors