LEVELING INTERMITTENT RENEWABLE ENERGY PRODUCTION THROUGH BIOMASS GASIFICATION-BASED HYBRID SYSTEMS

by

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A thesis submitted to the Faculty and the Board of Trustees of the Colorado School of Mines in partial fulfillment of the requirements for the degree of Master of Science (Engineering).

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This material is based upon work supported by the
National Renewable Energy Laboratory
Hydrogen Technologies and Systems Center

NOTICE

ABSTRACT

The increased use of intermittent renewable power in the US is forcing utilities to manage increasingly complex supply and demand interactions. This paper evaluates biomass pathways for hydrogen production and how they can be integrated with renewable resources to improve the efficiency, reliability, dispatchability, and cost of other renewable technologies.

The two hybrid concepts analyzed involve co-production of gaseous hydrogen and electric power from thermochemical-based biorefineries. Both of the concepts analyzed share the basic idea of combining intermittent wind-generated electricity with a biomass gasification plant. Wind availability overlaps biomass resource availability, making the use of locally produced wind electricity for gasification feasible.

The systems were studied in detail for process feasibility and economic performance. The best performing system was estimated to produce hydrogen at costs ($1.67/kg) within Department of Energy targets ($2.10/kg) for central biomass-derived hydrogen production, while also providing value-added energy services to the electric grid. The proposed hybrid systems seek to either fill energy shortfalls by supplying hydrogen to a peaking natural gas turbine or to absorb excess renewable power during low-demand hours.

Direct leveling of intermittent renewable electricity production is accomplished with either (1) an indirectly-heated biomass gasifier, or (2) a directly-heated biomass gasifier. The indirect gasification concepts studied were found to be cost competitive in cases where value is placed on controlling carbon emissions. Carbon values of $34-40 per tonne of CO_2

equivalent (CO_2e) make the systems studied cost competitive with steam methane reforming (SMR) to produce hydrogen. However, since non-hybrid biomass to hydrogen plants were found to be more cost competitive than the hybrid concepts studied herein, some additional value must be placed on energy peaking or sinking for these plants to be economically viable.

The direct gasification concept studied is unlikely to be cost competitive in the near future. High electrolyzer costs make the hybridization difficult to justify. Based on a direct replacement of the ASU with electrolyzers, hydrogen can be produced for $0.19 premium per kilogram. Additionally if non-renewable electricity is used, the hybrid system is a net CO_2e emitter.

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LIST OF ACRONYMS

ACESA	American Clean Energy and Security Act
AEO	Annual Energy Outlook
ASU	Air Separation Unit
BCL	Battelle-Columbus
BIGCC	Biomass Integrated Gasification Combined Cycle
CAES	Compressed Air Energy Storage
CNG	Compressed Natural Gas
CO_2e	Carbon Dioxide Equivalent Emission
DOE	Department of Energy
EIA	Energy Information Administration
GHE	Greenhouse Emissions
GHG	Greenhouse Gas
GIS	Geographic Information System
GTI	Gas Technology Institute
HTS	High Temperature Shift Reactor
H2A	Hydrogen Analysis Tool
ICC	Integrated Combined Cycle
ISO	Independent Transmission System Operator
LTS	Low Temperature Shift Reactor
MCFC	Molten Carbonate Fuel Cell
NG	Natural Gas
NREL	National Renewable Energy Laboratory
PEM	Proton Exchange Membrane
POx	Partial Oxidation
PSA	Pressure Swing Adsorption
PV	Photovoltaic
RTO	Regional Transmission Organization
SC	Simple Cycle

SMR	Steam Methane Reformer
SOFC	Solid Oxide Fuel Cell
TPD	Tons per Day
US	United States
WGS	Water Gas Shift
ZnO	Zinc-oxide

ACKNOWLEDGEMENTS

I would like to thank both my advisor Dr. Munoz and my co-advisor Dr. Braun for their support and guidance throughout the last two years. Without Dr. Munoz's faith in me I would never have started this journey. He has pushed me to broaden my perspective and examine my biases. Dr. Braun has taught me how to attack a challenge. He has consistently pushed me to understand more about the problem at hand.

Through all the long nights, early mornings, and uncertainty my wife Agata has remained supportive. I couldn't have done this without her sacrifices. And thanks to my boys Matthew and Isaac (who arrived along the way) for making me smile at just the right times.

Thanks to Michael Penev and Christopher Kinchen of the National Renewable Energy Laboratory for their contributions to this research. Paul Denholm provided a much needed crash course in electricity market operations and market price information. Donna Heimiller's GIS work helped to quickly identify markets for analysis.

Finally, several industry representatives reviewed an initial draft of this report and provided feedback. A sincere thanks to Frank Novachek of Xcel Energy, Bryan Olthof of the Harris Group, Dr. Rakesh Radhakrishnan of Navigant Consulting Inc., Lori Smith Schell of Empowered Energy, and Ching-Jen Tang of GE Global Research.

CHAPTER 1
INTRODUCTION

Hydrogen can be produced domestically from a wide variety of resources, including biomass, wind, solar, natural gas and coal. In support of the Department of Energy's (DOE) program to develop hydrogen production technologies, this thesis investigates hybrid hydrogen production systems. While other fuels such as synthetic natural gas or Fischer-Tropsch fuels may be worth future consideration, the National Renewable Energy Hydrogen Group funded this research and therefore hydrogen fuel was the focus.

Of the domestic resources available for hydrogen production, biomass shows significant promise. Recent assessments have shown that in excess of 400 million tons of biomass are currently available per year in the United States (Milbrandt 2005), which could be converted to roughly 30 million tons of hydrogen by thermochemical processing.[1] Some estimates predict that as much as 1 billion tons of biomass could be available in the future with changes to land management and agricultural practices (Perlack et al. 2005). In addition to high availability, thermochemical plants provide many opportunities for system integration.

This thesis addresses the definition and evaluation of opportunities for combined production of hydrogen and electric power by combining biomass conversion with other technologies, including wind, solar, coal and nuclear. The goal is to identify systems that could increase the efficiency and reliability or decrease the cost of hydrogen production, or

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[1] This conversion efficiency is based on values obtained from the various gasification systems described and analyzed as part of this report.

improve the sustainability of hydrogen production from non-renewable resources.

1.1 Analysis Methodology

Due to the open nature of the project objective, analysis was performed in consecutively refining steps. Initially, brainstorming and literature review were used to narrow the scope of research. Then, detailed techno-economic analysis was performed on a subset of concepts.

This report summarizes the screening of technologies to identify promising hybrid systems for further analysis. The screening was performed in three main stages. First, a list of possible power technologies (both renewable and non-renewable) was generated and used for brainstorming. The results of the initial brainstorming were recorded in an idea matrix and then used to guide a literature review of existing research on hybrid systems. Many of the ideas resulting from initial brainstorming were well documented in literature, and additional ideas were added to the list of possibilities as a result of the review.

Several hybrid systems quickly rose to the top as good candidates for further investigation. These ideas were discussed extensively within the project team and then compared using a Pugh decision matrix in order to narrow the options for further analysis.

The chosen concepts were then analyzed from both a technology and economics standpoint. From a technological standpoint, hybridized system inputs and outputs were determined using ASPEN Plus thermochemical software. These values were input into the National Renewable Energy Laboratory (NREL) Hydrogen Analysis Tool (H2A) to determine the cost of hydrogen production for each concept.

1.2 Initial Brainstorming

During initial brainstorming, the goal was to identify as many synthesis opportunities as possible between biomass conversion and other power technologies for the production of hydrogen and power. Biomass conversion was assumed to include gasification, pyrolysis[2], direct combustion and fermentation for the purposes of brainstorming. These five biomass conversion options were then compared with a list of possible power technologies including traditional renewables such as wind and solar but also including non-renewable coal, nuclear and other technologies.

Initially, research was not focused specifically on thermochemical conversion pathways. Combustion was included in initial brainstorming, even though it does not allow for fuel production directly, because hybridization might allow for some hydrogen byproduct to be produced. Fermentation was used to describe any form of biological conversion of biomass to fuels.

An idea matrix was constructed with biomass technologies along the top and other power systems listed along the side. This table was used for brainstorming binary combinations between biomass processing and other technologies. The summary table is shown in Figure 1.1. Combinations were ranked numerically and the results are shown graphically (darker cells correspond to higher perceived promise). The color gradients shown are based on the more detailed Numerical Idea Matrix in Appendix A. Technologies that had large resource availability, low greenhouse gas emissions, and were relatively near term were favored in the ranking process.

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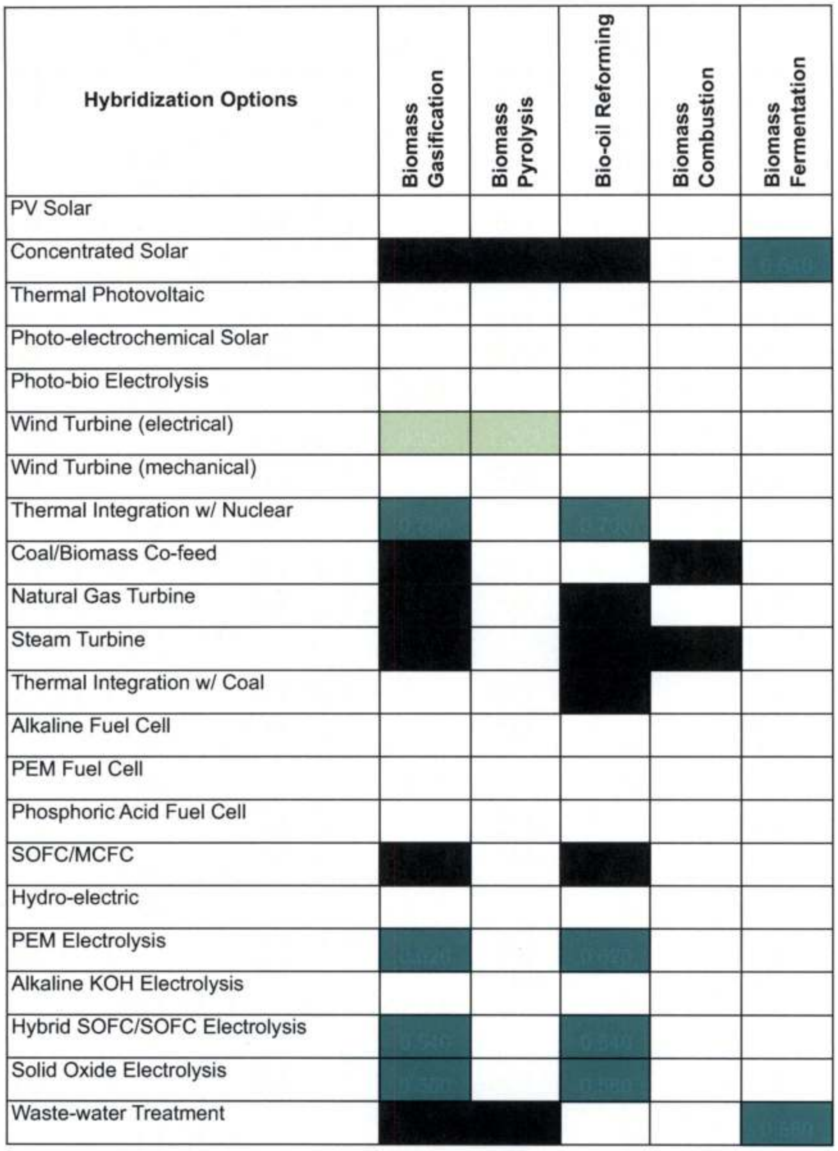
[2] Pyrolysis is the thermochemical decomposition of biomass in the absence of oxygen.



Hybridization Options	**Biomass Gasification**	**Biomass Pyrolysis**	**Bio-oil Reforming**	**Biomass Combustion**	**Biomass Fermentation**
PV Solar					
Concentrated Solar	■	■	■		▨
Thermal Photovoltaic					
Photo-electrochemical Solar					
Photo-bio Electrolysis					
Wind Turbine (electrical)	🟩	🟩			
Wind Turbine (mechanical)					
Thermal Integration w/ Nuclear	▨		▨		
Coal/Biomass Co-feed	■			■	
Natural Gas Turbine	■		■		
Steam Turbine	■		■	■	
Thermal Integration w/ Coal			■		
Alkaline Fuel Cell					
PEM Fuel Cell					
Phosphoric Acid Fuel Cell					
SOFC/MCFC	■		■		
Hydro-electric					
PEM Electrolysis	▨		▨		
Alkaline KOH Electrolysis					
Hybrid SOFC/SOFC Electrolysis	▨		▨		
Solid Oxide Electrolysis	▨		▨		
Waste-water Treatment	■	■			▨

Figure 1.1 - Idea Matrix Results

Initial brainstorming led to further consideration of the major concepts listed below:

- Combined wind power and biomass gasification for co-production of fuel and power.
- Combined electrolysis and biomass gasification for co-production of fuel and power.
- Combined coal and biomass gasification systems for co-production of fuel and power with carbon sequestration for both processes.

- Modified, integrated combined-cycle systems to be fed with syngas from biomass gasification for co-production of fuel and power.
- Combined, direct concentrated-solar and biomass gasification for fuel production.
- Co-location and thermal integration using steam from a nuclear reactor to feed bio-oil reforming to produce fuel.
- Co-location of biomass gasification and wastewater treatment for co-production of fuel and power.

CHAPTER 2
LITERATURE REVIEW

A literature review was undertaken to try to determine the current state of research with respect to biomass hybridization. The literature review showed that many of the previously discussed concepts have been, or are currently being, studied. However, there were several areas where research was either sparse or nonexistent.

Since all of the systems considered included some form of biomass conversion, the first step was to determine the availability of biomass resources. Figure 2.1 shows the areas of the US with greater than 2000 ton per day (TPD) of biomass available[3] within a 50-mile radius in green. Multiple types of biomass were considered including crop residues and woody biomass. Crop residues considered included corn, wheat, soybeans, cotton, sorghum, barley, oats, rice, rye, canola, beans, peas, peanuts, potatoes, safflower, sunflower, sugarcane, and flaxseed. It is important to note that estimates of residue were adjusted down to allow for soil erosion control, animal feed, bedding, and other existing farm uses (Milbrandt 2005). Woody biomass includes forest residues, primary mill residues and secondary mill residues. Primary mill residues are the bark and wood materials produced when logs are processed into lumber. Secondary mill residues consider the wood scraps from woodworking shops and factories such as furniture manufactures.

Particular attention was given to opportunities for hybridization of biomass pyrolysis and gasification plants. Pyrolysis is the thermochemical

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[3] 2000 TPD of bone-dry biomass is equal to approximately 430 MW of available energy based on lower heating value.

decomposition of biomass in the absence of oxygen. It is produces a mixture of synthesis gas and bio-oil at temperatures around 500°C. Gasification uses partial oxidation of the feedstock to provide heat for the reactor and is typically run at temperatures above 800°C. The higher temperature produces synthesis gas with very little bio-oil (which is considered an impurity or tar). Both of these technologies are promising alternatives for the production of 2nd generation bio-fuels from non-food crops.

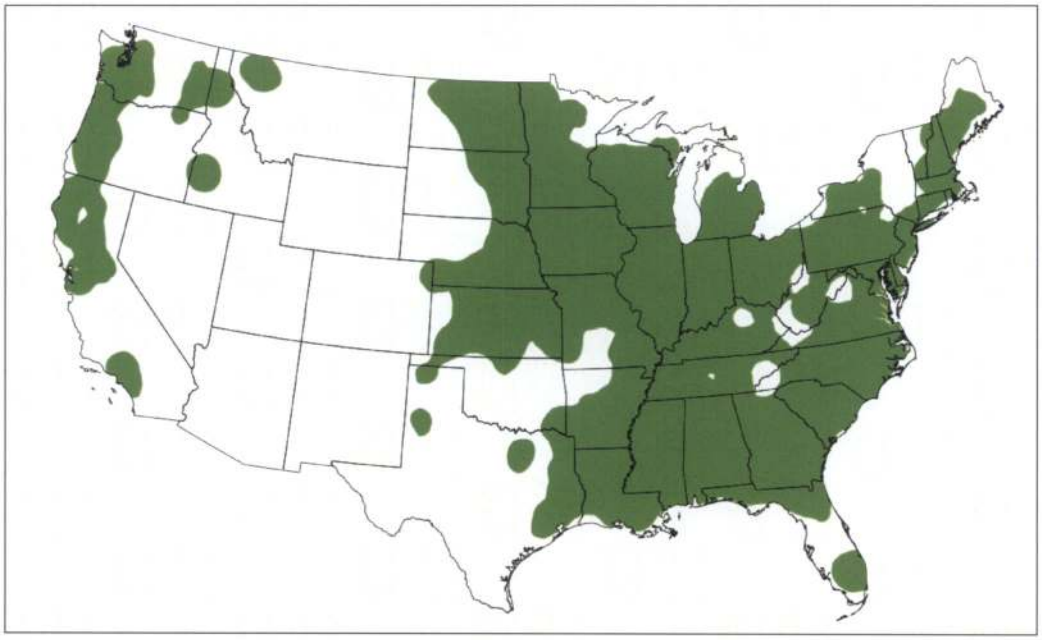

Figure 2.1 - Biomass Resources Availability[4]

2.1 Concentrated Solar – Biomass

Using solar energy to provide the heat for thermochemical biomass processing would reduce the environmental impact of both gasification and pyrolysis facilities compared to combustion heating. There are two main areas of research in this hybridization option, direct thermal transfer and indirect thermal transfer.

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[4] Additional NREL, biomass GIS data is available at www.nrel.gov/gis/biomass.html

In direct-thermal transfer, solar concentrators are focused into the reaction chamber of a pyrolyzer or gasifier. Several bench scale systems have been designed and tested to date with disappointing results (Lede 1999). These systems suffer from several technical problems including the amount of solar concentration needed to reach plausible reaction temperatures, solar intermittency, the need for a clear window into the reaction chamber, scalability concerns and the severe solar diffusion caused by particle movement within the reactor. In addition to technical challenges, capital costs for building a plant are expected to be significant. While technical hurdles remain, valuable research is ongoing. A good summary of the state of technology is provided in (Steinfeld 2005). In addition, research is ongoing at the University of Colorado on this topic (American Institute of Chemical Engineers).

Indirect thermal transfer relies on heating the outer walls of a reaction chamber or heating an intermediate used for thermal storage. These systems can provide a more consistent heat source at the expense of lower absolute temperature. Indirect thermal transfer systems are more likely to be used for biomass pyrolysis than gasification because of temperature limitations. One interesting option is the use of concentrated solar energy to heat molten salts, which can then be used as a pyrolyzing medium. Preliminary research suggests that using this approach a pyrolysis reactor could be run at steady state on solar energy alone (Adinberg et al. 2004).

Whether indirect or direct thermal transfer is used, both technologies require significant solar radiation and concentration. Rough concentration ratios for trough, tower, and dish concentrators are 100, 1000, and 3000 suns respectively (Masters 2004). All three technologies are technically feasible for power generation using a heated fluid in a traditional thermal-cycle. Trough concentration systems typically run at temperatures ranging from 300–400°C (National Renewable Energy Laboratory). The extreme temperatures needed for pyrolysis and gasification (500°C and 860°C respectively) favor tower or dish concentration systems. Tower concentrator systems are the most likely candidate for use with a stationary chemical reactor.

The US has significant solar resources at its disposal but they are generally concentrated in the deserts of the southwest where biomass availability is low. Plant location is further restricted because tower concentration systems require not only high solar radiation but also large areas of flat land for construction. With high transportation costs being one of the major obstacles to biomass as an energy source, the lack of local resources is problematic.

Figure 2.2 shows the areas with biomass resources greater than 2000 TPD within 50 miles in green. The orange color denotes solar resources in the Southwestern US of 6 kWh/m^2/day or greater direct normal radiation. The solar resources are further constrained to flat sites (areas with less than 1% land slope) excluding environmentally protected lands, urban areas, and water features. In short, the orange areas denote sites that might be capable of supporting a solar tower concentrator plant. Figure 2.2 shows that there is very little overlap of concentrated solar and biomass resources. Two exceptions to this can be found in a small part of southern California and pockets of northern Texas.

Biomass gasification and pyrolysis plants typically require external electrical power for operation when the plants are optimized for fuel production. Many of the biomass gasification and pyrolysis research papers reviewed mentioned that the use of renewable sources of power would further add to the environmental benefits of biomass processing.

While many papers mentioned using electricity produced by renewables, no papers were found directly addressing the combination of intermittent photovoltaic power with thermochemical processing.

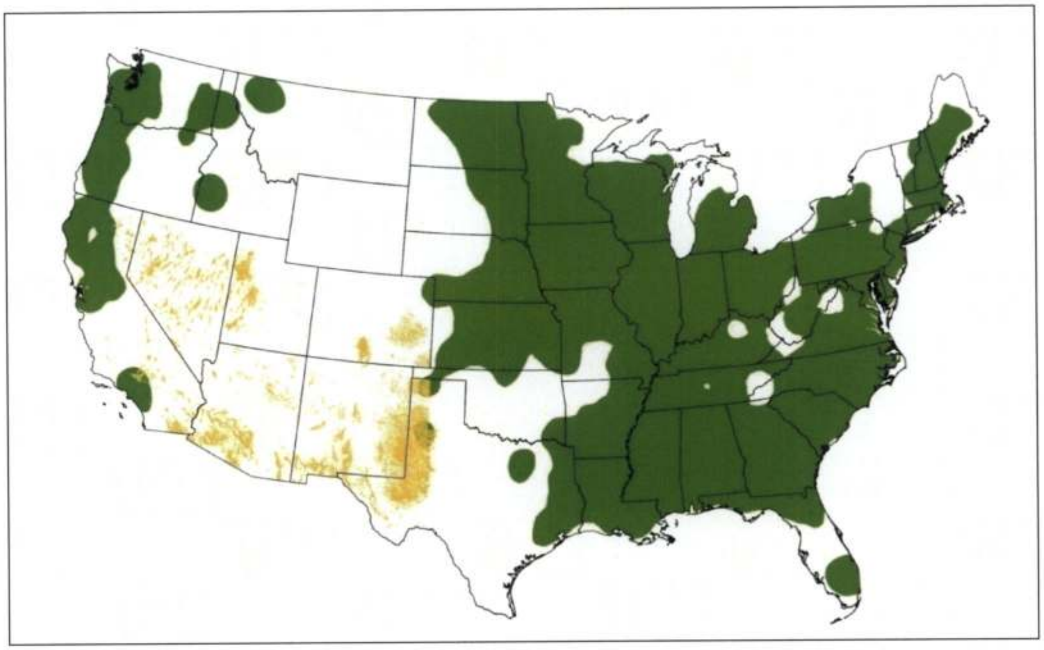

Figure 2.2 - Solar Tower Plant Locations vs. Biomass Resources

Use of photovoltaic (PV) electricity has several advantages over direct use of the radiation. The most significant advantage is the possibility of bringing the electricity to the biomass resources rather than having to ship biomass large distances. In addition, there is significant potential for distributed photovoltaic installation throughout the US (National Renewable Energy Laboratory).

The availability of low cost solar-generated electricity may be a challenge due to many factors. Peak solar radiation generally coincides with peak electricity demand, making the cost of the renewable electricity too high for cost-effective biorefinery usage. In addition, capital costs remain high for solar installations, driving up the baseline cost of solar generated electricity.

Photo-electrochemical water splitting is a future possibility for hydrogen production via direct water splitting. This technology was not considered viable for hybridization because of currently high-cost and low efficiency even at bench scale (Turner et al. 2008). In the future, it could provide a way to produce oxygen and hydrogen for biomass gasification without the high electricity requirements of both cryogenic air separation units and electrolyzers.

2.2 Integrated Gasification Combined Cycle

Biomass gasification for power generation is a more efficient route to power production than direct combustion. Extensive research has been done to date on using biomass and/or the syngas produced from a biomass gasifier to create power using either a gas turbine alone, a steam turbine alone or an integrated combined cycle approach. At least two major studies have been released by NREL directly addressing the technology, economics, and life-cycle implications of this type of hybrid power generation technology (Craig et al. 1996 and Spath et al. 2004). In addition to these assessments, many biomass to liquid fuel studies assume that unconverted syngas is burned in a gas turbine for power generation (Larson et al. 2005).

From a greenhouse gas emissions standpoint, biomass based power plants produce significantly fewer emissions than coal or natural gas systems. In fact, from a lifecycle perspective, a biomass IGCC produces approximately 94% fewer greenhouse gas emissions per kWh than a conventional coal plant and 90% fewer emissions than a natural gas ICC system (Spath et al. 2004). Even when carbon sequestration is used on fossil fuel plants, a biomass IGCC plant will produce fewer atmospheric greenhouse emissions (Spath et al. 2004). The major challenge for biomass based IGCC plants is the economies of scale limitations due to biomass availability. Biomass cannot be shipped for long distances economically because of its low energy density. Biomass IGCC plants are typically in the 10 – 60 MW range as compared to 500 MW coal gasification plants (Electricity Power Research Institute 2006).

2.3 Fuel Cell – Biomass

Coupling a fuel cell directly with the syngas output of a biomass gasification plant is a highly efficient way to produce electricity from biomass. With no moving parts and freedom from the Carnot limit, fuel cells can achieve much higher efficiencies than conventional turbines. In addition to high efficiencies, fuel cells run on a variety of fuels and typically have low maintenance requirements.

Molten carbonate (MCFC) and solid oxide (SOFC) fuel cells are the most likely candidates for combination with gasification because of their relatively low fuel-quality demands, high operating temperatures and tolerance of carbon monoxide (Seitarides et al. 2008). Owing to the high operating temperatures of these fuel cells (600 – 1000°C), it is typically most economical to produce combined heat and power with them. Several studies have been carried out on the possibility of combining gasification with high temperature fuel cells. Total biomass gasification to MCFC plant electrical efficiencies of approximately 40% have been reported in literature (Wang et al. 2008).

The major obstacles to both biomass SOFC and MCFC systems are cost, syngas cleaning, and durability. Both SOFC and MCFC systems are extremely sensitive to sulfur and some of the corrosive tars produced by gasification. The cost of syngas increases with increasing purity requirements, thus cleaning the syngas for fuel cell use could be a significant burden compared to combustion requirements. However, the syngas cleaning requirements for fuel cell systems are similar to those required for fuel synthesis from syngas. High temperature fuel cells are commercially available, but cost is still a major barrier to large-scale deployment especially when combined with the high costs of gasification equipment.

2.4 Electrolysis – Biomass

Directly heated gasification systems require a source of pure oxygen if they are to be used for fuel production. Currently, plants that use oxygen produce it with cryogenic air separation units (ASUs) (Ciferno et al. 2002). Electrolysis could provide an alternative to ASUs with the added benefit of producing a pure hydrogen stream. Initial research has shown that this hybridization option could be promising from both a technical and economic perspective (Gassner 2008). However, the feasibility of producing oxygen and hydrogen with electrolysis is heavily dependant on both the price of electricity and the value of the end products to the plant.

One other biomass gasification and electrolysis hybridization study was found in publication (Hulteberg et al. 2009). Based on a conversion of the papers results (in Swedish Krona) to dollars, the prices for hydrogen by gasification, electrolysis and hybridization were \$5.55, \$6.15 and \$5.85 per kilogram of hydrogen produced, respectively. The study was performed on a much smaller scale plant (approximately 37,000 kg/day H_2) than is being considered in this study.

At standard temperature and pressure, an ideal electrolyzer would use 39 kWh of electricity to produce one kilogram of hydrogen. The actual state of technology limits system efficiencies to between 56 – 73% meaning that approximately 53 – 70 kWh of electricity is needed for every kilogram of hydrogen produced (Kroposki et al. 2006).

To replace a single ASU for oxygen production, multiple electrolyzers would be needed. The largest commercial electrolyzer is produced by StatoilHydro (formerly NorskHydro) and produces a maximum flow rate of 43.6 kg/hr of hydrogen (174.4 kg/hr of oxygen) (StatoilHydro). A 2000 TPD biomass gasifier would require a large bank of these electrolyzers running at full capacity (the exact number of electrolyzers required is calculated in Section 6.1.2). According to NREL's most recent H2A forecourt[5] electrolysis analysis, one 174.4 kg/hr electrolyzer would cost approximately 2.5 million dollars installed with hardware. Therefore the electrolyzer bank for a 2000 TPD gasification plant would cost significantly more than a comparable ASU. One possibility for addressing these high capital costs is to use enriched air for gasification rather than pure oxygen.

Besides the high capital cost of electrolysis, there are other concerns with this hybridization. Water usage is a key concern with electrolyzer systems and would be especially pronounced when combined with the generally high water requirements of biomass processes and steam power production. Also, pressurized gasification plants use nitrogen from the ASU for pressurizing the biomass feed system. Another source of inert pressurization would be needed.

2.5 Wastewater Treatment – Biomass

Every year approximately 5.6 million dry tons of solid waste (or sludge) is produced in the United States (Bagchi et al. 2006). A significant amount of this sludge is either land filled or incinerated. Gasification could provide an alternative use for this readily available source of biomass.

The high water content of sludge is a significant challenge to traditional biomass gasification systems. Two options exist to overcome the water

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[5] Forecourt refers to a refueling station.

challenge. Municipal waste can be preprocessed and dried to levels acceptable to the reaction chamber before gasification or the gasifier can be run at much lower efficiencies and the water can be vaporized in the reaction chamber itself. Using a directly heated gasifier and wet biomass would result in a significant percentage of the input carbon being burned to heat water rather than produce syngas.

One promising alternative to traditional gasification is plasma gasification. Plasma gasification is typically done with electrically heated arc furnaces running at temperatures well above 950 degrees Celsius. It has been successfully used to produce high quality syngas and power from sewage sludge in the U.S., Canada, Malaysia and Japan. The most well known of these plants is located in Japan, and produces approximately 4 MW of grid electricity by processing 138 tons per day of sewage sludge. A good overview of the state of the technology and references to existing plants can be found in (Mountouris et. al. 2006).

Though less glamorous, biogas digesters are another option for turning sludge into useful gas. This 100 year-old technology has and is being used by households in China, India and other countries to produce natural gas for combustion in lanterns and stoves. Because the technology is simple and implementation is low-cost, this approach is being adopted by many wastewater treatment facilities here in the U.S.

2.6 Coal – Biomass

Due to the large existing coal infrastructure in the US, early combinations of coal and biomass involve co-firing or co-gasification of biomass with coal. Biomass can be co-fired in existing coal combustors but only in marginally small percentages due to feed problems (Wang et al. 2008). One way to address feed problems is to gasify biomass and then co-feed pulverized coal and biomass-produced syngas into existing combustors (Electricity Power Research Institute 2006). Another option is to torrify (or thermally pre-treat) the biomass, which produces a char that can be co-feed with coal slurry. A good overview of the practical issues of dual feed systems is provided in "Biomass cofiring: economics, policy and opportunities" (Hughes 2000). In addition to Hughes' paper, the white paper "Biomass Cofiring: A Renewable Alternative for Utilities" provides information about existing plants operating on both fuels (DOE 2000).

Co-gasification of coal and biomass has been a focus of recent research with several good papers published in the last few years (McLendon et al. 2004 and Valero et al. 2006). These dual feed systems help to lower the greenhouse gas emissions of the existing coal infrastructure while maintaining economies of scale and avoiding the difficulties of finding large, reliable quantities of biomass for power generation. When significant amounts of biomass are co-fed, problems can result from increased fouling of downstream processes and high alkali content in the product ash. In the Netherlands, the Buggenum coal gasification plant has reported co-feed percentages of up to 30% with only minor changes in plant power and waste output (Electric Power Research Institute 2006). One recent paper proposed coal gasification with biomass co-feed for production of fuel and power (Cormos 2009).

Another synthesis possibility is thermal integration of biomass gasification or bio-oil reforming facilities with existing coal fired power plants. Biomass gasification, whether directly or indirectly heated, requires a steam source that could come directly from a coal power plant. The major challenges to this type of integration are the added capital cost, the low steam temperatures relative to gasification requirements and the mismatch in scale between biomass availability and steam production. Biomass availability could be addressed by gasification of bio-oil produced from multiple off-site pyrolysis units. Papers directly addressing the technological and economic possibilities of this system were not found during the literature review.

2.7 Nuclear – Biomass

Thermal integration and co-location of biomass processing with nuclear energy is a promising hybridization option. The presence of near carbon neutral power and steam from the reactor could significantly increase the efficiency of a biomass plant. Charles Forsberg makes a strong case for this concept (Forsberg 2007).

Nuclear energy currently provides 20% of the electricity in the United States (Nuclear Energy Institute). According to the Nuclear Energy Institute, there are 104 nuclear reactors in the United States with another 30 plants currently seeking federal license approval. All of these existing plants provide a reliable source of electricity and could provide low pressure, low temperature steam to a biomass processing facility. Currently this steam is a waste stream that must be condensed after the last turbine cycle for US plants. Most gasification plants would need to upgrade the steam quality before it entered the reactor, however ethanol plants could use the low quality steam directly (Forsberg 2007).

For economic reasons, the nuclear plant would need to be near either large biomass resources or access to low cost barge transportation in order to benefit from hybridization. Figure 2.3 shows an overlay of existing nuclear facilities (marked with black dots) versus biomass availability. It shows that at least 6 of the existing nuclear plants are located in areas with greater than 1,500 TPD of biomass available. Approximately 23 of the existing reactor sites are situated in areas that could support biomass plants of 750 TPD or larger.

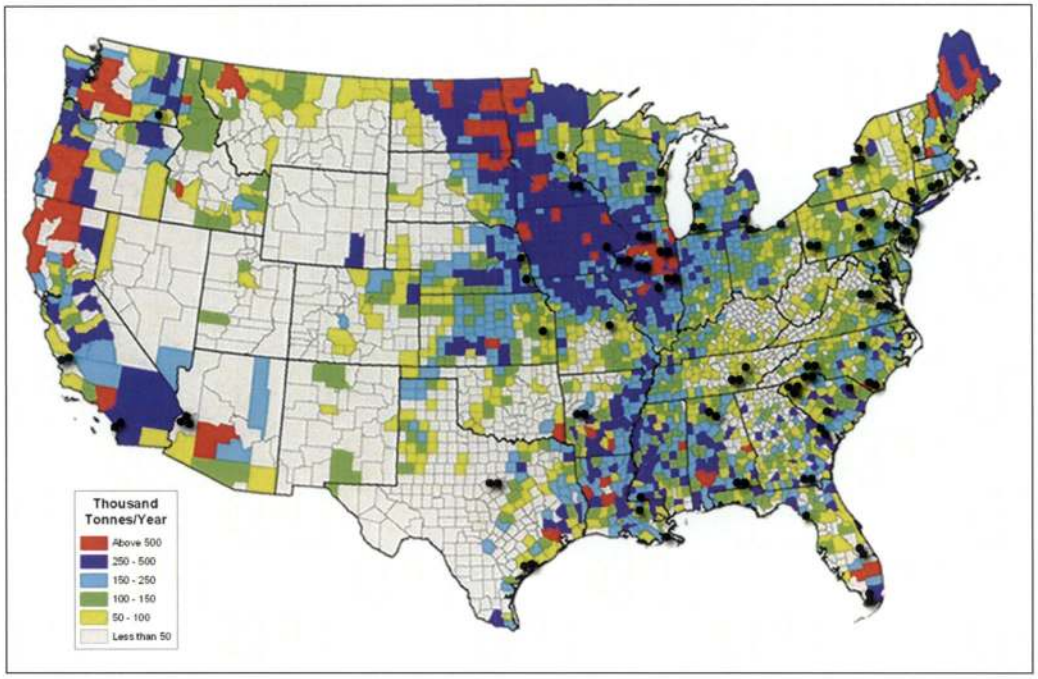

Figure 2.3 - Nuclear Plant Locations vs. Biomass Resources

Even with significant biomass resource availability, there will be a scale mismatch between biomass availability and steam production. As mentioned in the coal-biomass section of the paper, scale issues could be addressed by gasifying bio-oil produced from multiple off-site pyrolysis facilities (the "hub and spoke" concept). One way to take advantage of the scale mismatch might be to pull a slipstream of super-critical steam from the nuclear power plant steam cycle for use in gasification. No documentation or research was found on this concept. Significant nuclear plant modifications would need to be made for this type of integration.

Nuclear power plants rely on non-renewable uranium resources to create heat and subsequently power. For any biomass-nuclear hybridization to be reasonable, sufficient domestic uranium resources must be available in the long term. According to Nuclear Energy Association estimates, fuel availability is not a concern for several centuries (Price 2002). Reserves could last significantly longer with improvements in mining technology, reactor design and increased fuel rod recycling.

Major challenges to such a system include US resistance to new nuclear plants and security concerns involved with additional on-site processing. It may be possible to address both issues with biomass. Placing the biomass plant outside of the secure perimeter and piping the steam over the fence could negate security concerns. Creating an additional source of farm income in rural areas may go a long way towards overcoming local resistance to nuclear reactors.

2.8 Wind – Biomass

Biomass gasification and pyrolysis plants typically require external power for operation when the plants are optimized for fuel production. Many of the biomass gasification and pyrolysis papers reviewed mentioned that the use of renewable sources of power would further add to the environmental benefits of thermochemical biomass processing. While many papers mentioned using electricity produced by renewables, few looked at how to directly couple intermittent wind power with thermochemical processing.

Wind turbines have quickly become a widely accepted, commercial source of renewable energy in the US. Over the last 29 years US utilities have vastly improved their knowledge and ability to manage intermittent electricity sources. However, there are significant issues that remain if large-scale wind power is pursued in the US. These issues are addressed in detail in (20% Wind Energy by 2030, 2008).

The two most significant issues with wind power are its location and its intermittency. The vast majority of land-based wind resources are found in the rural areas of the middle US (Department of Energy). In order to successfully utilize these resources electricity must be transported long distances to demand centers. Additionally, the intermittency of wind means installing too much capacity will create grid instability unless suitable grid leveling options are available.

Transportation of wind-generated power can be accomplished via the electrical grid or by converting the electricity to a transportable fuel. Using the national electric grid to transport the power would require significant updates to the national infrastructure. Additional high voltage transmission lines would be needed in many locations to connect wind resources with urban areas (20% Wind by 2030, 2008). Another option is to convert intermittent electricity into a fuel. Several studies have been done recently on using electrolyzers to create hydrogen from wind-generated electricity (Levene et al. 2006).

Intermittency of wind electricity can cause challenges for the power grid if proper leveling options are not available. One option, commonly used today, is to use natural gas turbines to maintain system reliability. Gas turbines are readily available, and can be brought on and offline very rapidly. Another option is to use batteries and/or electrolyzers to store power during peak winds for use during low or no-wind conditions (Fingersh 2004).

One of the only papers found to address the challenges of wind directly with biomass processing is (Denholm 2006). The proposed system would use compressed air energy storage to store off-peak electricity generated by wind. This energy would then be used as needed by a properly designed biomass gasification plant.

Portions of national wind resources lie in areas that also have biomass availability. Because of this, the shift from viewing wind electricity as an external source of electrons to trying to find direct synthesis between the two technologies appears to be a promising area of research. As a first step in this direction, maps were constructed that overlaid class 4 or better wind resources on biomass resources. For both maps, the wind resources are shown in red and exclude potentially sensitive environmental lands, wind on water features and stranded wind resources (i.e. small isolated areas). The green shading on both maps indicate that greater than 2000 TPD of the specified type of biomass is available within 50-miles. There is some “wash out” of wind data because the spatial resolution on the wind resources is much finer than the biomass, so the flecks of red can be difficult to observe.

Figure 2.4 shows woody biomass resources versus available wind. Woody biomass includes forest residues, primary mill residues and secondary mill residues. Primary mill residues are the bark and wood materials produced when logs are processed into lumber. Secondary mill residues consider the wood scraps from woodworking shops and factories such as furniture manufacturers. As can be seen on the maps, there are small pockets of the northwest and northeast where both class 4 or greater wind and sufficient woody biomass exist for a co-located, combined system.

Figure 2.5 shows agricultural (crop) residue biomass resources versus available wind. Crop residues considered included corn, wheat, soybeans, cotton, sorghum, barley, oats, rice, rye, canola, beans, peas, peanuts, potatoes, safflower, sunflower, sugarcane, and flaxseed. It is important to note that estimates of residue were adjusted down to allow for soil erosion control, animal feed, bedding and other existing farm uses (Milbrandt 2005). There is significantly more overlap of agricultural biomass with wind than woody biomass and wind.

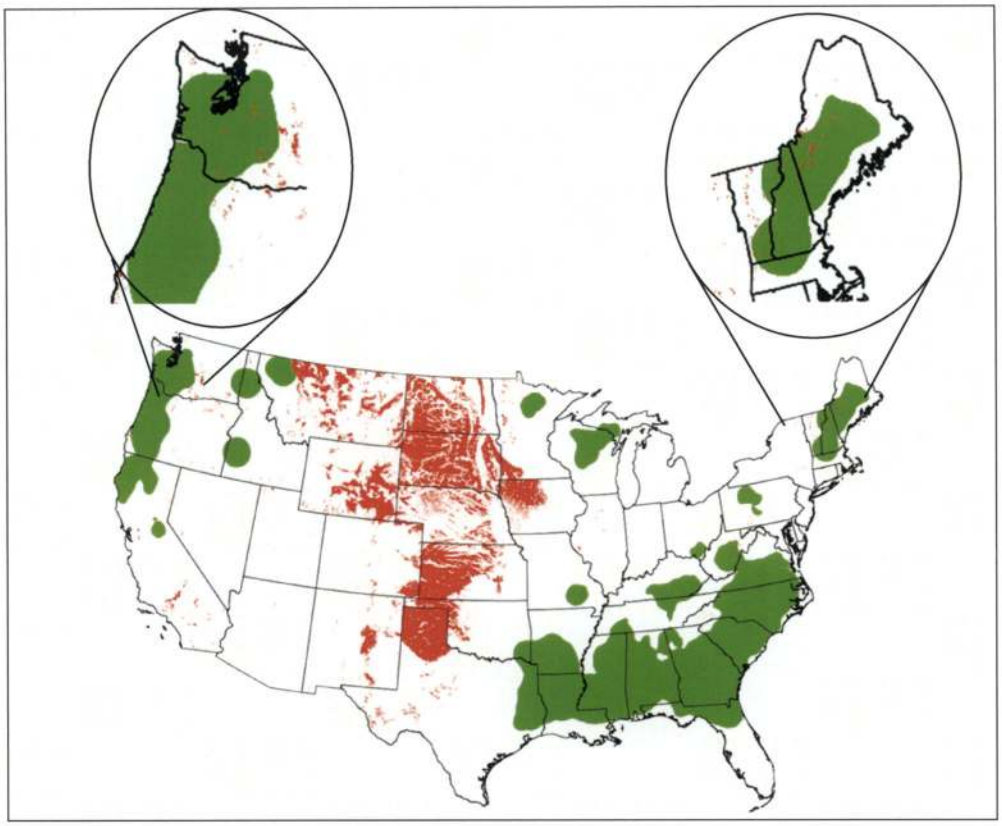

Figure 2.4 - Wind Resources vs. Woody Biomass Resources

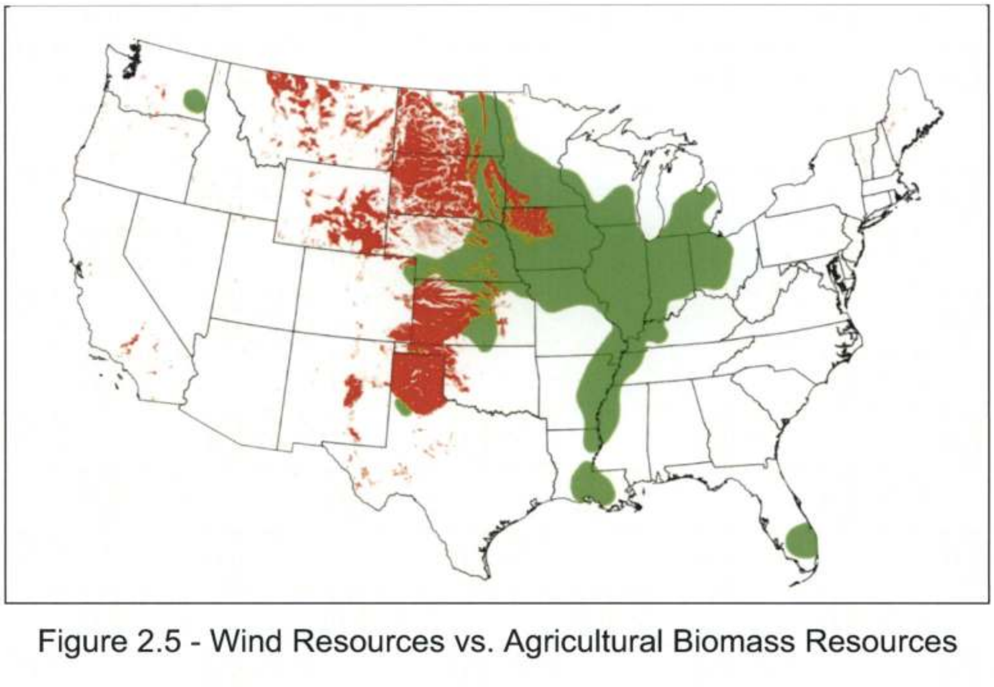

Figure 2.5 - Wind Resources vs. Agricultural Biomass Resources

2.9 Industrial Hybridization

It may be advantageous to look to industries that currently have high heat processes, biomass waste streams or large steam requirements for synthesis possibilities. One example is replacing industrial gas in limekilns with synthesis gas from the gasification of hog fuel (Gribik et al. 2007). This research is outside of the scope of this thesis research but may be a promising future area of research.

CHAPTER 3
CONCEPT SELECTION

Using the literature review process to refine and augment initial brainstorming, a list of promising hybridization options was compiled. This initial list of options did not focus specifically on hydrogen production. Instead, concepts that produced hydrogen, electricity, or a combination of the two were all considered. The options are listed below:

1. Combined wind power and indirectly-heated, biomass gasification plant for co-production of fuel and power. Electric heaters could be used to increase the efficiency of hydrogen production in the biomass plant. A syngas-fed gas turbine could provide peaking electricity.
2. Use of an electrolyzer in place of an air separation unit (ASU) for a directly heated biomass gasifier for co-production of fuels and power. In addition to providing oxygen for the gasification reaction, the hydrogen produced can be used to increase fuel production of the biomass to hydrogen plant.
3. The system described in concept 2 above could be combined with wind so that additional oxygen and hydrogen can be produced during off-peak hours for use during times of high electricity demand.
4. Combined biomass gasification and solid oxide or molten carbonate fuel cells. The syngas produced by gasification will produce power via the fuel cell. This system should provide higher overall electrical efficiency than a similar combined cycle system.
5. Thermally-integrated coal power plant and biomass/bio-oil gasification system. Possible opportunities for research include examining if gasification of bio-oil could be used to improve the

kinetics of coal gasification, if waste steam could be used for gasification, and the life cycle implications of the combination. Carbon sequestration could be used to benefit both processes.

6. Combined concentrated solar and molten-salt biomass pyrolysis so that the heat needed for biomass reforming is generated from a renewable, carbon neutral source. Syngas generated from the system would be used for hydrogen production.
7. Direct, concentrated solar energy for biomass gasification. Synas produced from the gasifier could be used for fuel or power production.
8. Co-located and thermally-integrated bio-oil reforming using the supercritical water from a nuclear reactor to create hydrogen fuel (with a distributed pyrolysis system).
9. Co-location of biomass gasification and wastewater treatment for co-production of hydrogen and power.

To narrow the hybrid systems mentioned above down to two for further investigation a Pugh decision matrix was constructed.[6] A Pugh decision matrix is used to compare multiple alternatives to a chosen baseline case. For this project all systems were compared to a biomass gasification integrated combined-cycle system. Each system was then ranked on criteria including greenhouse gas emissions, feedstock renewability and availability, reliability, cost, fuel production (syngas output per unit biomass) and the state of the technology. Rankings were input based on better (+1), worse (-1) or neutral (0) compared to the biomass IGCC plant. It is common practice to apply weighting factors to the criteria when needed. Because of the variety of systems involved, and in order to select the most generally promising systems, all weighting factors (w) were set equal (value of 1) for this study. In addition to the concept titles listed

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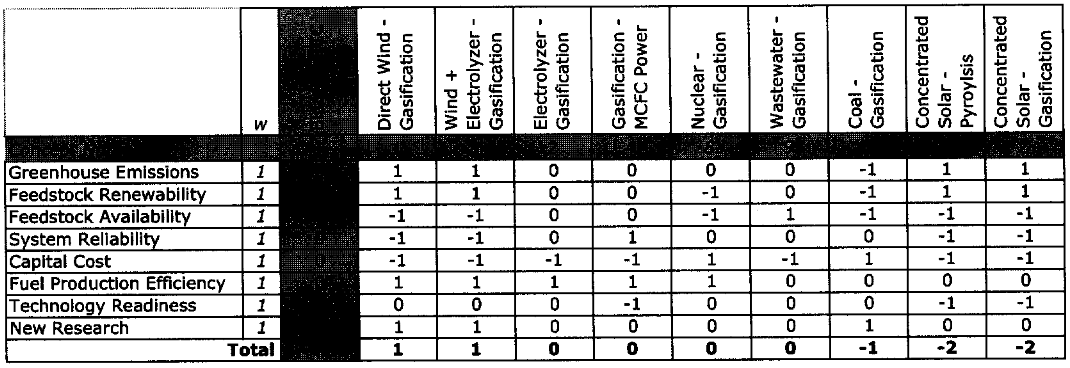
[6] For more information on this technique see: www.sixsigmafirst.com/PughMatrix.html

along the top of the table, the concept number (referencing the list above) is shown in italic.



	W		Direct Wind - Gasification	Wind + Electrolyzer - Gasification	Electrolyzer - Gasification	Gasification - MCFC Power	Nuclear - Gasification	Wastewater - Gasification	Coal - Gasification	Concentrated Solar - Pyroylsis	Concentrated Solar - Gasification
Greenhouse Emissions	*1*		1	1	0	0	0	0	-1	1	1
Feedstock Renewability	*1*		1	1	0	0	-1	0	-1	1	1
Feedstock Availability	*1*		-1	-1	0	0	-1	1	-1	-1	-1
System Reliability	*1*		-1	-1	0	1	0	0	0	-1	-1
Capital Cost	*1*		-1	-1	-1	-1	1	-1	1	-1	-1
Fuel Production Efficiency	*1*		1	1	1	1	1	0	0	0	0
Technology Readiness	*1*		0	0	0	-1	0	0	0	-1	-1
New Research	*1*		1	1	0	0	0	0	1	0	0
Total			**1**	**1**	**0**	**0**	**0**	**0**	**-1**	**-2**	**-2**

Figure 3.1 - Pugh Decision Matrix

It can be seen from the matrix that the concentrated solar systems scored low due to their high cost, intermittency and concerns with the state of the technologies involved. Coal based systems were low on the decision matrix because of the greenhouse gas emissions involved and concerns with the scaling disparity between available local biomass resources and coal plant outputs. Electrolyzer – gasification and wastewater (or sludge) gasification, and MCFC power systems were all seen as alternatives to biomass combined cycle technology. The possibility of using nuclear steam for gasification is also an alternative to the baseline, but concerns remain about the nonrenewable fuel for nuclear power and scaling disparity between biomass resources and plant outputs.

Direct wind and wind/electrolyzer combinations with biomass gasification rose to the top of the decision matrix due to several factors. Little research has been done in these areas to assess economics and technical feasibility. In addition to novelty, the use of renewable wind resources to power a renewable biomass process lays the foundation for truly renewable fuel production. Finally, both cases have the possibility to increase syngas and/or fuel yields from a given amount of biomass

The two concepts chosen for further analysis can be stated as:

1. Direct grid leveling of intermittent wind power with an indirectly-heated biomass gasification plant. The plant will produce both electricity and fuel.
2. Using an electrolyzer in place of an air separation unit (ASU) for a directly heated biomass gasifier for co-production of fuel and power.

Both of the concepts chosen for further analysis share the basic idea of combining wind-generated electricity with a biomass gasification plant. Wind availability overlaps biomass resource availability (Figures 2.4 and 2.5), making the use of locally produced wind electricity for gasification feasible. In addition, gasification plants provide multiple opportunities for electricity use.

While wind power is a promising and largely commercial renewable source of energy, its penetration of the grid will cause some unique challenges. These challenges include management of intermittency with peaking units and, in the extreme case, finding use for electricity produced by wind when there is no demand. Managing intermittency will drive utilities to invest in additional peaking units and increase the need for interruptible customers, energy storage options, and dispatchable loads. Finally, wind in many parts of the country is located far from demand centers and is therefore a stranded resource. Direct synthesis between the two technologies could allow a hybrid system to manage local intermittency or capture stranded resources.

Figure 3.2 shows an hourly wind generation profile from Norfolk Nebraska versus the electricity demand for the same hours as reported by the Midwest ISO. Three lines are plotted showing the wind profile at 1%, 20%, and 100% wind penetration to the market. At current levels (1%) wind energy has little if any effect on the broader energy market. At 20% wind power, significant peaking units will be needed to manage the unpredictable wind generation. Dispatchable demand may also become a sought after service to the grid at night and any other times when high wind is not in phase with demand. While 100% wind power is not a possibility on a regional scale, the curve is shown to emphasize the effect that wind power can have in localized pockets of the rural grid where there may be significant wind generation with low demand.

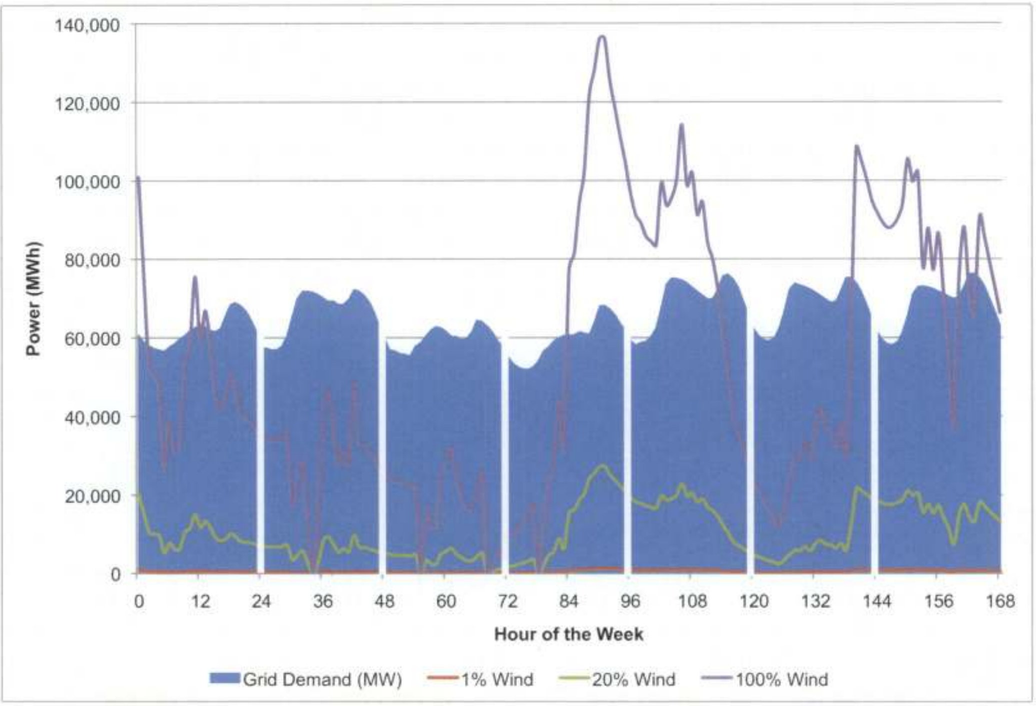


Figure 3.2 - Wind Generation vs. Demand for a One-week Period

The proposed hybrid systems attempt to do one of two things:

1. Fill wind energy shortfalls and replace the natural gas turbines that would generally be used for this peaking purpose with a renewable alternative.
2. Absorb wind generated power that there is no demand for. Biomass gasification aided with wind electricity could “sink” this energy into fuel and provide dispatchable demand for the local utility.

CHAPTER 4
ANALYSIS OBJECTIVES

The two hybrid systems chosen for detailed analysis utilize different biomass gasification architectures to either absorb or provide electricity to the grid and produce hydrogen fuel. Hydrogen fuel was chosen because the National Renewable Energy Laboratory Hydrogen Technologies and Systems Center funded this research. The goal of the analysis is to determine the technical and economic feasibility of the proposed hybrid systems.

A plant capacity of 2000 ton per day (TPD) of biomass was selected for the analysis based on resource availability. Assuming a fifty-mile collection radius, there are multiple locations throughout the country that could support this level of biomass requirement. At scales larger than 2000 TPD the number of possible plant locations in the US becomes severely limited (Milbrandt 2005).

Based on a fixed biomass throughput, baseline directly-heated and indirect-heated gasification plants were determined. The baseline systems were optimized for hydrogen production using near term (~2012 timeframe) technologies. For the indirectly-heated gasifier, a published ASPEN Plus system study was used as the baseline (Spath et al. 2005). Since there was not an existing ASPEN Plus model available for the directly-heated gasifier case, a model was constructed as part of this thesis. With indirectly and directly-heated baseline cases in place, each baseline was modified to incorporate the proposed hybrid concepts.

The indirectly-heated gasifier hybrid system consists of two parts: (1) producing peaking electricity intermittently with a gas turbine and (2) sinking electricity into electric heaters intermittently to boost fuel production efficiency. Each of these parts was analyzed separately as independent options. The indirectly-heated gasifier hybrid system is meant to be grid connected and provide peaking electricity and dispachable demand (e.g. an on demand load) to the local utility the help manage intermittent wind resources.

The directly-heated gasifier hybrid system involves replacement of the air separation unit in the baseline design with electrolyzers. This change allows for extra production of hydrogen and intermittent operation. The analysis for this system assumed direct replacement of the ASU without intermittent operation. From there estimates were made of the effect of intermittent operation on plant economics. The electrolyzer bank could be either grid-connected to provide a dispachable demand to the utility or it could be tied directly to a stranded wind source to convert wind into fuel.

The objectives for both systems analyzed were to determine:

- If the proposed hybridization is technically feasible given existing or near-term technology.
- To quantify the energy production or energy absorption possible with a 2000 TPD biorefinery.
- The capital cost implications of each hybridization option.
- To compare the cost of hydrogen production with the proposed hybrid systems to that of the baseline systems and other competing systems such as steam methane reforming (SMR) and electrolysis.

The following discussion addresses each of these objectives in turn for each of the proposed hybrid systems. Control strategies and the dynamic effects of intermittent operation were not considered. Instead, the results of this thesis are intended to help inform the decision of whether the proposed systems deserve further consideration.

CHAPTER 5

INDIRECT GASIFIER HYBRID SYSTEM

This concept investigates two possible changes to a biomass to hydrogen plant based on indirect gasification architecture. The first modification is to allow switching between fuel production and electricity production based on grid demand. This is accomplished by routing some or all of the synthesis gas from the gasifier to a gas turbine instead of the fuel production reactors. In addition to power production, modifications that allow use of additional cheap or surplus electricity by the gasifier are investigated. A block diagram of the proposed system is shown in Figure 5.1.

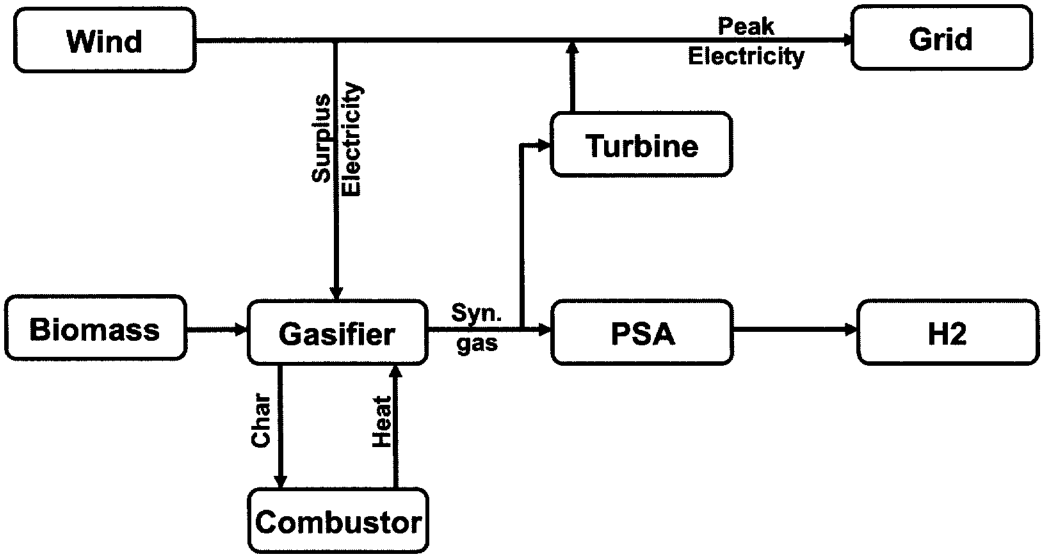


Figure 5.1 – Indirect Hybrid Block Diagram

Indirectly-heated gasification is a two-stage fluidized-bed process where the heat needed for reaction is produced by burning char in a separate chamber to heat sand. This hot sand is then circulated through the reaction chamber to drive reaction kinetics. The layout can be seen in Figure 5.2.

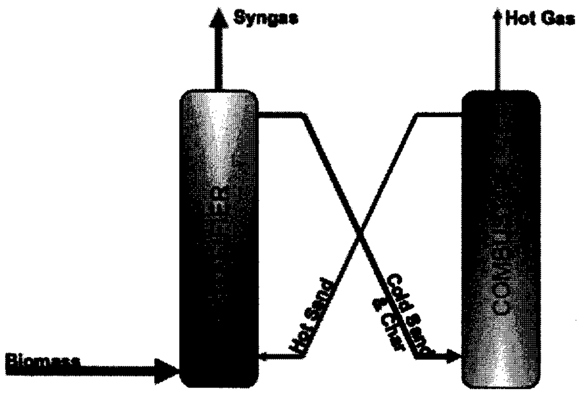


Figure 5.2 - Indirect Gasifier Diagram

Electricity supplied to the gasifier during periods of low demand (low purchase price) will be used to heat the gasifier reaction chamber. As the temperature of the gasifier is increased, the proportions of syngas, char, and tar produced by the gasifier from a given amount of biomass change via a known relationship for a given system. Adding heat energy will create additional syngas, which will increase plant efficiency.

The ideal plant would continuously adjust both feed use and fuel production to optimize the plant economics. Electricity would be produced instead of hydrogen only when electricity was the more profitable product and vice-versa. Similarly, electricity would be used for heating (or sunk) only when electricity costs were low enough that the additional efficiency provided by the electric heat offsets the cost of that electricity. The feed and product selection decision is summarized in Figure 5.3.

Analysis of this concept was separated into the peaking and sinking modifications. The two modifications were analyzed individually to highlight their respective effect on plant economics.

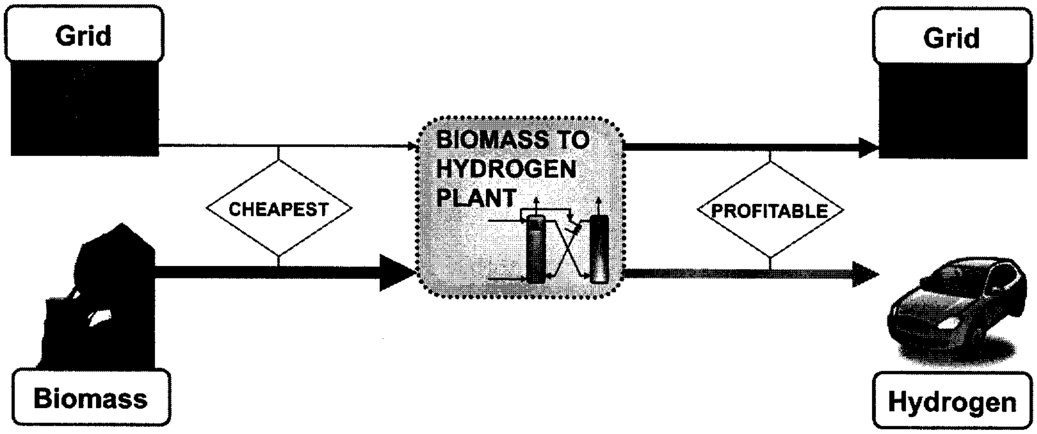


Figure 5.3 - Concept 1 Feed and Product Selection

5.1 Peaking Modifications

Previous NREL studies have looked at the possibility of using a 2000 TPD (ton per day) woody biomass plant for dedicated power production and for dedicated hydrogen production (Craig et al. 1996 and Spath et al. 2005). These studies have assumed steady-state operation of the biomass plant. The current analysis differs from previous studies by alternating between hydrogen production and electricity production based on market demand. Figure 5.4 shows the proposed combination of the two previous studies for the current analysis. The biomass to hydrogen plant design proposed by (Spath et al. 2005) was used as the baseline design. Modifications were then made to this design to allow intermittent power production. The modifications investigated in this thesis are highlighted in red in Figure 5.4.

Biomass entering the plant is dried, then gasified to produce syngas in an atmospheric pressure, indirectly heated gasifier. Tar is thermally cracked in a catalytic tar cracker. The tar cracker must be heated because of the relatively low temperature (870°C) of the syngas leaving the gasifier. Any tar remaining after the tar cracker and any particulate is removed from the syngas with a wet scrub. Sulfur is removed from the syngas in a two-step process of LO-CAT desulfurization followed by ZnO polishing beds. After sulfur has been removed, the carbon monoxide in the syngas stream is shifted towards hydrogen. Finally, the hydrogen is stripped from the syngas and the remaining (purge) gas is burned to provide heat to the tar cracker.

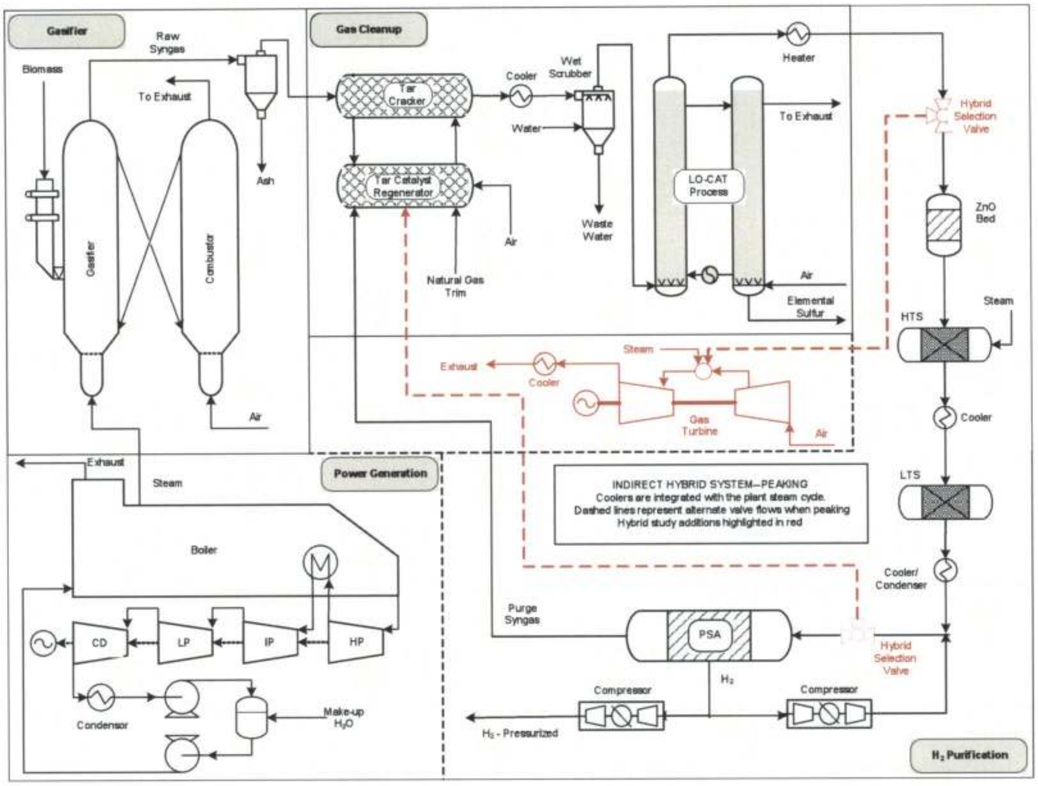


Figure 5.4 – Proposed Peaking Hybrid Process

The proposed plant modifications could increase the economics of biomass utilization. The syngas produced by the gasifier can be used to produce hydrogen fuel or it can be used in a gas turbine to provide peaking electricity, depending on which option will maximize profit. The switching is accomplished by adding valves to reroute the majority of the syngas stream to a gas turbine on demand. Switching, rather than a two train system, was used to provide truly intermittent operation. A two-train system would simply produce both electricity and hydrogen at some constant proportion.

The synthesis gas composition available for use in all calculations is based on the biomass to hydrogen indirectly-heated gasifier study previously completed by NREL (Spath et al. 2005). The "goal design", flows and specifications were used based on the suggestion of the NREL biomass group and are referred to as the "base plant" or "base design".

5.1.1 Plant Design

Multiple locations were considered for redirecting the syngas flow over to the turbine during power generation. Tar reforming is required because tars are extremely corrosive and could adversely affect the turbine combustor. In addition, the tar represents a significant portion of the potential syngas energy, so cracking the tars converts that energy into a compatible form. Therefore, directly after the tar reformer was the first possible location to split the syngas stream for combustion. Other possible locations included:

1. Directly after the compression chain before the LO-CAT System
2. Between the LO-CAT and ZnO sulfur removal steps
3. Before the Pre-PSA Knock-out drum
4. Directly before the PSA unit

Table 5.1 shows the syngas flow, properties, composition and approximate heating values for each option. As syngas is taken further down the reactor chain, the heating value of the gas available decreases. A trade-off must be made between the emissions produced from burning tar and sulfur containing syngas and lower power production if pure syngas is burned. For this analysis, syngas from the LO-CAT reactor is sent to the gas turbine. By placing the turbine after one or both the sulfur removal steps, emissions of SO_x can be significantly reduced.

The options of burning syngas directly prior to the PSA process or burning the hydrogen product directly in a gas turbine were also considered but each would result in less power production due to the lower heating value of the fuel stream. In addition, using pure hydrogen as fuel in a gas turbine would require significant steam or nitrogen co-feeding and cooling modifications that would involve additional study to estimate (Chiesa et al. 2005).

Table 5.1 - Turbine Feed Stream Possibilities

Option Number	*1*	*2*	*3*	*4*	*CNG*
Stream Number	*317*	*326*	*417*	*420*	*427*
Flow (kg/sec)	23.38	23.27	40.68	31.00	-
Pressure (bar)	29.3	28.3	25.5	24.8	1.0
Temp (°C)	60	48.9	43.3	43.3	15.5
		- Mass Fraction (%) -			
H_2	6.61	6.64	5.92	7.77	-
H_2O	0.72	0.31	24.05	0.35	-
CO	54.21	54.48	1.73	2.27	-
N_2	0.14	0.14	0.08	0.11	1.8
CO_2	36.04	36.22	66.96	87.87	1.3
H_2S	0.09	-	-	-	-
NH_3	0.02	0.02	0.01	-	-
CH_4	1.71	1.72	0.98	1.29	91.4
C_2H_6	-	-	-	-	5.5
C_2H_4	0.42	0.43	0.24	0.32	-
C_2H_2	0.04	0.04	0.03	0.02	-
C_6H_6	-	-	-	-	-
		- Energy Content (MJ/kg) -			
LHV	14.5	14.6	7.89	10.4	48.3

The system will be binary, meaning that it either produces power or hydrogen but not both at the same time. Switching between the two modes presents several technical challenges due to the high degree of thermal integration in the base plant. The most significant technical challenges identified included:

1. Down stream water gas shift (WGS) catalyst beds are extremely sensitive to air exposure and therefore syngas flow must be maintained or they must be effectively sealed if shut down.

2. The pressure swing adsorption off-gas from hydrogen production is burned (with a small amount of trimming natural gas) to heat the Tar Reformer Catalyst Regenerator. This energy must be replaced when the PSA is not running.
3. The base design includes a thermally integrated steam cycle, which is partially fed by syngas cooling steps downstream of the split location. Loss of this heat energy to the steam turbine will adversely affect the plant power system.

In the base design, pressure swing adsorption (PSA) off-gas is burned with a small natural gas feed in the tar reformer catalyst regenerator. The natural gas is needed because of the low heating value of the PSA off-gas (approximately 3 MJ/kg). The energy balance and mass flows into the catalyst regenerator unit in the base design are summarized in Figure 5.5.

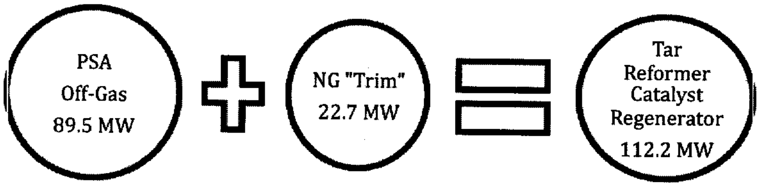


Figure 5.5 - Tar Reformer Catalyst Regenerator Flows

If all of the syngas stream is diverted to a gas turbine for power generation for any period of time then the tar reformer will cease to function. Instead, some fraction of the syngas stream must be used to make up the lost energy flow when the PSA system is shut down. The proposed solution divides the syngas stream directly after the LO-CAT sulfur removal step so that a fraction is sent downstream through the water gas shift reactors and then sent to the tar reformer catalyst regenerator to be burned rather than going through the PSA unit. This approach should effectively keep the sensitive WGS catalysts from air and also meet the energy demands of the tar reformer system. In addition, it is assumed to keep all base design systems warm and active (except for the PSA unit) until peaking is completed.

The pressure ratio between the feed and purge gas of a PSA is critical. Higher purge pressures result in significantly lower hydrogen recovery rates. The current design uses a feed pressure of 25-bar (360 psi) and a purge pressure of 1.4-bar (20 psi). Because pressure ratio is so critical to maintain proper function of the PSA unit, continuing to produce hydrogen with a smaller flow (lower pressure) does not appear to be a viable option. Because the reactor runs at low temperature and has little condensable water, a properly insulated PSA unit should be capable of being shut down for several hours at a time.

In order to maintain the base plant steam system, some portion of the exhaust gases from the gas turbine will be run through a heat exchanger. The size and design of the heat exchanger will be such that it exactly replaces heat loses due to the smaller flow in the water gas shift heat exchangers.

5.1.2 Modeling

ASPEN Plus simulation software (AspenTech) was used to test the feasibility of running the design scenario previously described. The existing biomass to hydrogen simulation was modified to include the power generation system running at steady state. Detailed discussion of the ASPEN modifications can be found in Appendix B.

Splitting the syngas flow after the LO-CAT reactor was done so that the flow of trimming natural gas to the tar cracker and the flow of PSA off-gas were both replaced with syngas. Specifically, enough syngas flow was maintained through the water-gas shift reactors to exactly meet the energy demands of the tar reformer catalyst regenerator. The PSA will be shut down during power generation. The remaining flow at the split is sent to the gas turbine. Figure 5.6 shows the split of syngas that meets the design requirements.

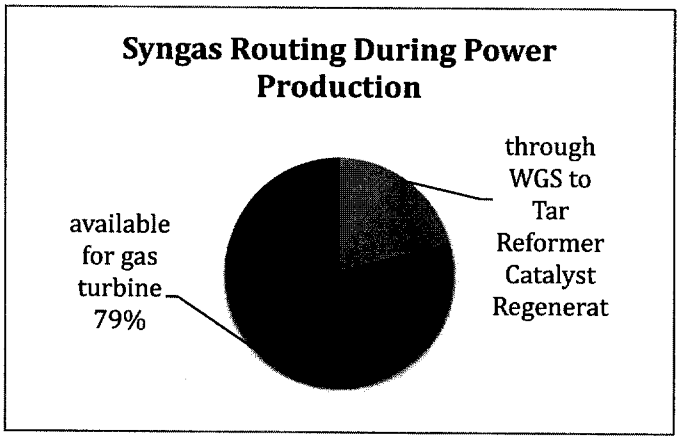


Figure 5.6 - Syngas Stream Split

Given the quantity and properties of the syngas available for combustion, a gas turbine was selected for the plant. Simple cycle gas turbines are typically used for peak shaving applications because they can be quickly brought on and off-line. Combined cycle systems have significant start-up times due to the large amount thermal mass involved with steam generation. One possible alternative is to run the combined cycle system constantly but at a lower power level until needed. While the NREL BIGCC study (Craig et al. 1996) used an integrated combined cycle system (ICC) for power generation, a simple cycle gas turbine was selected for this analysis because of the need to cycle power production on and off rapidly. The turbines considered are summarized in Table 5.2.

The GE MS6001FA (6FA) simple cycle turbine was chosen for this analysis. It is rated at 75.9 MW power output using natural gas (GE 2009). Both the 6FA and 7EA turbines have the same approximate power output but the F class turbines are significantly more efficient than older E class. Approximately 100 F class turbine units are functioning worldwide with more than 2 million combined operating hours. F class turbines have been documented to provide high reliability (GE 2009). Both the 6FA and the 7EA turbines can be run on synthesis gas. The 6B turbine was initially thought to be the proper size but was discarded as an option once the actual split fractions were determined in ASPEN. The energy content in the syngas is not sufficient to use the larger 7FA family of turbines.

Table 5.2 - GE Turbine Specifications[1]

	MS6001B	*MS6001FA*	*MS7001EA*
Output (MW)	42.1 (42)	75.9 (90)	85.1 (90)
Heat Rate (kJ/kWh)	11,226	10,332	11,002
Pressure Ratio	12.2:1	15.7:1	12.7:1
Mass Flow (kg/sec)	141	204	294
Turbine Speed (rpm)	5,163	5,254	3,600
Exhaust Temperature (°C)	548	603	536
Turbine Inlet Temp (°C)	1104	1288	1113
ICC Version	S106B	106FA	107EA
ICC Power Output (MW)	64.3 (63)	118.1 (130)	130 (130)

[1] Heavy duty gas turbine products. 2009. General Electric Company.

Several special considerations must be made when running gas turbines on a non-standard fuel. The turbine is typically limited by some inlet choke flow that is determined by the compressor flow and mass flow of natural gas under standard conditions. When a non-standard (low heating value) fuel is burned, a higher fuel feed rate is typically needed, which can cause choking of the turbine and compressor stall. The 6FA can get up to a 20% up-rate when run on syngas (from 75 to 90 MW) depending on the syngas composition. Allowing more mass flow through the turbine produces the extra power output. As more mass is forced through the turbine, the compression ratio increases since the turbine and compressor are coupled. According to GE technical papers (Brdar et al.

2000) and discussions with a GE representative (Ching-Jen Tang, April 2009), it is best to assume less than 14% deviation from catalogue flow rates to avoid compressor stall.

In addition to high fuel flow rates; cooling issues can arise depending on the syngas composition. Increased burner temperatures can shorten the service life of a turbine as well as drastically increasing NO_x and SO_x emissions. Based on data from previous GE gasification projects (Drdar et al. 2000) and previous research done at NREL (Craig et al. 1996), syngas compositions similar to the ones used in this study are typically humidified with steam before combustion. Steam was added to the syngas so that the final fuel gas was 20% H_2O by weight. This corresponds to a lower heating value of approximately 11 MJ/kg (174 Btu/ft^3).

To estimate the power output of a GE 6FA or similar gas turbine running on the available syngas composition, an ASPEN Plus model was made for the turbine and inserted into the simulation. Details of the ASPEN model and its calibration can be found in Appendix C.

Several important questions remain about the concept that cannot be addressed with this thermodynamic model. The most important question is how the system will behave dynamically when switched between hydrogen production and power production. This question cannot be addressed with the existing model; either physical testing or dynamic simulation (with additional detailed plant design and component performance characteristics) would be required. Also, the turbine outputs for this simulation are realistic and representative but detailed combustion analysis and testing would be required to determine the plant power output more precisely.

The final results are summarized in Table 5.3. When making hydrogen fuel, the plant would have the major input and output variables shown in the H_2 Production column. The Power Production column shows the input and output variables when the PSA unit is shut down.

Table 5.3 - Baseline Peaking Plant I/O[1]

		H_2 Production Mode	*Power Production Mode*
Inputs (kW)	Biomass Feed[2]	433,971	433,971
	NG Feed[3]	22,903	-
	Electricity	10,287	-
Outputs (kW)	Electricity	-	77,400
	H_2	232,074	-

[1] Where applicable, mass flows were converted to energy flows with the following lower heating values.
[2] A biomass LHV of 17.25 MJ/kg bone-dry wood was used.
[3] Natural Gas was assumed to have an LHV of 47.14 MJ/kg.
[4] Hydrogen was assumed to have an LHV of 120.21 MJ/kg.

The power production value of 77.4 MW represents what a "rubber" turbine[7] with GE F-class efficiencies and an assumed nameplate capacity of 80.6 MW could produce if all of the available syngas where utilized. A lower value of 72 MW would be produced if a GE 6FA turbine were used for the plant. This number is lower because there is a small amount of syngas available that cannot be used by a 6FA turbine.

When running in a peaking capacity, the plant has the relatively low total efficiency of 17.8%. This is compared to an efficiency of 49.7% when producing hydrogen and an expected turbine efficiency of approximately 32%. The extremely low power production efficiency is the result of multiple factors including the fact that 21% of the syngas stream is used to maintain the water gas shift reactors and tar cracker rather than for power

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[7] The term "rubber" turbine refers to the fact that the turbine size was set to exactly match the available fuel stream available as opposed to using an existing, stock frame size that could not utilize all of the available fuel.

production. In addition, a portion of the power output is used to provide power to the plant that is provided by the grid during hydrogen production.

The heating value of the syngas decreases after the water-gas shift reactors. This decrease is due primarily to the conversion of carbon monoxide into other species, If, rather than sending the syngas for the tar reformer catalyst regenerator through the water-gas shift reactor before combustion, the syngas were burned immediately, less gas would be needed (17% rather than 21% of the total syngas flow). Shutting off the water-gas shift reactors completely would make more syngas available to the "rubber" turbine and 82.3 MW could be produced. This range of values (~70 to 85 MW) will be used for sensitivity analysis in the economic model.

5.1.3 Alternative Design Scenario

Simple cycle gas turbines are typically used for peaking applications because they can be quickly cycled on and off. Combined cycle systems have significant start-up times (for example it can take over 8 hours to cold start a 500 MW combined cycle system when gas turbines alone can start in 12-15 minutes) due to the large amount thermal mass involved with steam generation. One possible alternative is to run the combined cycle system constantly but at a lower power level until needed. This scenario was briefly investigated to provide data for economic analysis.

The 106FA combined cycle from GE is rated at 118 MW nominal output, which is up-rated to 130 MW when run on synthesis gas (Heavy duty gas turbine products, 2009). Using the ASPEN Plus simulations already constructed, and specifications for the 106FA, estimates were made on the power output possible if a combined cycle were used instead of a simple cycle turbine.

The proposed system would require a steam turbine to operate at approximately 30% of its rated power during hydrogen production and 100% of its rated power during power production. While this is an extreme swing, it is possible with existing technology assuming large efficiency losses and complex control systems are acceptable. The following approximations were made based on discussions with a GE representative (Ching-Jen Tang, April 2009):

- When the steam cycle is operating at 100% rated power its thermal efficiency is 40%
- When the steam cycle is operating at 30% rated power its thermal efficiency is 30%
- Approximately 1 hour will be required to ramp up from 30% to 100% rated power.

Base on the assumptions listed above, power production model outputs were derived for this alternative scenario. A summary of the values is given in Table 5.4. The plant is assumed to operate in power production mode 20% of the time and in hydrogen production mode the additional 80% of the year.

Table 5.4 - Peaking Plant I/O Combined Cycle

		H_2 Production Mode	*Power Production Mode*
Inputs (kW)	Biomass Feed	433,971	433,971
	NG Feed	22,903	-
	Electricity	17,781	-
Outputs (kW)	Electricity	-	125,710
	H_2	232,074	-

The power production value of 125.7 MW represents what a "rubber" turbine with GE F-class efficiencies could produce if all of the available syngas where utilized and exhaust heat recovered. A lower value of 110 MW would be produced if a lower efficiency steam cycle were used for the plant.

When running in a peaking capacity, the plant has a significantly better total efficiency than that of the gas turbine only. The increase from 17.8% to 29% should increase the economic viability of the system. However, it is still low compared to an efficiency of 48.9% when producing hydrogen.

If, rather than sending the syngas for the tar reformer catalyst regenerator through the water-gas shift reactor before combustion, the syngas was burned immediately, less gas would be needed. This would make more syngas available to the "rubber" turbine and 144.98 MW could be produced. This range of values (~110 to 150 MW) will be used for sensitivity analysis in the economic model.

5.1.4 Capital Costs

The additional costs to the existing Central Biomass Goal H2A analysis are summarized in Table 5.5. The gas turbine numbers were taken from the 2006 GTW Handbook and then adjusted to 2005 dollars using the Chemical Engineering Plant Cost Index.

Table 5.5 – Capital Cost of Gas Turbines

	2006 Dollars	*2005 Dollars*
Uninstalled Gas Turbine ($/kW)	247.00	231.48
Installation Factor	1.8	1.8
Total Installed Cost ($/kW)	444.60	416.66

The cost associated with additional steam turbine capacity (used in the combined cycle alternative design scenario) was taken directly from the baseline Biomass to Hydrogen report (Spath et al. 2005) and shown in Table 5.6.

Table 5.6 - Capital Cost of Steam Turbines

	2002 Dollars	*2005 Dollars*
Steam Turbine ($/kW)	474.34	561.39

Using these values and heat exchanger prices taken from the baseline Biomass to Hydrogen report, the additional capital costs for both the simple cycle and combined cycle peaking systems were calculated and summarized in Table 5.7.

Table 5.7 - Concept 1 Peaking Capital Costs

Combined Cycle Additional Capital Costs		
	Simple Cycle	**Combined Cycle**
Turbine	$ 32,249,794	$ 32,249,794
Exhaust BFW Preheater	$ 247,729	$ 247,729
Steam Turbine	$ 0	$ 27,120,801
Total	**$ 32,497,523**	**$ 59,618,324**

5.2 Sinking Modifications

Indirectly-heated gasification is a two-stage process where the heat needed for reaction is produced by burning char in a separate chamber to heat sand. This hot sand is then circulated through the reaction chamber to drive reaction kinetics. The goal of the system is for the gasifier to produce enough char to heat the reaction zone to an optimal temperature.

There is a direct correlation between the reactor temperature and the amount of syngas produced from a given amount of biomass. Higher reaction temperatures favor syngas production over char and tar production. Lower reaction temperatures cause increased tar formation and char. The relationship, as reported by previous correlations (Bain 1992), can be seen in Figure 5.7. Because of this correlation to temperature, indirectly-heated gasifiers will reach an equilibrium temperature if left in steady-state.

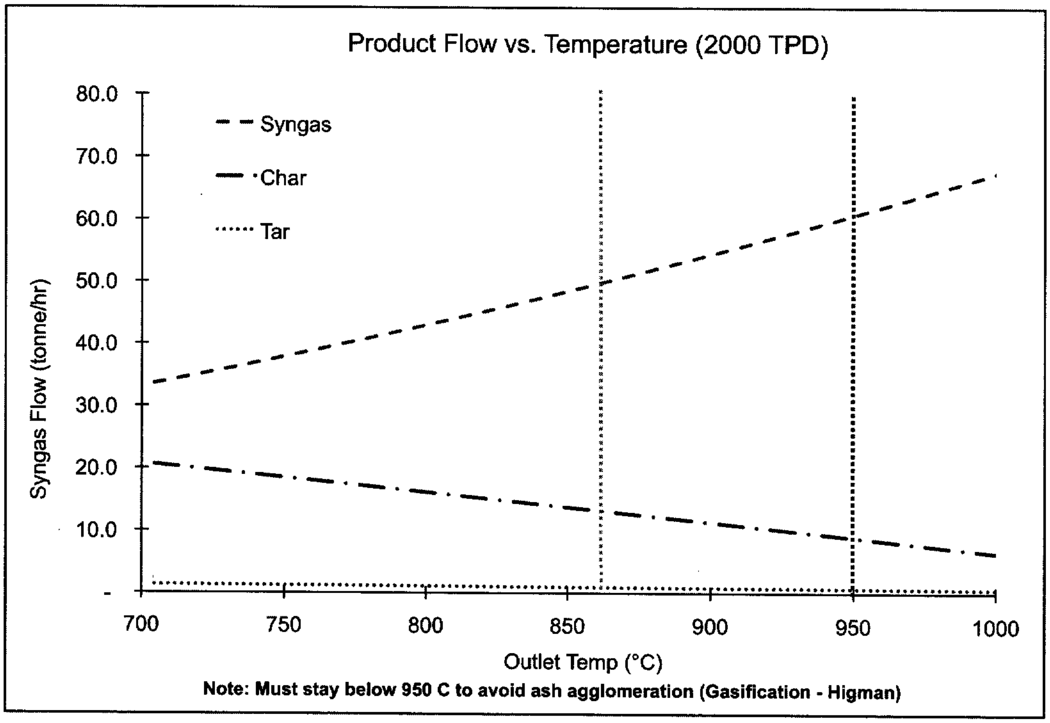

Figure 5.7 – Indirect Gasifier Products vs. Temperature

The equilibrium temperature that a gasifier finds may be a less-than-optimal reaction temperature so that enough char can be produced to provide the required heat. Adding heat energy will displace the need for high char production and break some of the char and tar into additional syngas, which will increase process efficiency. Alternatively, electric heating could provide a replacement for any syngas recycle or natural gas trimming currently used to push the gasifier to an optimal reaction temperature. Use of excess wind power to provide the electricity would be one way to create a dispatchable load.

5.2.1 Plant Design

The original system concept involved electrically-heating the gasifier freeboard or sand recirculation path to increase the gasifier operating temperature. The proposed system modification can be seen in Figure 5.8.

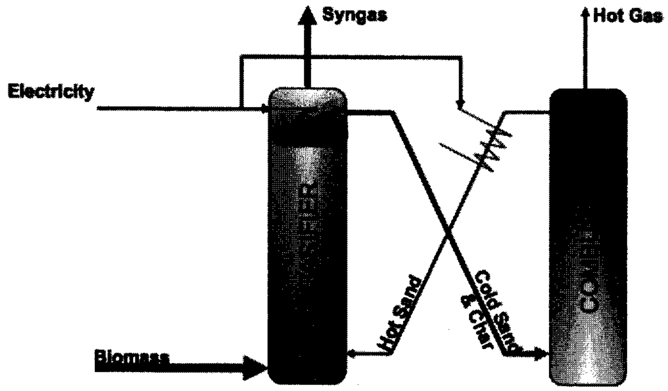

Figure 5.8 – Initially Proposed Sinking Modification

Based on the correlation data available at the time (Bain 1996), the equilibrium operating condition of the gasifier produced enough char to keep the reactor at an optimal 780°C. That temperature is high enough for primary tar destruction and low enough to avoid ash agglomeration (which occurs at temperatures around 950°C) (Higmann 2008). Electric heaters could be used to increase the gasifier temperature above equilibrium, but below 950°C, thus increasing the amount of syngas produced from each unit of biomass.

The initially proposed system (Figure 5.8) would produce a constant mass flow of syngas using a varying amount of biomass feedstock based on electricity availability. During times of low electricity demand, excess wind electricity can be used directly by the gasifier to increase the efficiency (syngas yield per ton of biomass). In times of high electricity demand, the gasifier would not use any electricity, instead it would be operated at typical BCL equilibrium conditions.

Using correlation data and a perfectly efficient electric heater (i.e. 1 kWh electricity equals 1 kWh of heat), an initial estimate of the amount of biomass that could be displaced by electric heat was made. As heat is added to the gasifier the temperature increases which in turn increases the amount of syngas produced by each unit of biomass. If the amount of syngas produced is held constant, then less biomass is required to produce the same flow of syngas. Figure 5.9 shows this correlation.

The relationship is approximately linear and shows that 1 kWh of electricity replaces up to 3.9 kWh of biomass (LHV basis) for a 2000 TPD indirect gasifier. In economic terms, biomass valued at $50/ton is worth approximately 1.01 ¢/kWh, meaning that electric heating could be preferable at any electricity costs lower than 3.94 ¢/kWh. These initial calculations made the prospect of intermittent electric heating seem promising even though use of high-grade energy (i.e. electricity) for heating is thermodynamically inefficient.

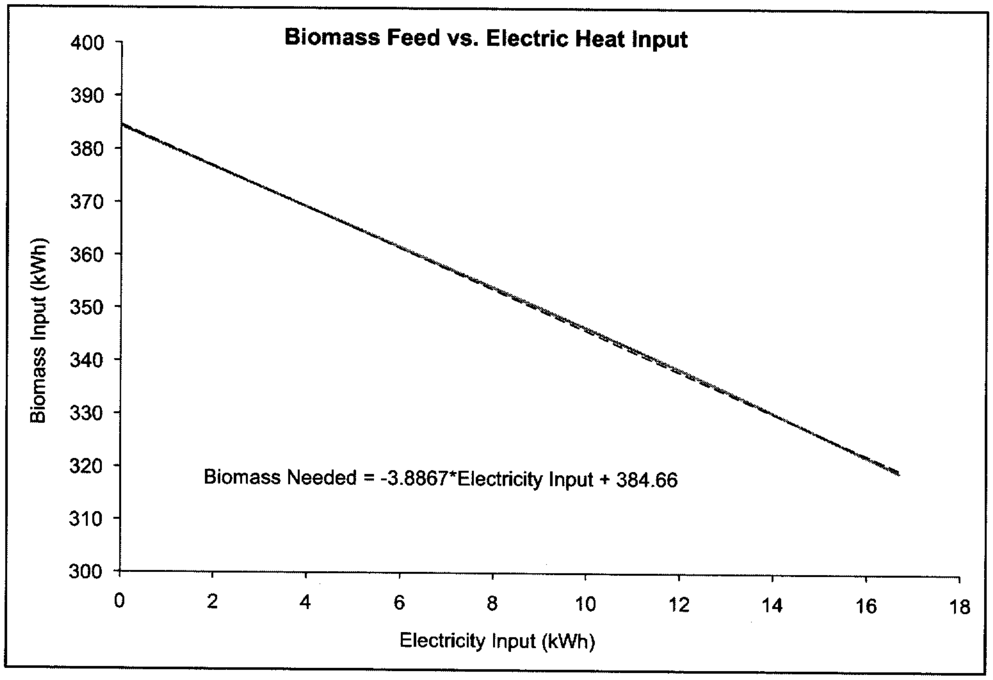

Figure 5.9 - Relationship of Heat Addition and Biomass Usage

Initial research into how to add electric heat yielded interesting findings. Embedding typical resistive heaters may not be feasible in the corrosive environment of the gasifier. However, one possibility is to use the fluidized bed itself as a "resistive element" through which to pass the electricity. The Institute of Gas Technology considered this possibility in the 1970's before discarding it. Their reports show that it was a technically feasible option but that electricity prices made it less cost effective than burning biomass. Using electricity selectively may change the economics.

As work began on finding pathways to electric heating of the gasifier, contact was made with the NREL Biomass Center. In the process of discussion, it came to light that the Biomass to Hydrogen model was being updated to include new yield correlations for the gasifier based on data collected in the Thermochemical Process Development Unit (Kinchin et al. 2009). The results of the updated model reveal that the gasifier does not produce enough char to maintain gasification temperatures, therefore raw syngas must be diverted and combusted to supplement the heat delivered to the gasifier by the char combustor. This new information dramatically altered the research path for the current study and negated the previous concept. A diagram of the updated gasifier with recycled syngas is shown in Figure 5.10.

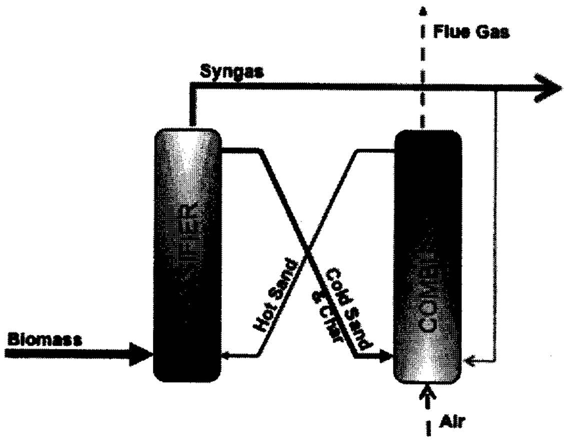

Figure 5.10 - Updated Indirect Gasifier Diagram

Based on the updated model, research shifted to supplementing the heat delivered to the gasifier by the char combustor with electrically generated heat rather than diverting and combusting syngas. The design effort initially focused on adding heat directly to the olivine (sand) as it returned to the gasifier. This approach proved unreasonable for two reasons:

- Electrically heating the olivine with currently available collar heaters (wrap around piping, Fig. 5.11) is not possible because commercial units are not available in the temperature and power range necessary for this application. While a dedicated olivine-heating vessel can be envisioned and assumed to exist, the efficiency and capital cost for such a unit will be difficult to estimate.
- The hot product gases from combusting the char and diverted syngas are used to dry the incoming biomass. If the diverted syngas is replaced with electric heaters, the olivine will receive the necessary heat to maintain gasification temperatures (approximately 870 C), but the combustion product gases used to dry the incoming biomass will be limited to product gases from the char combustor alone, which are not sufficient to for drying the biomass.

If the model is to be based on currently available technology, the most likely design will employ high power, high temperature electric air heaters (Figure 5.11). If the combustion air used in the syngas and char combustors is preheated, more heat can be delivered to the olivine per kilogram of char or syngas combusted, resulting in a reduced amount of syngas that must be recycled.

Combustion air pre-heaters are actually air duct heaters that are available up to 2.2 MW. The electric air heaters can also heat air for use in the biomass dryer; therefore several design scenarios using electric heaters are possible

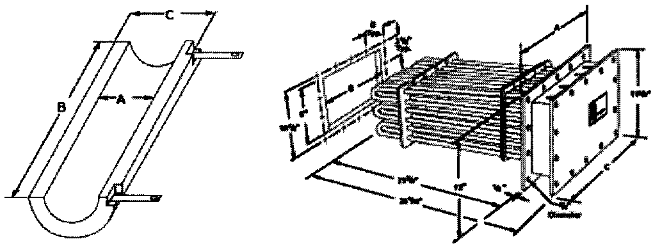

Figure 5.11 - Collar Heater and Air Duct Heater (Watlow 2000)

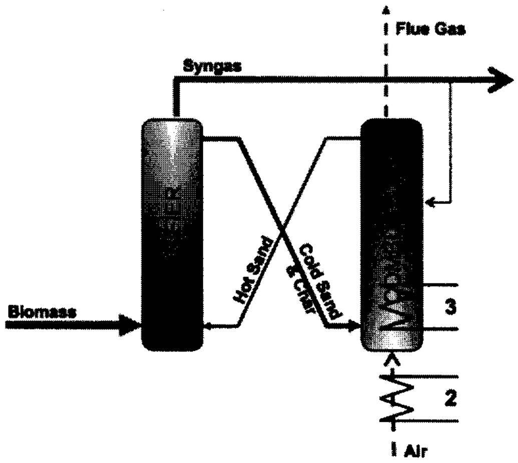

Figure 5.12 - Final Indirect Gasifier Sinking Modifications

1. **Base Case** No electric heat added. This design scenario is identical to the updated Biomass to Hydrogen model reported in (Kinchin et al. 2009). It provides a basis for comparing the design cases that employ electric heating.

2. **Electric Air Heaters Only** This design case limits the electric heat assist to currently available technology (e.g. the electric air duct heaters).

3. **Electric Air Heaters and Electric Olivine Heating Vessel** This design scenario uses electric air heaters to preheat the air entering the

combustor, electric air heaters to heat air used to dry the biomass, and an envisioned electric olivine heating vessel. This design scenario eliminates the need to divert any syngas for heating purposes. All heat required for gasification and biomass drying is supplied by combusting char, electrically heating air, and electrically heating olivine.

5.2.2 Modeling

ASPEN Plus thermodynamic simulation software was used to test the feasibility of running the design scenarios previously described. As mentioned above, these design scenarios were simulated using the updated Biomass to Hydrogen model (Kinchin et al. 2009) not the 2005 Biomass to Hydrogen model. An overview of the findings is given below:

Electric Air Heaters Only: Results of this design case were very promising. The amount of syngas that must be diverted is reduced by about 45%, and enough hot combustion products are still produced to dry the incoming biomass. Table 5.8 gives the detailed plant input and output values from the ASPEN Plus simulation. The plant is assumed to operate with electric air heaters on 20% of the time and without them the additional 80% of the year.

Table 5.8 - Sinking Plant I/O with Electric Heaters Only

	Units	*No Electric Heat*	*Electric Air Heaters Only*
Biomass Feed In	kg/hr	83,333	83,333
Electricity In	kW	6,794	53,661
Natural Gas In	kg/hr	3,265	3,744
Electric Heat Demand	kW	-	77,629
Electric Heater Power	kW	-	86,573
Power for Syngas Combustion Air Blower	kW	1,858	3,885
Hydrogen Out	kg/hr	7,134	8,005

Electric Air Heaters and Electric Olivine Heating Vessel: As expected, results of this scenario indicate increased hydrogen yield, but at the expense of increased power requirements. Table 5.9 gives the detailed plant input and output values from the ASPEN Plus simulation. The plant is assumed to operate with electric heating 20% of the time and without the heaters the additional 80% of the year.

Table 5.9 - Sinking Plant I/O with All Electric Heat

	Units	*No Electric Heat*	*All Electric Heat*
Feed In	kg/hr	83,333	83,333
Electricity In	kW	6,794	110,927
Natural Gas In	kg/hr	3,265	4,348
Electric Heat Demand	kW	0	98,820
Electric Air Heater Power	kW	0	86,573
Electric Heater Tank Power	kW	0	23,633
Power for Syngas Combustion Air Blower	kW	1,858	5,828
Hydrogen Out	kg/hr	7,134	9,306

Another way to compare gasifier results is through cold gas efficiency. Cold gas efficiency is defined as the chemical energy in the syngas leaving the gasifier divided by the energy in the biomass feedstock (and electric heat added in this case). Specifically, cold gas efficiency was calculated by dividing the sum of the chemical energy output on a lower heating value basis divided by the energy in the biomass feed on a lower heating value basis plus the electric heat added. The baseline indirectly-heated gasifier (without electric heat) has a lower cold gas efficiency than the electrically heated gasifier. The baseline cold gas efficiency was 67% and it increased to 71.5% when electric heat fully replaced the syngas recycle stream.

In order to enter these scenarios into the economic model, and so that the sinking and peaking cases could be directly compared, the ASPEN results were scaled to the same baseline case that was used in the peaking analysis (reference Figure 5.4 without the modifications highlighted in red). In addition, the economic model requires all inputs in terms of kW so energy content of the mass flows was calculated using the lower heating value. The scaled and converted results are shown below for each of the two cases under consideration.

Table 5.10 - Final Sinking I/O for Air Heaters Only

	Units	*Baseline*	*Heaters*
Feed In	*kW*	433,971	433,971
Electricity In	*kW*	10,287	57,154
Natural Gas In	*kW*	22,903	27,977
Hydrogen Out	*kW*	232,074	261,150

Table 5.11 - Final Sinking I/O for All Electric

	Units	*Baseline*	*Heaters*
Feed In	*kW*	433,971	433,971
Electricity In	*kW*	10,287	114,420
Natural Gas In	*kW*	22,903	34,367
Hydrogen Out	*kW*	232,074	304,598

Unlike the peaking modification, which produced lower efficiency, adding electric heat to the plant actually increases the total plant efficiency (energy in/energy out) from 49.7% to 50.3% (or 52.2% for the all electric option). Each additional unit of energy input as electricity produces 0.56 to 0.63 units of hydrogen energy output. Therefore, while providing a dispatchable load service to the local utility, the plant actually operates more efficiently.

5.2.3 Capital Costs

The electric heater capital costs came from a quote provided by Watlow for their largest, high-temperature air duct heater. Based on that quote, a 2.2 MW air duct heater costs $250,000 and has an electricity to heated air conversion efficiency of 90%. Additional unit savings are not expected with increasing scale so a scaling factor of 0.9 was used. Finally, an installation factor of 2.47 was assumed for all cost estimates. The installation factor of 2.47 was used for consistency with other H2A analyses (Spath et al. 2005). Based on these assumptions, the total additional capital costs for electrically heating the combustion air came to $15.8 million.

For the all-electric heat case the same costing assumptions were used for electric duct heaters. The cost of adding electric heating to the olivine was estimated by taking the cost of the Inconel heating coil required for delivering the given heat and multiplying it by 2.5. This is a very rough way to approximate the system, but because little is known of the actual design of such a heater, it provides a starting point. If the system is extremely close to economical or sensitive to capital cost then sensitivity studies will be needed on this value. These assumptions yield an additional capital cost of $17.8 million. Detailed costing information for both scenarios can be found in Appendix D.

CHAPTER 6

DIRECT GASIFIER HYBRID SYSTEM

This concept is based on directly-heated gasifier architecture. Directly-heated gasifiers typically have a single combustion/reaction chamber and burn a small portion of the biomass feed to create heat. A source of pure oxygen is required for combustion if the syngas is to be used for fuel production. Electrolysis could provide an alternative to an air separation unit with the added benefit of producing an additional pure hydrogen stream. The initial concept block diagram is shown in Figure 6.1.

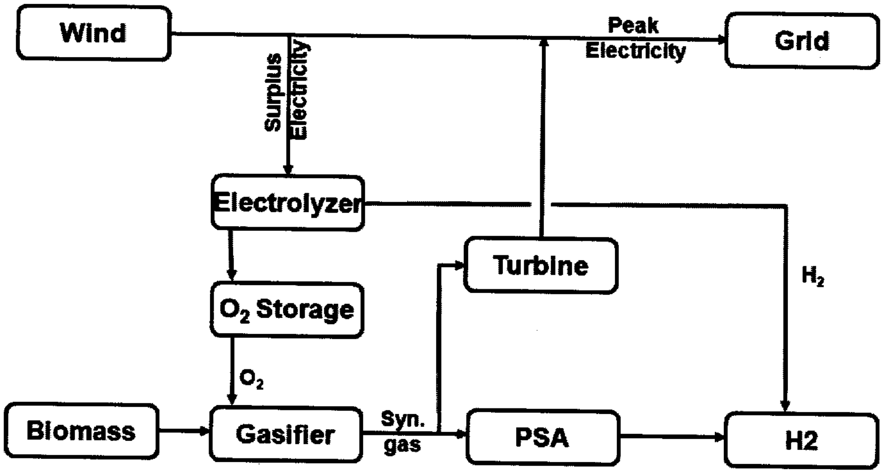

Figure 6.1 – Direct Hybrid System Block Diagram

There has been some research to date on the feasibility of combining electrolysis with gasification (Gassner 2008). The Gassner paper concluded that the economic feasibility of the combination was highly dependent on the price of available electricity.

The proposed hybrid system could directly address electricity price dependence by running the electrolysis system intermittently. Electricity available during periods of low electricity demand (low purchase price) will be used by electrolyzers to produce oxygen and hydrogen for use by the gasifier and or stored for later use. During periods of peak electricity demand, the stored oxygen will be used to create syngas rather than running the electrolyzers.

6.1 Plant Design

In order to characterize the possible benefits of this concept a "baseline" gasification plant and a hybrid plant were simulated in ASPEN Plus and compared. Both the baseline and hybrid plants have the same general structure; biomass is gasified to produce syngas in a high-pressure gasifier. Tars are removed and the syngas is cooled before entering a sulfur removal step. After sulfur has been removed, the carbon monoxide in the syngas stream is shifted towards hydrogen. Finally, the hydrogen is stripped from the syngas and the remaining gas is burned to generate power. A block diagram of the baseline plant is shown in Figure 6.2.

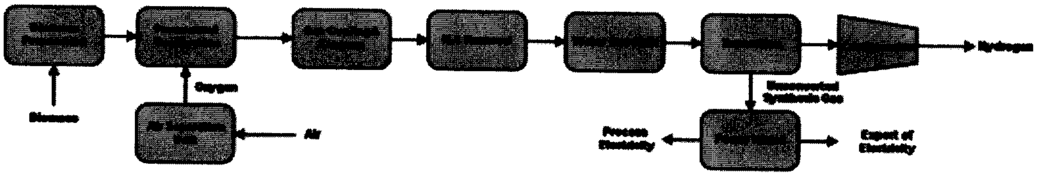

Figure 6.2 – Baseline System Block Diagram

In the mid 1980's the Institute of Gas Technology (IGT) developed a pressurized, steam-and-oxygen blown, fluidized-bed biomass gasification process for the Pacific Northwest Laboratory (PNL) and the United States Department of Energy (DOE). A twelve-ton per day process research unit (PRU) was built and tested using wood chips for feedstock. The results of these tests are publically available (Evans et al. 1988 and Bain 1992) and were used as the basis for the directly-heated gasifier design used in this report.

The ability to feed biomass against pressure is arguably the largest technical hurdle to pressurized biomass gasification. The only proven technique for feeding low-density feedstock into a pressurized reactor is using lock-hoppers (Lau et al. 2003). Nitrogen is a common inert pressurization agent for use with lock-hoppers. It is available from the cryogenic air separation unit (ASU) for oxygen-blown gasifiers.

Figure 6.3 shows the general arrangement of the assumed gasifier design. A bed temperature of 860°C was chosen so that primary tars would be cracked inside the bed but ash agglomeration could be kept to a minimum. The gasifier operates at 24-bar and incorporates several modifications to the experimental results published by Evans et al. A dual lock-hopper feed system was substituted for the single lock-hopper design used in the PRU. Switching to the dual lock-hopper system and optimizing the lock-hoppers will result in less inert gas usage per kilogram of biomass fed (Lau et al. 2003). In addition, partial oxidation was added to the gasifier freeboard. Based on literature review findings, it is assumed that a freeboard outlet temperature of 1030°C would result in an 88% decrease in tar content (Pan et al. 1999) and a 60% reduction in ammonia leaving the gasifier (Devi et al. 2003). The increased freeboard outlet temperature should not cause additional ash agglomeration because it is well above the particles in the fluidized bed.

The biomass feedstock used in this analysis is untreated hybrid-poplar wood chips. Data on the feedstock composition was taken from the Phyllis biomass properties database (Phyllis Biomass Datapage). It is similar, but not identical, to the feedstock used in the indirectly-heated cases. Cold gas efficiency for this gasifier design was 71.3%. Cold gas efficiency is defined as the chemical energy in the syngas leaving the gasifier divided by the energy in the biomass feedstock.

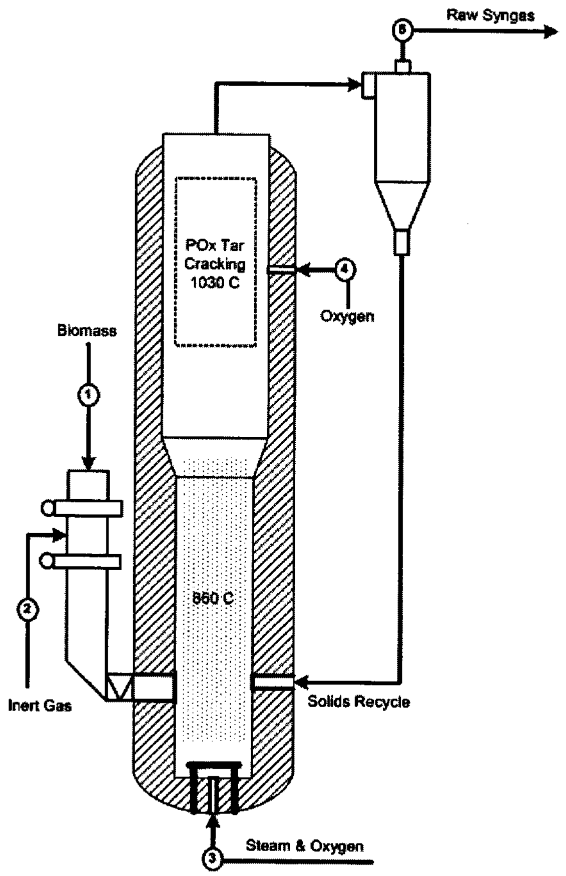

Figure 6.3 – Direct Gasifier Diagram

Low moisture content for the biomass was assumed in the direct model so that drying of the biomass was not required before gasification. The indirect case feedstock assumed moisture contents of farmed trees as high as 50 weight-percent at the plant gate being dried before gasification (Spath et al. 2005). Drying (and therefore energy) requirements are highly dependent on the feedstock type. Therefore, the direct biomass gasification baseline case has an energy advantage over the indirectly-heated biomass gasification baseline.

After the syngas leaves the gasifier, tar decomposition/removal is accomplished in two tar-decomposition reactors followed by a wet scrubbing step. The decomposition reactors catalytically breakdown the tars and the wet scrubbing step removes the residual tar that is not decomposed in the reactors.

The first tar decomposition reactor is a fixed-bed reactor packed with a dolomite silica sand mixture. Based on laboratory scale experiments a tar decomposition percentage of 62.5% is possible (Devi et al. 2005). The syngas stream in the baseline model is at higher temperatures than the ones reported in literature and, based on the trends in the laboratory scale experiments, it is expected a higher conversion percentage of 65% is possible.

The second decomposition reactor is a fixed-bed reactor with a nickel-based catalyst. Nickel-based catalyst is significantly more reactive than dolomite but is also more sensitive to impurities in the syngas. By removing a large percentage of the tars in the dolomite bed prior to the Ni-based catalyst reactor, tar concentration levels are reduced to less than 10 g/Nm^3, which is a level below which literature reports that catalyst deactivation is a concern (Aznar et al. 1998 and Wang et al. 2000). The reactor converts tar, ammonia, ethane, ethylene, and methane at the conversion efficiencies shown in Table 6.1.

Table 6.1 - Tar Decomposition Reactor Conversion Efficiencies

Compound	*Conversion Efficiency*
NH_3	95%
Tar	99%
C_2H_4	90%
C_2H_6	90%
CH_4	90%

A wet scrubber removes remaining particulate and tars in the syngas stream. Wet scrubbing is the only commercial method currently available for removing particulates and tars at the operating conditions present in this study (Milne et al. 1998). The assumed scrubbing performance can be seen in Table 6.2.

Table 6.2 - Wet Scrubber Removal Efficiencies

Compound	*Removal Efficiency*
NH_3	99%
Tar	95.8%
Particulates	99.9%

A significant amount of water is required for wet scrubbing and it is contaminated during the scrubbing process. A venturi wet scrubbing system was used for calculations. Approximately 106-gallons of water are circulated through the venturi for each cubic meter of gas (Liu 2005). The bulk of the water can be recycled but some must be purged to avoid contaminant build-up. A purge of 5% of the flow was assumed.

Sulfur compounds must be removed from the syngas prior to the water-gas shift reactors to avoid poisoning the catalyst. Hydrogen sulfide (H_2S) is the dominant sulfur compound produced by biomass gasification. One of the most economical methods to remove H_2S from syngas is to use a LO-CAT process followed by a ZnO polishing bed (Spath et al. 2005).

The LO-CAT process is capable of reducing H_2S to approximately 10-ppmv (Spath et al. 2005). While this purity level is sufficient for combustion, it is not sufficient to protect downstream fuel production catalyst and so a second polishing step is required (Van der Drift et al. 2006). Depending on the reactor design and H_2S concentrations in the syngas, ZnO beds are capable of producing a syngas with a H_2S

concentration in the ppb range (Spath et al. 2005). For the catalysts considered in this study it is sufficient for the ZnO reactors to reduce H_2S concentrations to the 1-ppmv range.

The LO-CAT process is significantly less expensive than the Rectisol, Selexol, or Amine-type sulfur removal processes with which it competes. This was the primary reason for its selection. However the LO-CAT process does not provide for CO_2 capture.

To produce the maximum amount of hydrogen possible from the syngas, a two-stage water-gas shift system is used to shift some of the available CO to hydrogen. Commercial high temperature WGS systems are typically fixed bed reactors that use an iron or chromium oxide-based catalyst. For high temperature WGS, catalyst activity drops off for temperatures less than 300°C. Commercial low temperature WGS systems operate at temperatures less than 250°C and use zinc or copper oxide-based catalysts.

Hydrogen is separated from the syngas using a pressure swing adsorption (PSA) system. Pressure swing adsorption is a commercial technology in wide use today. Special adsorptive materials are used to preferentially adsorb hydrogen at high pressure. Then the reactor swings to low pressure to release, or desorb, the hydrogen. Because the process is non-continuous, multiple units in parallel are required.

For optimum performance, several constraints must be met. Absorption efficiency decreases with increasing temperature, and the absorptive materials are extremely sensitive to entrained liquids. Therefore, syngas is cooled and condensable liquids removed with a flash drum before entering the PSA. A pressure ratio of four to one is maintained so that 90% of the inlet hydrogen is captured at purity greater than 99.9% (Spath et al. 2005).

In order to achieve 90% hydrogen capture from the PSA it is important that there be greater than 70% molar fraction in the syngas inlet stream (Spath et al. 2005). The syngas leaving the final water gas shift reactor is significantly below that level. To meet the 70% requirement, some of the product hydrogen is compressed and recycled to the PSA inlet.

The PSA off gas is composed primarily of carbon dioxide, carbon monoxide and nitrogen. Its quality is too low for use in a gas turbine (Drdar et al. 2000) and so it is burned to provide heat and power to the plant.

6.1.1 Baseline Plant

As previously mentioned, a source of pure oxygen is required if the syngas is to be used for fuel production so that nitrogen dilution does not effect downstream processing. Currently, plants that use oxygen produce it with cryogenic air separation units (ASU).

A cryogenic air separation unit (ASU) is used to produce the oxygen at 95% purity for the baseline study. In addition to producing oxygen, a nitrogen stream of 97% purity is available as a byproduct from the ASU and is used as the inert pressurization agent for the lock-hopper system. According to one source, 287–375 kWh of electricity are needed for every tonne of oxygen produced (American Water Works Association). Based on this 350 kWh per tonne of oxygen was assumed. This value does not include compression of the oxygen and nitrogen product gases, which exit the ASU at approximately 5-bar (Smith et al. 2001).

Figure 6.4 shows the envisioned biomass to hydrogen pathway using the direct gasifier and a cryogenic air separation unit. The major differences between this pathway and that detailed in Figure 5.4 is the addition of an ASU and the alternative gasifier architecture.

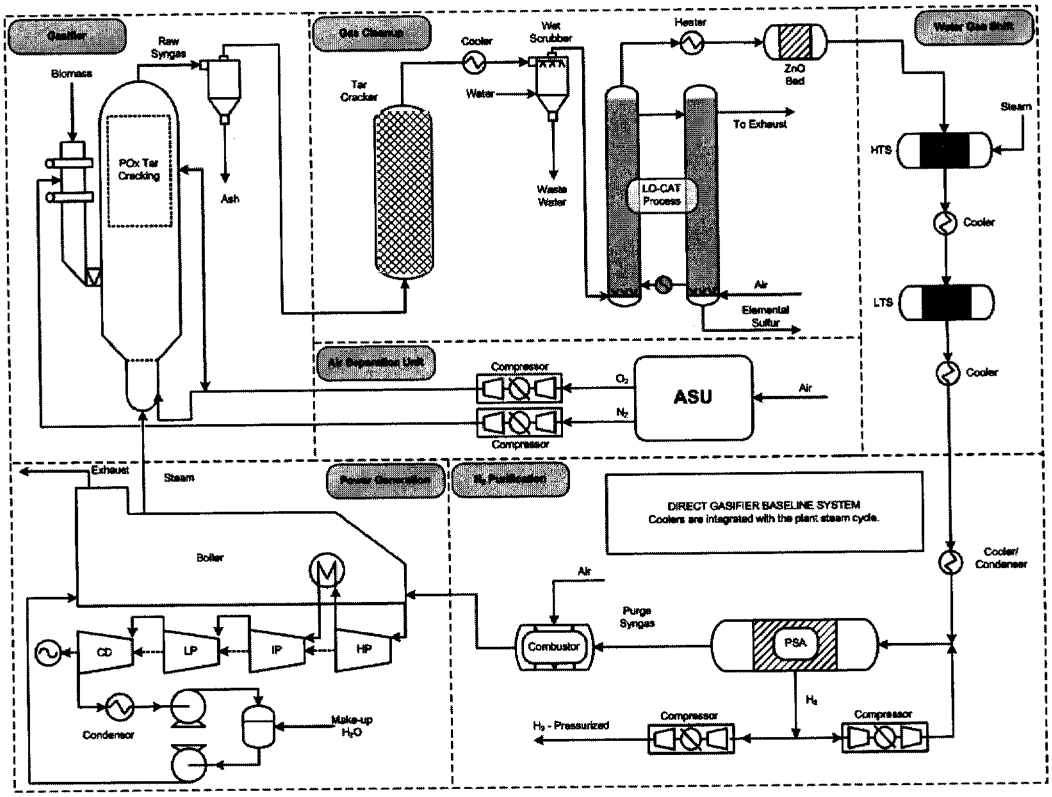

Figure 6.4 – Direct Gasifier Biomass to Hydrogen Plant

6.1.2 Hybrid Plant

Electrolysis could provide an alternative to an air separation unit with the added benefit of producing a pure hydrogen stream. Key changes to the plant would be required including (1) replacing the entire ASU with an electrolyzer bank and (2) replacing the LO-CAT/ZnO sulfur removal steps with a two-stage Selexol plant. The sulfur removal change is driven by the need for an inert gas for feed pressurization. The envisioned plant is shown if Figure 6.5.

To replace a single ASU unit for oxygen production, multiple electrolyzers are needed. The largest commercial electrolyzer is produced by StatoilHydro (formerly NorskHydro) and produces a maximum flow rate of 174 kg/hr of oxygen (43.6 kg/hr of hydrogen).

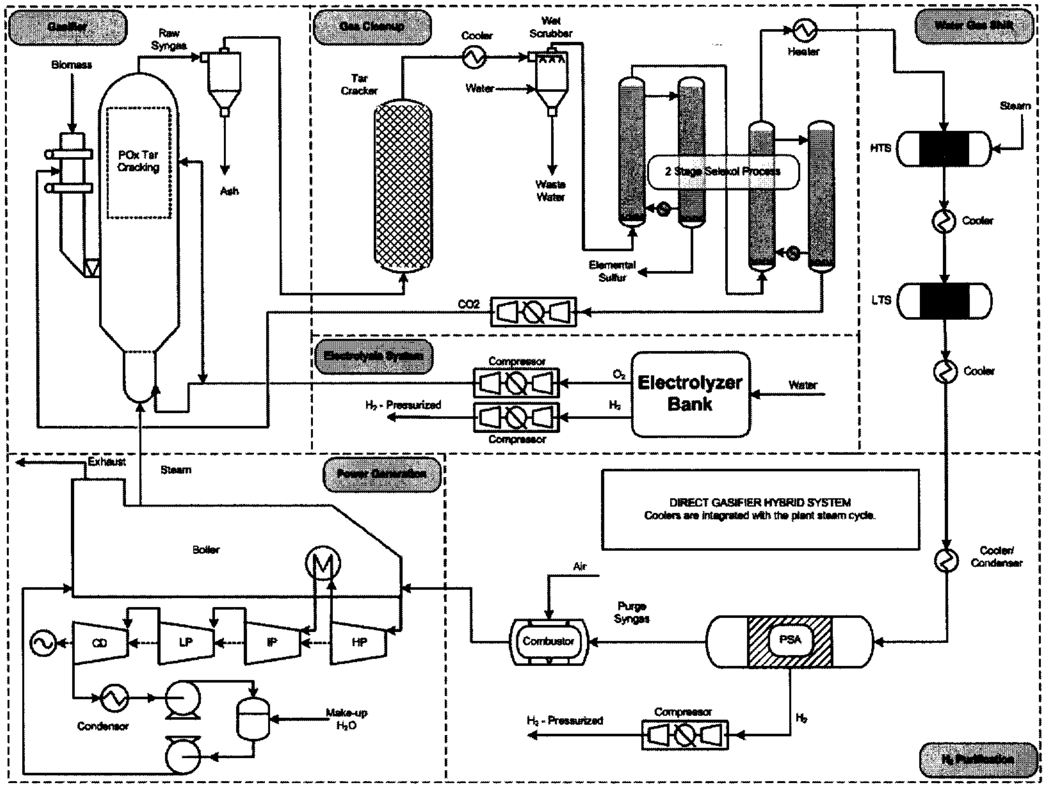

Figure 6.5 – Direct Gasifier Hybrid Concept

The StatoilHydro electrolyzer is an alkaline, low-temperature system. While significant research is ongoing for high-temperature, advanced electrolysis systems, commercial technology was chosen for this analysis. Low-pressure products were assumed, however alkaline electrolysis units have operated at pressures as high as 448-bar according to literature (Kroposki et al. 2006).

Based on ASPEN simulations, a 2000 TPD fluidized-bed, biomass gasifier with POx injection would need approximately 27,800 kg/hr of oxygen supplied. For this design, 160 electrolyzers would be needed to replace the air separation unit. This number assumes no spare hardware is required and that all electrolysis units run at design capacity 100% of the time.

Selexol is a well-proven process that uses a dimethyl ether-based solvent to remove sulfur compounds from the syngas stream. The process can be configured in a number of ways depending on the level of sulfur removal required. Sub part-per-million H_2S outlet concentrations are possible using Selexol (Kubek et al. 2000). It may also be configured to remove both sulfur compounds and CO_2. In this configuration, CO_2 removal rates of up to 90% can be achieved (Manning et al. 1991).

The baseline direct gasification model (Figure 6.4) assumed 0.15 kilogram of nitrogen was required per kilogram of biomass for feed pressurization. However, since no nitrogen stream is available from the ASU, CO_2 will be used for feed pressurization instead. Using CO_2 rather than N_2 as the inert gas requires that either occupy the same volume. Based on ideal gas assumptions, 0.23 kg CO_2 is needed per kg biomass fed.

For the proposed plant configuration to work, the Selexol process must produce enough CO_2 to replace the inert N_2 feed previously used for biomass feed pressurization. ASPEN Plus simulations confirm that there is more CO_2 available than needed for biomass pressurization. Excess CO_2 is vented to the atmosphere.

6.1.3 Integrated Steam Cycle

A steam cycle was thermally integrated with the plant to capture the significant heat rejected in the fuel synthesis process. In order to maximize power production, a Pinch analysis was performed for each of the cases.

Pinch analysis is a set of techniques developed to optimize the energy efficiency achieved by process integration (Kemp 2007). Using stream temperatures and heat content requirements of each plant, "hot" and "cold" streams are defined; A "hot" stream is one that must be cooled and a "cold" stream requires heating. For this analysis, the hot streams were defined for each plant by the individual gas clean-up process requirements. Once the hot streams were identified, a composite curve was constructed. This composite curve represents the combined total of all available thermal energy in the system.

A similar approach was taken in constructing the cold composite curve. The cold composite curve represents the combined total of all the heating requirements in the process. In order to maximize power production the amount of steam produced was set to recover as much thermal energy as possible. Due to the low temperatures of the condensation steps and other low temperature cooling requirements an on-site water-cooling system had to be provided in the model.

The composite curves for both cases can be seen in Figures 6.6 and 6.7. A pinch temperature of 20°C was used for both cases. The pinch temperature is the minimum allowable temperature deference between the hot and cold composite curves that still allows for heat transfer. Looking at the horizontal axis, significantly less heat energy is available in the hybrid system then is available for the baseline meaning that less electric power can be produced.

The change from nitrogen as an inert pressurization agent to CO_2 reduced the power production of the plant. The reduction in available energy was due to both a change in the specific heat of the syngas and a change in the syngas quantity being cooled. The most significant of those variables was the reduction of mass flow through the heat exchanger network. The Selexol plant in the hybrid case removes ninety-percent of the CO_2 in the syngas. This gas flowed through the entire system previously and stored heat that is now vented.

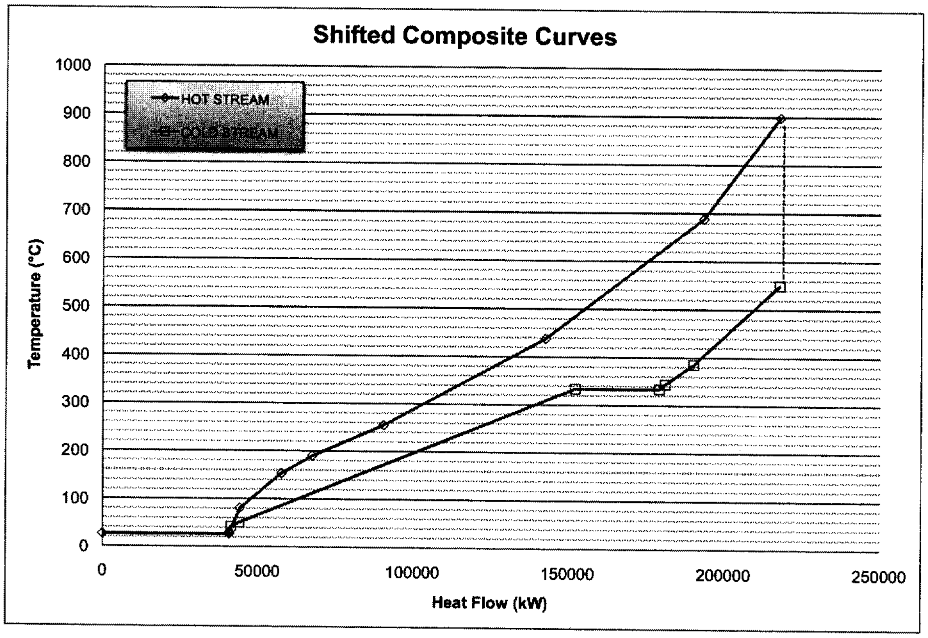

Figure 6.6 – Baseline Pinch Analysis Composite Curve

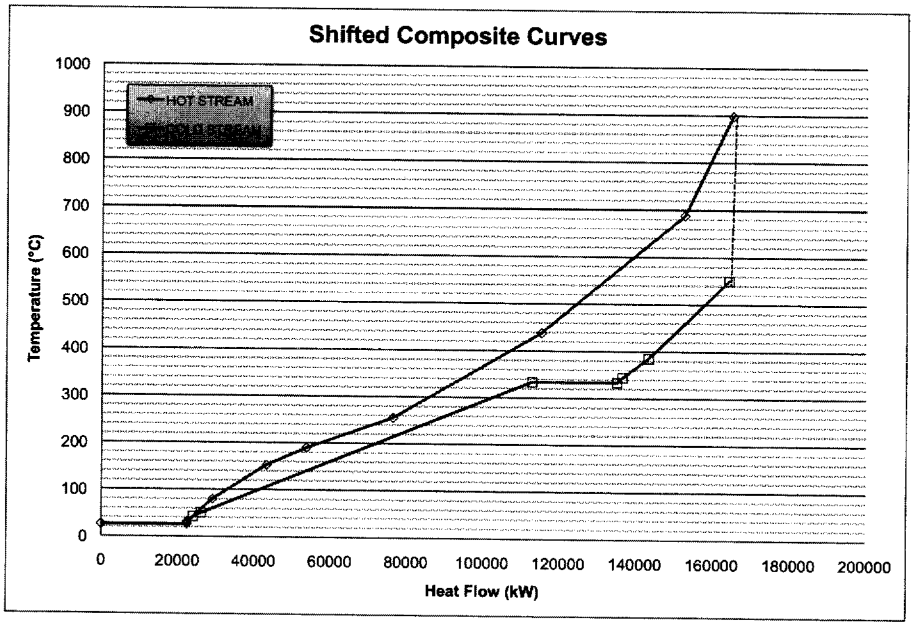

Figure 6.7 – Hybrid Pinch Analysis Composite Curve

6.2 Modeling

In order to estimate the input and output values for each plant, an ASPEN Plus model was developed based on Figures 6.4 and 6.5 respectively. The ASPEN Plus files for both models can be found on the included CD-ROM.

As previously mentioned, the electrolyzer bank for the hybrid system consists of 160 electrolyzers. Input and output parameters from an electrolysis plant of that size were determined from the published "Future Central Hydrogen Production from Grid Electrolysis" H2A analysis. Based on ASPEN Plus simulations and the electrolysis H2A case, the plant input and output values for the hybrid system were estimated. The values for each individual plant and the hybrid case are given in Table 6.3.

Table 6.3 – Direct Gasifier Hybrid Operating Requirements and Costs

	Units	*Baseline Biomass*	*Baseline Electrolysis*	*Hybrid System*
Capacity Factor	*%*	90	97	90
Biomass In	*kg/kg H_2*	15.3	0	5.6
Electricity In	*kWh/kg H_2*	0	44.7	25.4
Cooling Water	*gal/kg H_2*	170.6	293.9	237.0
Process Water	*gal/kg H_2*	5.3	2.9	4.0
Total Variable Operating Costs	*MM$/year*	$34.8	$117.8	$159.9
Electricity Out	*kWh/kg H_2*	3.1	0	0
Hydrogen Out	*kg/day*	118,344	167,360	322,440

6.3 Capital Costs

Multiple sources were used to estimate capital costs. All electrolyzer costs were calculated using the previously mentioned future central hydrogen production from electrolysis case. The majority of costs associated with the fluidized bed gasifier and feed preparation were taken from a recent publication in Biofuels Bioproducts and Biorefining (Jin et al, 2009). Gas cleanup costs were scaled based on previous systems studies completed at NREL (Spath et al. 2005). Selexol prices were based on (Department of Energy 2007). Finally, steam cycle and cooling costs were determined from the Pinch Analysis. An overview of the costs can be found in Table 6.4. Detailed costing information for the baseline fluidized-bed gasifier plant and the hybrid plant can be found in Appendices E and F.

Table 6.4 - Direct Gasifier Capital Costs

Plant Area	*Baseline*	*Hybrid*
Feed Preparation, Handling	$ 27,897,950	$ 27,897,950
Gasification, Tar Reforming, Quench	$ 22,723,289	$ 22,723,289
ASU or Electrolyzer Bank	$ 21,339,385	$99,162,176
Gas Cleanup	$ 29,906,771	$ 58,701,843
Shift and PSA	$ 18,626,072	$ 18,626,072
Steam System and Power Generation	$ 20,423,378	$ 20,423,378
Cooling Water and Other Utilities	$ 2,113,753	$ 3,713,021
Buildings and Structures	$ 6,368,900	$ 6,368,900
Total	**$ 149,399,497**	**$ 257,616,627**

Based on the initial cost information summarized in Table 6.4, direct replacement of the ASU with an electrolyzer bank causes a 72% increase in the total capital investment to build the gasification plant. While there are additional revenues created due to hydrogen produced by the electrolyzer bank, it is unlikely that the revenues will offset the dramatic increase in capital costs shown in Table 6.4.

To determine if electrolysis might become the economically preferable option over cryogenic air separation at a different scale, cost curves for both technologies were constructed. Figure 6.8 shows the combined capital and operating costs of each system per tonne of oxygen produced. The ASU cost curve was constructed by fitting data from multiple literature sources with a power function. Details of the fit can be found in Appendix G. Electrolyzer costs were linear with increasing scale (and therefore a flat line on the graph) because additional individual electrolyzer units were added. Using the economic assumptions in this study, there is not a point at which electrolysis will be favored from a cost standpoint. However, using electrolysis rather than an ASU for smaller plant scales appears to be favored as the cost per tonne of oxygen increases for ASU units at small scales.

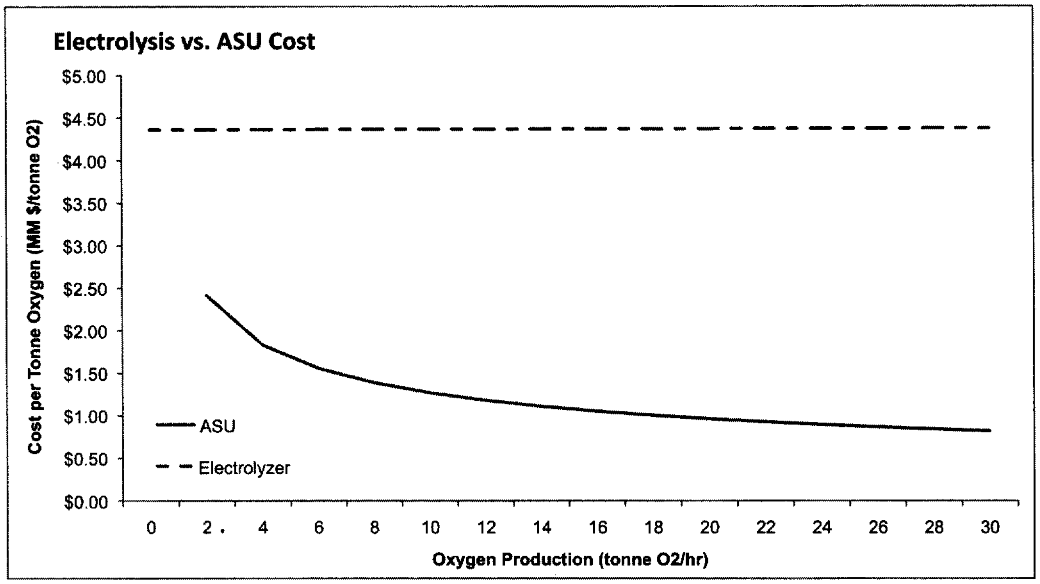

Figure 6.8 – Cost of Electrolysis vs. Cost of ASU

Below oxygen requirements of 50 ton/day (1.9 tonne/hr) cryogenic air separation becomes uneconomic. At these small scales other competitors enter the market like LIN-assist cryogenic plants and non-cryogenic air separation techniques (Universal Industrial Gases). More analysis on this smaller scale could provide different results. However, because these additional technologies are economically favored at small scale to cryogenic air separation it is unlikely that electrolysis will be cost competitive if the ASU trend holds.

CHAPTER 7

ECONOMIC MODEL

The hybrid system concepts studied herein respond dynamically to fluctuations in the energy market, either absorbing or providing electricity on demand. To simulate this switching, a binary model was created for each proposed system based on a specified peaking or sinking duty. Duty is defined as the percent of hours per year where either sinking or peaking mode is used. The models were created in excel and switch between a baseline mode and one of the hybrid modes discussed in Chapters 5 and 6.

Leveling of wind-generated electricity while producing hydrogen fuel is the goal of the concepts studied. However, the cost of electricity was used to determine the hybrid duty cycle instead of wind availability. This choice was made so that the cost of electricity could be easily quantified. All concepts were assumed to be grid connected for the analysis. For the directly-heated gasifier hybrid system the effect of stranded operation is discussed briefly in the results.

Regional Transmission Organization (RTO) day-ahead prices were used for the cost-of-electricity when available. These prices represent the market value of electricity to the utilities in an area on an hourly, averaged basis. The hybrid systems must be profitable at this low price point to be able to trade electricity on the market. When day-ahead market information was unavailable, load lambda data was used in its place. Load lambda data gives the cost of producing one unit of electricity to the utility for each hour of the year.

Based on GIS research reported earlier (see Figure 2.5 and Figure 2.6), three areas appear to have promising quantities of both wind and biomass. These areas are the Northeast, Midwest, and Northwest. Cost-of-electricity data for each location is shown in Figure 7.1.

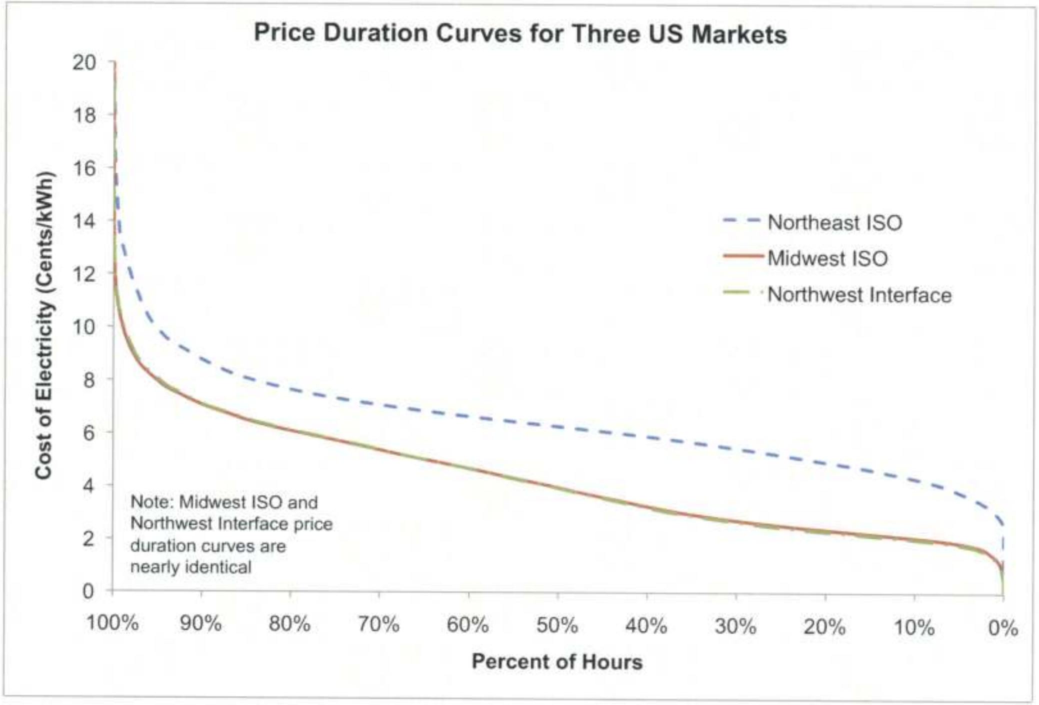

Figure 7.1 - Regional Price Duration Curves

Each profile shown above was based on 2007 year-end data. Costs for the Northeast were based on NE ISO day-ahead data. Midwest ISO data was directly available from the regional office itself. To estimate the northwest, where no RTO currently exists, load lambda data was used from the Northwest Interface. More detailed information on the grid data can be found in Appendix H.

At current levels, wind penetration does not dramatically alter the cost-of-electricity on the regional market. In the near future, wind will continue to be balanced by additional natural gas turbines to provide the majority of peaking electricity. Based on the most recent Energy Information

Administration (EIA) Annual Energy Outlook (AEO) report (EIA 2009), electricity prices are expected stabilize at current levels and then remain unchanged for several years. Looking further ahead, the AEO predicts the real cost of electricity to increase only 1.4 – 1.8 cents per kilowatt-hour between 2015 and 2030 in constant dollars (EIA 2009). This level of price increase is addressed in the sensitivity analysis found at the end of the report.

In June of 2009 the House of Representatives passed the American Clean Energy and Security Act of 2009 (ACESA). ACESA, is an extremely complex bill that attempts to regulate greenhouse gases with a combination of markets, efficiency programs, and incentives. Though the bill has the potential to change the rate of renewable energy deployment, it is not expected to significantly affect the AEO electricity projections (EIA 2009).

Modern grid power can be roughly divided into base load and peaking electricity. Base load power is produced by "always on" generation facilities like coal and nuclear plants. Base load power is currently the lowest cost electricity available. Peaking electricity is provided for the most part by natural gas turbines, which can be turned off and on quickly. However, other more expensive sources of electricity such as wind and solar also compete in this higher-price region.

For proper operation of the hybrid systems studied, the distribution of peaking or sinking needs versus base load demand must be heterogeneous. Figure 7.2 shows the time distribution of costs over the course of the year for the Northeast. Because the red high priced electricity and dark blue low cost electricity occur for only a few hours at a time, and appear relatively predictable, the switching proposed should be reasonable from a system control standpoint.

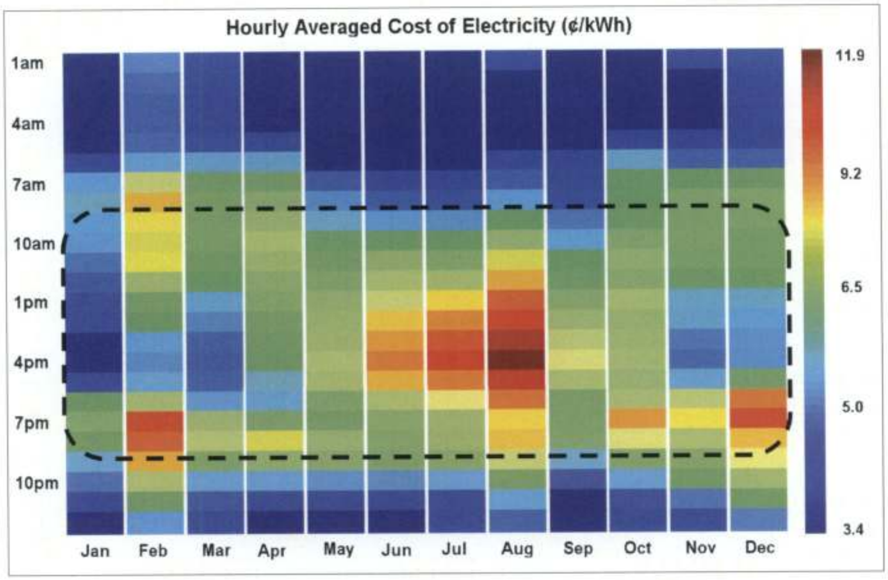

Figure 7.2 - Time Diagram of Electricity Cost

To assess the market potential of each of the proposed hybridizations, the yearly inputs and outputs for each plant were entered into the H2A Analysis Tool[8] along with applicable capital costs. The resulting cost of hydrogen produced, in \$/kg, was compared to both the published Future Central Hydrogen Production via Biomass Gasification H2A results and also the Future Central Hydrogen Production from Natural Gas without CO_2 Sequestration H2A results. If the cost of hydrogen produced by a hybrid system is lower than the Future Central Biomass to Hydrogen case then the cost of additional equipment (incremental costs) for hybridization were fully offset by added income or efficiency that resulted from the hybridization.

The H2A analysis, as it was run, makes several assumptions including that current electricity prices are representative of those at the time of actual plant construction and that sufficient market demand for hydrogen exists that all product can be sold. The major economic assumptions are

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[8] Additional information available at: http://www.hydrogen.energy.gov/h2a_analysis.html

summarized in Table 7.1. All additional assumptions for maintenance, land and financial variables were taken directly from the existing, published Future Biomass to Hydrogen or Future Central Electrolysis H2A cases. In addition, no price is associated with carbon emissions or carbon avoided and no value is added inherently due to the dual-mode hybrid operation.

Table 7.1 - Economic Assumptions

Parameter	*Value*
Internal Rate of Return (after-tax)	10%
Debt/Equity	0%/100%
Plant Life	40 years
Depreciation	MACRS
Depreciation Recovery Period	20 years
Construction Period	2 years
1st year	75% (25% for electrolysis)
2nd year	25% (75% for electrolysis)
Start-up Time	12 months
Revenues	50%
Variable Costs	75%
Fixed Costs	100%
Working Capital	15% of Total Capital Investment
Inflation Rate	1.9%
Total Taxes	38.9%
Decommissioning Costs	10% of depreciable capital
Salvage Value	10% of total capital investment

The effects of carbon costs on system results were taken into account separately as an adjustment to the H2A results. The amount of CO_2 equivalent (CO_2e) emissions per kilogram of hydrogen produced was tracked for all cases analyzed. The emissions for each hybrid system vary not only with the type of hybridization but also with the amount of time spent in each mode of operation. Values for these emissions are given with the detailed results.

In addition to the CO_2e emissions from the plant, there was assumed to be value associated with carbon emissions avoided due to the renewable nature of any fuel or electricity production. Based on the regional average grid mix, anywhere from 483 to 724 kg of CO_2e are emitted per kWh of electricity produced. Table 7.2 details the grid production mix and corresponding emissions for each location studied. One kilogram of hydrogen has the approximate energy equivalent of one gallon of gasoline. However, because hydrogen can be used in fuel cells with much higher efficiency, one kilogram of hydrogen could actually offset about two gallons of gasoline.[9] Burning two gallons of gasoline produces 17.84 kg CO_2e. These numbers were used as carbon credits for each kilogram of hydrogen or MWh of electricity produced.

Table 7.2 - Electricity Greenhouse Emissions by Region

	Kg CO_2e per kWh[1]	*NE ISO*[2]	*MISO*[3]	*NW Interface*[4]
Coal	952.5	15.0%	52.2%	58.0%
Oil/Petroleum	893.1	18.0%	9.3%	1.0%
Natural Gas	599.2	30.0%	23.9%	19.7%
Nuclear	-	28.0%	7.9%	1.0%
Renewable	-	9.0%	6.7%	20.3%
Average kg CO_2e/kWh		483.4	723.5	679.4

[1] Based on HyARC Energy Constants and Assumptions that are part of the H2A Analysis Tool.
[2] Based on market reports available at http://www.iso-ne.com/, Accessed August 10, 2009.
[3] Based on market reports available at http://www.midwestiso.org/, Accessed August 10, 2009.
[4] Approximation based on http://www.pacificorp.com/File/File89760.pdf, Accessed August 10, 2009.

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[9] Based on technology projections to 2020, gasoline vehicles are expected to use approximately 1.94 MJ/km traveled as opposed to 1.00 MJ/km traveled for a hydrogen fuel cell vehicle (Shäfer et al. 2006).

CHAPTER 8
ECONOMIC RESULTS

Neither the indirect gasification based hybrid system nor the direct gasification based hybrid system produced hydrogen for less than a non-hybrid plant could. In all cases a premium was paid for hybridization that could not be offset by the increased functionality. The results for each system and the non-hybrid baseline cases are detailed below.

8.1 Indirect Hybrid System – Peaking

The proposed system switches between (1) hydrogen production and (2) electricity production (peaking) driven by the cost of electricity available on the grid. Based on discussions with Xcel Energy (Frank Novachek, August 2009) a peaking duty of 20% was used for the analysis. Table 8.1 summarizes the major model inputs by region for the indirect hybrid system. A contract rate was used for any peaking electricity produced by the plant. This is common practice in the current electricity market and provides a premium price for dispatchable peaking assets.

Based on the economic inputs above, and the plant inputs and outputs previously discussed, the cost of hydrogen production in each area was calculated. The results are shown in Table 8.2 along with the cost of hydrogen production for a baseline, non-hybridized biomass to hydrogen plant and a steam methane reforming plant.

Regardless of whether gas turbine or a swinging steam cycle is used, the additional capital costs of hybridization cannot be justified today in any of the locations studied. There is a 17-24 cent premium on hydrogen produced by the proposed hybrid system compared with a non-hybrid biomass to hydrogen gasification plant. Areas with higher priced electricity move this hybridization closer to economic feasibility with the best results found in the Northeast.

Table 8.1 - Indirect Hybrid System - Peaking H2A Inputs

	NE ISO	*MISO*	*NW Interface*
Peaking Duty	20%	20%	20%
Peaking Electricity Value (¢/kWh)	14.0	12.0	11.5
Utility Electricity Cost (¢/kWh)	5.92	3.62	3.69
Utility Natural Gas Cost ($/nm^3)	0.32	0.32	0.32
Cost of Biomass ($/ton)[1]	48.83	48.83	48.83

[1] Value taken from the Biomass 2009 Multi-Year Research, Development and Demonstration Plan (EERE 2009). The 2012 target value is $50.70 per ton of dry woody biomass in 2007 dollars. Taken to 2005 dollars with 1.09% inflation this yields $48.83 per ton.

Table 8.2 - Indirect Hybrid System – Peaking Cost of Hydrogen Results

$/kg H_2	*NE ISO*	*MISO*	*NW Interface*
SMR	1.40	1.40	1.40
Biomass to Hydrogen Baseline	1.64	1.64	1.64
Gas Turbine Hybrid System	1.84	1.86	1.88
ICC Hybrid System	1.81	1.85	1.87

The previous results assumed that carbon has no value. As carbon emissions are controlled, carbon costs could be a significant factor. If carbon costs are taken into account, then the hybrid system results substantially improve. When the hybrid plant is producing electricity, the plant uses no natural gas or electricity whereas when the plant is producing hydrogen, natural gas is used for balancing the heat duty of the plant and electricity is required to run compressors.

Taking the differences in carbon emissions into account, a value of \$34-40 per tonne of CO_2e makes the cost of hydrogen for the proposed hybrid system equal that of SMR depending on the location and power output of the system. At \$37 per tonne of CO_2e the simple-cycle hybridization becomes cost competitive with a methane steam reforming plant in the Northeast. At \$34 per tonne of CO_2e the combined-cycle system becomes cost competitive with SMR in the Northeast. The effect of carbon costs on the cost of hydrogen are shown with the sensitivity results in Figures 8.1 and 8.2.

It is important to note that the baseline Biomass to Hydrogen plant only requires approximately \$23-25 per tonne of CO_2e value to be cost competitive with hydrogen produced by SMR. This means that peaking hybridization will only be economically promising when there is some value placed on the additional functionality of dual mode operation.

In order to characterize the effect of various technical and economic assumptions, a sensitivity analysis was performed for both the gas turbine and integrated combined-cycle peaking systems. Figures 8.1 and 8.2 show the results for both the price of hydrogen and greenhouse gas emissions.

The sensitivity analysis for the gas turbine system (Figure 8.1) showed that capital costs, the cost of biomass, and the price which peaking electricity can be sold for are key inputs. Capital costs, when varied +/- 30-percent cause the cost of hydrogen to vary 25-cents. The cost of biomass was varied from \$40 to \$60 per ton based on projections in the Biomass 2009 Multi-Year Research, Development and Demonstration Plan and causes the cost of hydrogen to vary by less than 20-cents. The price of peaking electricity and the plant capacity factor also cause fluctuations of less than 20-cents. Changes in the cost of electricity bought by the plant, the cost of natural gas, peaking power output, and turbine peaking duty cause hydrogen costs to vary less than 5-percent.

Greenhouse gas emissions are affected by the gas turbine peaking duty. During hydrogen production, a natural gas trim is used to maintain the tar cracker catalyst regenerator temperature. When producing peaking electricity, the natural gas trim is replaced with synthesis gas. This switch means that higher peaking duty results in less natural gas use and therefore fewer net emissions.

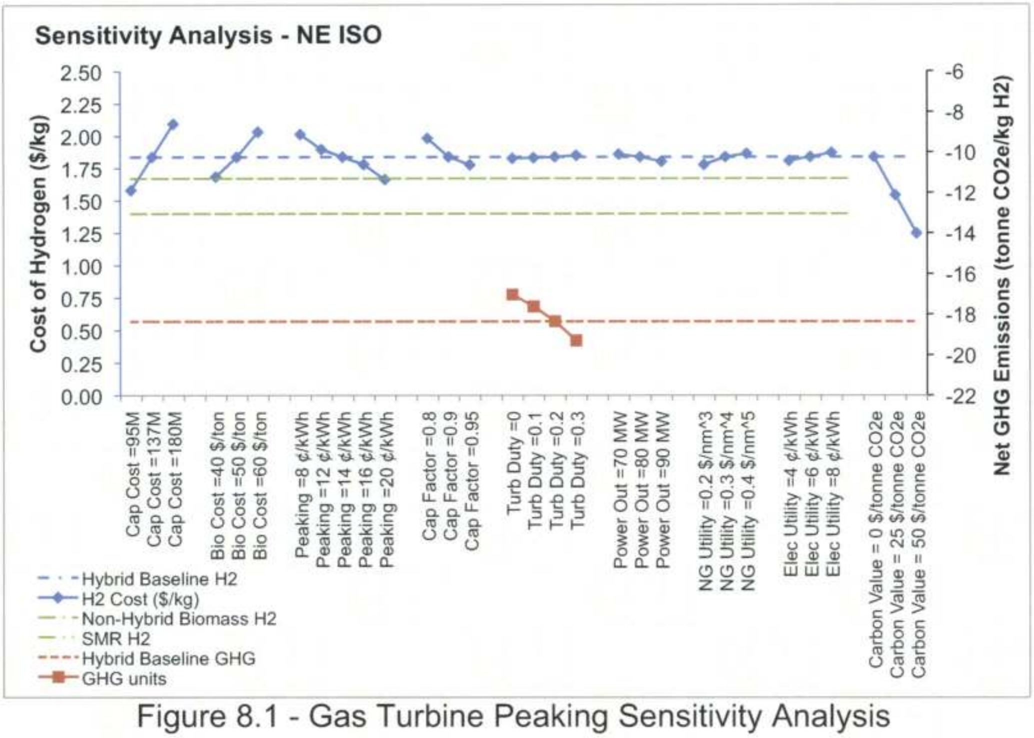

Figure 8.1 - Gas Turbine Peaking Sensitivity Analysis

For the integrated combined cycle peaking system (Figure 8.2) the results were significantly more volatile than the gas turbine system. Most of the trends discussed for the gas turbine system held but were amplified.

One difference was that the peaking duty had a much more significant effect on the cost of hydrogen, resulting in variations up to 22-cents. This shows that the cost of hydrogen produced is extremely sensitive to the amount of peaking power produced by the system.

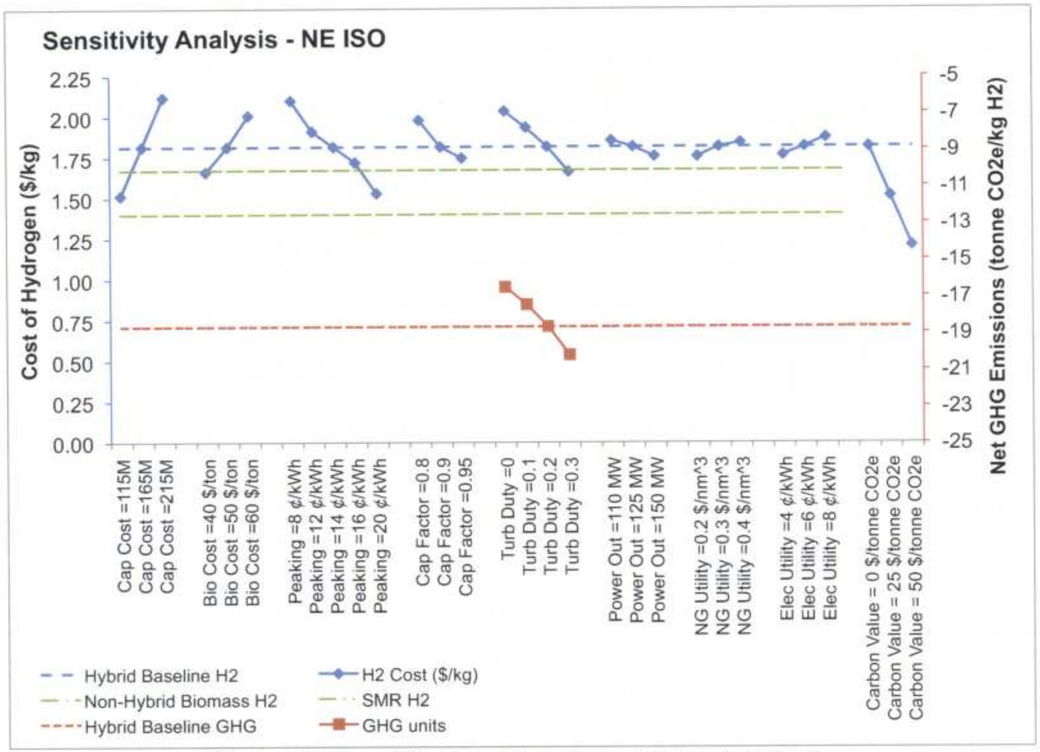

Figure 8.2 - ICC Peaking Sensitivity Analysis

8.2 Indirect Hybrid System – Sinking

The proposed system switches between (1) hydrogen production with syngas recycling for gasifier heat and (2) hydrogen production with electrical heating (sinking) to decrease or fully replace the syngas recycle. This sinking ability can best be described as a dispatchable load or demand from the viewpoint of the grid. Based on discussions with Xcel Energy (Frank Novachek, August 2009), the utility will pay to have wind-generated electricity used at times (the electricity has a negative cost) and having a dispatchable load would provide a valuable service to the utility. Xcel Energy is required by Colorado's renewable portfolio standard to accept renewable energy when it is available. The requirement to accept wind electricity regardless of grid demand is not universal and therefore negative cost wind electricity may not be available in other states.

No similar system was found for comparison; therefore a sinking duty of 20% was used as a starting point for the analysis. Table 8.3 summarizes the major economic model inputs by region. Unlike peaking electricity, where a contract rate was used for electricity produced by the plant, the sinking analysis simply used the average cost-of-electricity for the cheapest 20% of hours as the sinking electricity cost. The utility electricity cost was the average of the remaining 80% of the hours. These costs-of-electricity would be valid if a plant operator used accurate day-ahead energy market forecasts to schedule plant operation.

Table 8.3 - Indirect Hybrid System - Sinking H2A Inputs

	NE ISO	*MISO*	*NW Interface*
Plant Capacity Factor	90%	90%	90%
Sinking Duty	20%	20%	20%
Sinking Electricity Cost (¢/kWh)	4.35	2.20	2.18
Utility Electricity Cost (¢/kWh)	7.05	4.87	4.69
Utility Natural Gas Cost ($/nm^3)	0.32	0.32	0.32
Cost of Biomass ($/ton)	48.83	48.83	48.83

Given the economic inputs in Table 8.3, and the plant inputs and outputs detailed in Section 5.2, the cost of hydrogen production in each area was calculated. The results are shown in Table 8.4 along with the cost of hydrogen production for a baseline, non-hybridized biomass to hydrogen plant and a steam methane reforming plant.

Table 8.4 - Indirect Hybrid System – Sinking Cost of Hydrogen Results

\$/kg H_2	*NE ISO*	*MISO*	*NW Interface*
SMR	1.40	1.40	1.40
Biomass to Hydrogen Baseline	1.64	1.64	1.64
Electric Air Heater Hybrid System	1.75	1.67	1.68
All Electric Heat Hybrid System	1.77	1.68	1.67

There is a 3-13 cent premium on hydrogen produced by the proposed hybrid system compared with a non-hybrid biomass to hydrogen gasification plant. The additional capital costs of the sinking hybridization are not fully offset by additional revenue in any of the locations studied. However, the marginal costs found are small enough that it is difficult to draw any definitive conclusion. Areas with lower cost electricity move this hybridization closer to economic feasibility with the best results found in the Northwest.

Because the premium is small (about 5%) it may be acceptable in the long term. Recent studies have shown that there is inherent value added, or welfare effects, for electricity storage capacity. Whether similar value is added by the proposed sinking hybrid is unknown. A more likely parallel would be the idea of “interruptible customers” which get discounted electricity rates in return for intermittent power supply. A similar contractual agreement could be envisioned for the proposed hybrid where discounted electricity rates would be provided in return for intermittent demand.

The proposed system is not a direct competitor with storage systems such as pumped hydro or compressed air energy storage (CAES). Energy storage systems attempt to profit by market arbitrage (selling electricity back to the grid at a higher price than it was bought) whereas the proposed hybrid system sinks cheap electricity into transportation fuel. Sinking low cost electricity into fuel could be considered cross-market arbitrage and the most similar system to this would be electrolysis. Compared to electrolysis, the proposed system is significantly less expensive and has the added benefit of running without electric heat when electricity costs are too high.

Assuming that all sinking electricity is renewable and taking the additional differences in carbon emissions into account, a value of $26-35 per tonne of CO_2e makes the proposed hybrid cost competitive depending on the location and amount of dispatchable demand of the system. At $27 per tonne of CO_2e the combustion air-heater hybridization becomes cost competitive with a methane steam reforming plant in the Northwest. At $26 per tonne of CO_2e the all-electric heat system becomes cost competitive with SMR in the Northwest. The effect of carbon cost on the cost of hydrogen produced is shown along with sensitivity results on Figures 8.3 and 8.4.

The baseline Biomass to Hydrogen plant only requires approximately $23-25 per tonne of CO_2e value to be cost competitive with hydrogen produced by SMR. Thus, the sinking hybridization will only be economically promising when there is some value placed on the additional functionality of dual mode operation.

In order to characterize the effect of various technical and economic assumptions, a sensitivity analysis was performed for both the sinking systems. Figures 8.3 and 8.4 show the results for both the price of hydrogen and greenhouse gas emissions.

The sensitivity analysis for both heating systems (Figures 8.3 and 8.4) showed that capital costs, the cost of biomass, and the plant capacity factor for are key inputs. Capital costs, when varied +/- 30-percent cause the cost of hydrogen to vary 18-cents. The cost of biomass was varied from $40 per ton to $60 per ton based on projections in the Biomass 2009 Multi-Year Research, Development and Demonstration Plan and causes the cost of hydrogen to vary by less than 15-cents.

Changes in the sinking duty, the added hydrogen production due to sinking, the cost of natural gas and the cost of electricity cause hydrogen costs to vary less than 3-percent in both cases. There were no major differences in sensitivity between the two cases.

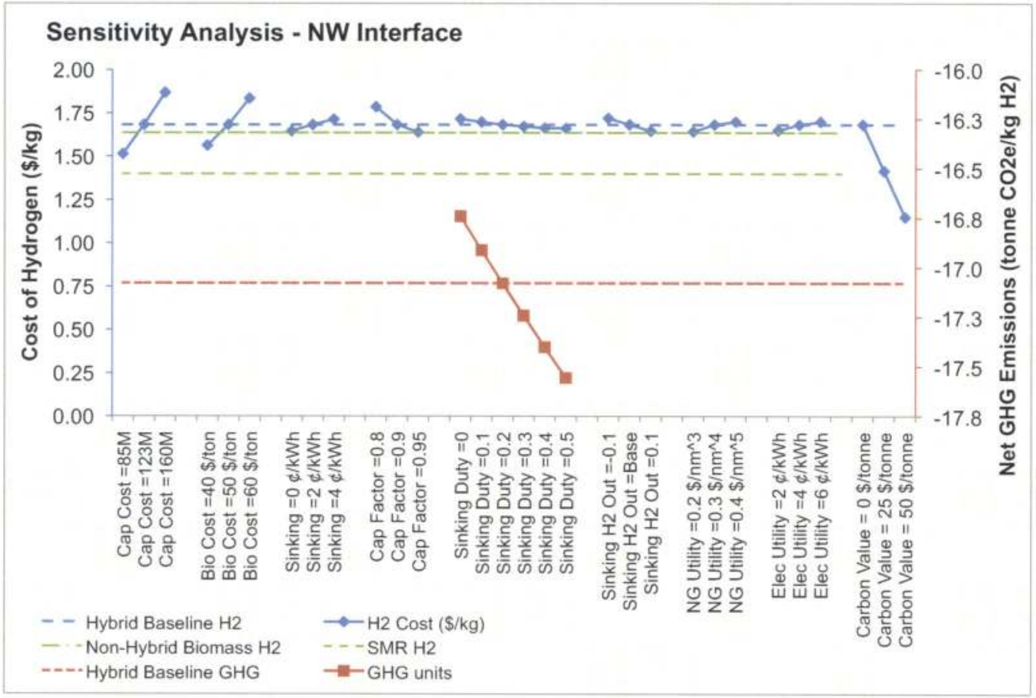

Figure 8.3 – Air Heater Sinking Sensitivity Analysis

From the sinking duty sensitivity analysis, it appears that an increased sinking duty would be preferable to the 20% assumption made. Minimum hydrogen production prices occur for both systems when the electric heating systems are run approximately 40% of the time. Running the electric heaters for more time increases plant hydrogen production and decreases plant GHG emissions.

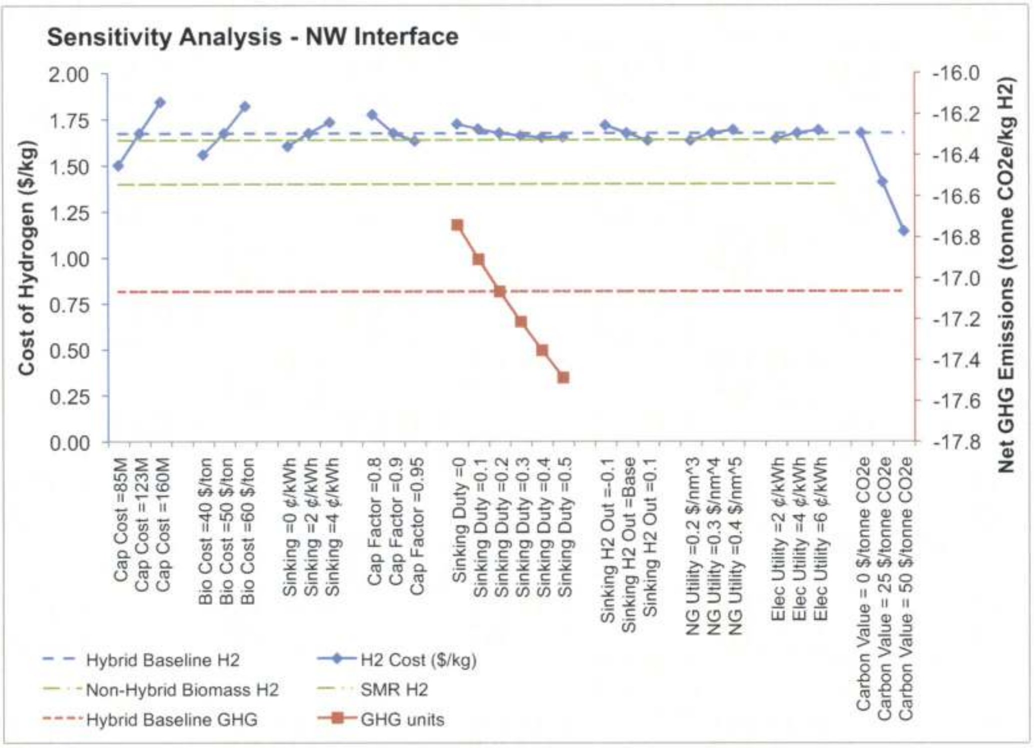

Figure 8.4 – All Electric Sinking Sensitivity Analysis

8.3 Direct Gasifier Hybrid System

Based on the model inputs and outputs described in Section 6.2, an H2A analysis was completed to determine the baseline cost of hydrogen produced by a directly heated gasification plant that uses electrolysis regardless of electricity cost and the cost of hydrogen for a directly heated gasification plant that uses a traditional ASU. Table 8.5 summarizes the major model inputs. The plant was assumed to be located in the Midwest ISO region.

Table 8.5 - Direct Hybrid System - H2A Inputs

	MISO
Electrolyzer Duty (%)	100
Utility Electricity Cost (¢/kWh)	4.33
Cost of Biomass ($/ton)	48.83

The results of the analysis, along with the associated carbon emissions, are listed in Table 8.6. Costs for production of hydrogen via SMR and for an electrolysis plant only are also shown for reference purposes.

Table 8.6 – Cost of Hydrogen & CO_2e Emissions from Direct Gasification

	$/kg H_2	*Net $CO_2e/kg\ H_2$*
SMR	$1.40	-6.8
Biomass Gasification	$2.13	-19.6
Electrolysis	$2.59	14.5 (-17.8)
Electrolysis + Gasification	$2.32	0.24 (-18.2)

At $2.32 per kilogram, hydrogen produced by a direct gasifier-electrolyzer hybrid plant is significantly more expensive than that produced by SMR. Electricity costs account for 36.9% of the overall cost of hydrogen produced (or $0.86). Therefore, if electricity costs could be halved by intermittent operation, the savings would bring the cost of hydrogen to $1.89 per kilogram of hydrogen. At this price, the hybrid system could compete with hydrogen produced by a standard direct gasification plant. It's important to note that these costs are approximately 15% higher than the $1.64 per kilogram of hydrogen estimated for the indirectly-heated baseline gasification plant used for comparison of the indirect hybrid analysis (see Table 8.4).

Unfortunately, to halve the cost of electricity used by the plant, the electrolyzer bank could be run only at the bottom of the area's price duration curve. The previous sinking analysis assumed that extra electricity was used in only the cheapest 20% of hours in a year and that resulted in half-price sinking electricity.

The original concept called for the electrolyzer bank to be directly coupled with stranded wind resources. Typical capacity factors for wind-generated power are in the 30-35% range (Kroposki et al. 2006). The effect of intermittent operation on the cost of hydrogen production was investigated using electrolyzer duties of 20, 35, and 50%. The smaller the duty, the larger the number electrolyzers needed to produce the same amount of oxygen per year.

The effect of reduced duty on the cost of hydrogen is shown in Figure 8.5. Two curves are plotted. The first assumes a constant electricity price of \$4.33/MWh regardless of the electrolyzer duty. This assumption would represent the scenario if the electrolyzers were directly coupled with a stranded wind resource. The second curve uses Midwest ISO market data so that the cost of electricity fluctuates with duty. Table 8.7 lists the data used. These numbers do not consider the cost of oxygen storage, as these costs are assumed small compared to the increased electrolyzer capital costs and are identical for similar duty.

Table 8.7 – Cost of Hydrogen vs. Electrolyzer Duty

Duty (%)	*Constant \$/MWh*	*\$/kg H_2*	*MISO \$/MWh*	*\$/kg H_2*
100	\$4.32	\$2.32	\$4.32	\$2.32
50	\$4.32	\$2.59	\$2.77	\$2.18
35	\$4.32	\$2.83	\$2.43	\$2.32
20	\$4.32	\$3.42	\$2.20	\$2.85

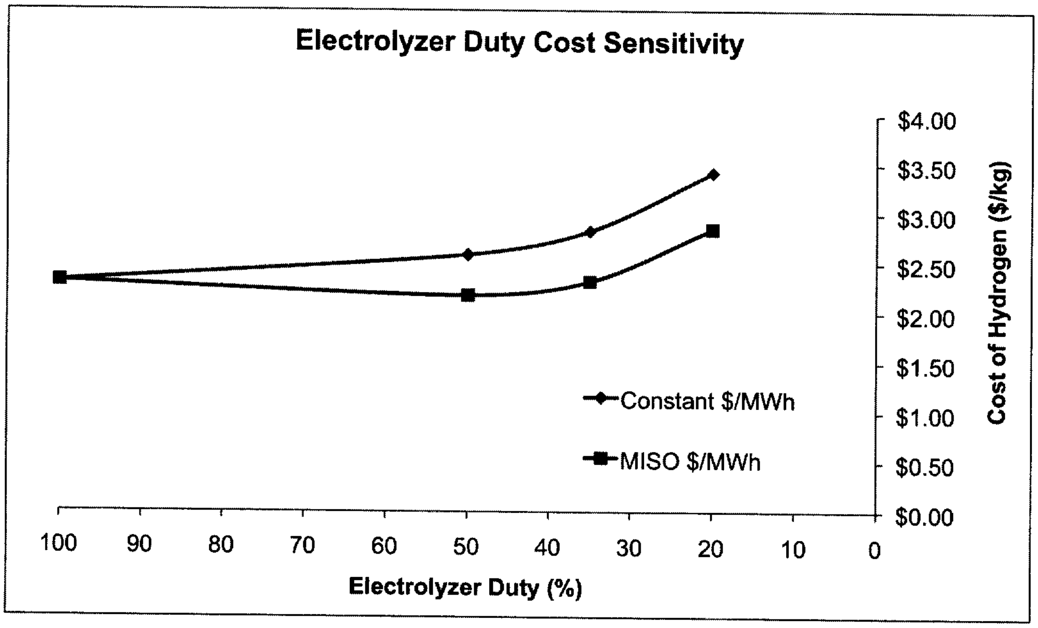

Figure 8.5 – Electrolyzer Duty vs. Cost of Hydrogen

Intermittent operation significantly adds to plant capital costs. If the price of electricity available to the plant does not decrease with intermittent electrolyzer operation, then the cost of hydrogen increases by 22% for a 35% duty. However, if the cost of electricity can be reduced by intermittent operation then the analysis shows that intermittent operation may be preferable to 100% duty. This result means that the initial concept of capturing stranded resources is less favorable economically than a grid connected sinking system.

If grid electricity is used then carbon emissions actually increase relative to biomass gasification alone for this hybridization. If upstream emissions for electricity production are taken into account, electrolysis and this proposed hybrid are both net CO_2e emitters. This makes their justification by carbon value impossible unless only renewable electricity is used for operation (renewable energy values are shown in parentheses in Table 8.6). Since renewable wind energy is inherently intermittent, it is unreasonable to assume renewable electricity is used without adding the necessary equipment for intermittent operation. One renewable energy exception would be the use of hydroelectric power for the systems.

Similar to the other hybrid systems investigated, hybridization results in a price premium. There must be some additional justification for hybridization such as welfare effects or utility demands. The baseline biomass to hydrogen case with an ASU requires only \$35 per tonne of CO_2e to be cost competitive with SMR produced hydrogen.

Figure 8.6 shows the results of a sensitivity analysis performed on the baseline biomass to hydrogen via direct gasification model. Similar to the previous gasification systems, capital-cost uncertainties have the largest effect on the price of hydrogen produced. The sell price of electricity, electricity production and water consumption have very little effect on overall economics of the plant.

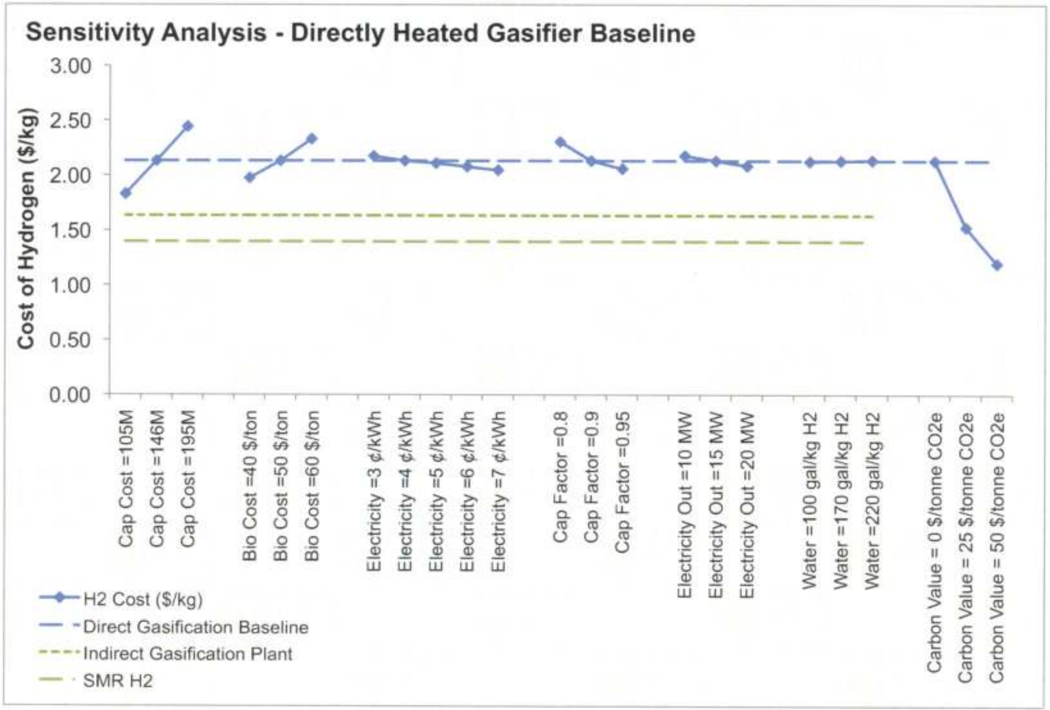

Figure 8.6 – Direct Gasifier Baseline Sensitivity Analysis

In contrast to the baseline plant, electrolyzer-based hybrid system economics are extremely sensitive to the cost of electricity (see Figure 8.7). Fluctuations of plus or minus one-cent in the cost of electricity cause the price of hydrogen to fluctuate more than 25-cents per kilogram. Lines are included for both electrolysis alone and the baseline biomass plant for comparison. If the price of electricity were to drop below 0.4-cents per kilowatt-hour then the proposed hybrid system could economically compete with a non-hybrid biomass to hydrogen plant. However, this low price level is unlikely and any energy market fluctuations would have a dramatic effect on economic viability.

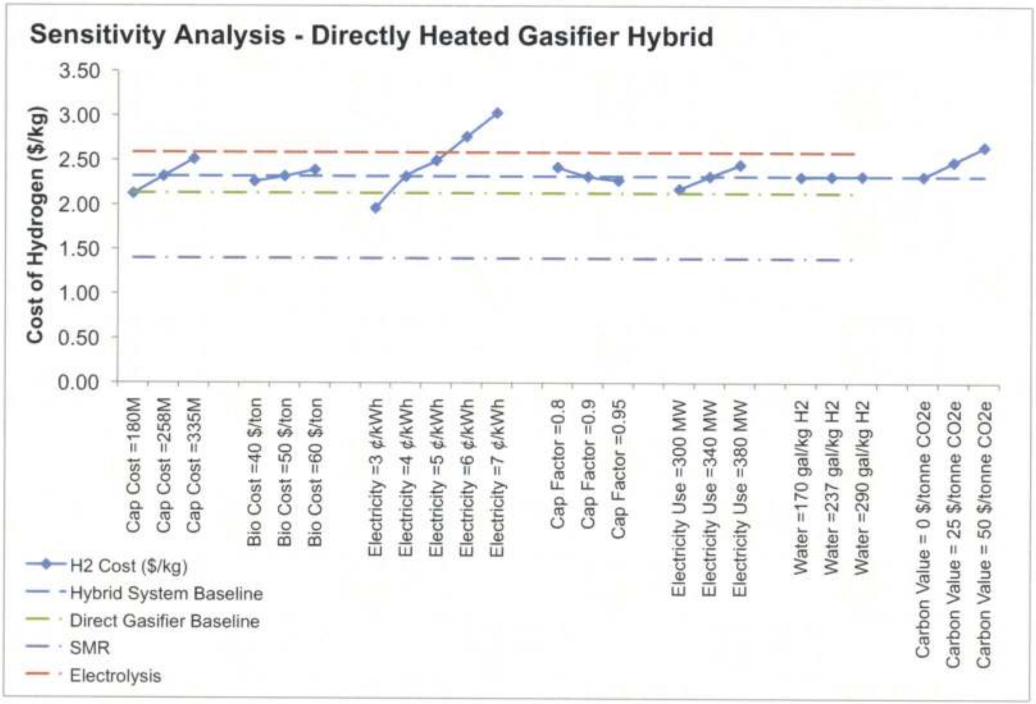

Figure 8.7 – Direct Gasifier Hybrid Sensitivity Analysis

Further intermittent simulation was not performed on the concept after these initial findings. For the use of electrolysis to make sense in this setting, there must be a significant benefit to the ability to operate intermittently. One envisioned possibility would be to use enriched air rather than pure oxygen for gasification. This would prevent the production of fuel but would be acceptable for power production and the electrolyzers could produce a slipstream of hydrogen fuel.

CHAPTER 9
CONCLUSIONS

This thesis evaluated biomass pathways for hydrogen production and how they can be integrated with renewable resources to improve the efficiency, dispatchability, and cost of other renewable technologies. Many potential hybrid systems were considered before narrowing the focus to two. The selected systems were studied for process feasibility and economic performance. The best performing system was estimated to produce hydrogen at costs as low as $1.67/kg while also providing value-added energy services to the electric grid.

Of the domestic resources available for hydrogen production, biomass shows significant promise. Recent assessments have shown that in excess of 400 million tons of biomass are currently available annually in the United States, which could be converted to roughly 30 million tons of hydrogen by thermochemical processing.[10] Thermochemical plants provide many opportunities for system integration.

A matrix considering the combination of biomass processing technologies and how they could be hybridized with other technologies was generated at the beginning of research. The matrix contained many potential binary technology combinations. These were ranked based on criteria such as resource availability, technology maturity, and hybridization benefits. Some of the top concepts included:

- Combined wind power and biomass gasification for co-production of fuel and power.

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[10] This conversion efficiency is based on values obtained from the various gasification systems described and analyzed as part of this research.

- Combined electrolysis and biomass gasification for co-production of fuel and power.
- Combined coal and biomass/bio-oil gasification systems for co-production of fuel and power with carbon sequestration for both processes.
- Co-location and thermal integration using steam from a nuclear reactor to feed bio-oil reforming to produce fuel.

These results were narrowed using a decision matrix. Direct wind and wind/electrolyzer combinations with biomass gasification rose to the top of the decision matrix due to several factors. The concepts provide renewable fuel and power, supplement grid demand and also can take up excess electricity. The two concepts chosen for detailed analysis can be summarized as:

- Direct grid leveling of intermittent wind power with an indirectly-heated biomass gasification plant. The plant will produce both electricity and hydrogen.
- Using an electrolyzer in place of an air separation unit (ASU) with a directly heated fluidized-bed biomass gasifier for co-production of fuel and power.

The two hybrid concepts analyzed involve co-production of gaseous hydrogen and electric power from thermochemical-based biorefineries. Both of the concepts analyzed share the basic idea of combining intermittent wind-generated electricity with a biomass gasification plant. Wind availability overlaps biomass resource availability in three areas of the U.S., making the use of locally produced wind electricity for gasification feasible. The proposed hybrid systems attempt to do one of two things:

1. Fill wind energy shortfalls by burning syngas in a natural gas turbine.

2. Absorb excess renewable power during low-demand hours.

The indirectly-heated gasifier hybrid system consists of two parts: (1) producing peaking electricity intermittently with a gas turbine and (2) sinking electricity into electric heaters intermittently to boost fuel production efficiency. Each of these parts was analyzed separately as independent options. The indirectly-heated gasifier hybrid system is meant to be grid connected and provide peaking electricity and dispachable demand (e.g. an on demand load) to the local utility the help manage intermittent wind resources.

The indirect gasification concepts studied could be cost competitive in the near future as value is placed on controlling carbon emissions. Carbon values of just under $40 per tonne of CO_2e make the systems studied cost competitive with steam methane reforming (SMR) to produce hydrogen. However, a non-hybrid biomass to hydrogen plant will be more cost competitive in general so there must be some additional value placed on peaking or sinking for these plants to be economically attractive. This additional value is likely to become a reality as additional intermittent renewable energy sources, such as wind, are added to the national grid.

For the proposed peaking system, increased electricity production dramatically improves the economics so the additional complexity of a swinging integrated combined cycle system may be justified. However, further work is needed to investigate the dynamic response and control of the proposed plants. Development of control strategies for these plants would be a significant undertaking and was not addressed by the steady state models used in this study.

The directly-heated gasifier hybrid system involves replacement of the air separation unit with electrolyzers. This change allows for extra production of hydrogen and intermittent operation. The analysis for this system assumed direct replacement of the ASU without intermittent operation. From there estimates were made of the effect of intermittent operation on plant economics. The electrolyzer bank could be either grid-connected to provide a dispachable demand to the utility or it could be tied directly to a stranded wind source to convert wind into fuel.

The direct gasification concept studied is unlikely to be cost competitive in the near future. Intermittent operation and stranded operation were found to have worse economics than grid connected systems. High electrolyzer costs make the possible electricity cost savings of intermittent operation difficult to justify. Based on a direct replacement of the ASU with electrolyzers, hydrogen can be produced for $2.32 per kilogram. However, using grid electricity, the hybrid system is a net CO_2e emitter. For the use of electrolysis to make sense in this setting, there must be a significant benefit to the ability to operate intermittently.

The study results and the baseline values used for comparison are summarized in the table below. A range of values is given for systems where multiple configurations and locations were studied. Cost numbers reflect current electric grid prices and demand. In the future, additional value may be placed on grid leveling services.

This analysis could be extended into other fuels (dimethyl ether, Fischer-Tropsch fuels), as well as to coal gasification. More detailed process modeling to cover dynamic response and controls systems would also solidify the analysis. The maps developed in the course of this study show that there are areas of the U.S. where wind and biomass resources overlap. It would be useful to develop more detailed GIS information to determine the actual amount of available wind energy resources that overlap with biomass.

Table 9.1 – Summary of Results

Cost of Hydrogen	*($/kg)*
Steam Methane Reforming Baseline	1.40
Indirectly Heated Biomass Gasification to Hydrogen Baseline	1.64
Directly Heated Biomass Gasification to Hydrogen Baseline	2.13
Electrolysis Baseline	2.59
Indirect Hybrid System for Peaking Electricity	1.81–1.88
Indirect Hybrid System for Sinking Electricity	1.67-1.77
Direct Hybrid Electrolysis System	2.32

As mentioned earlier in the paper, hybrid biomass systems encompass a wide range of possibilities. Significant opportunities exist for future work including:

- Determining if there is inherent value added in electricity sinking/peaking that is not directly considered by this analysis. Sioshansi et al. looked at the value of electricity storage in detail (Sioshansi et al. 2009) however; the market arbitrage they studied is significantly different then turning electricity into fuel. What is the value of "cross-market" arbitrage?
- Coal-biomass hybridization research received significant interest from multiple industry reviewers. Some research has been done on coal-biomass hybridization, as discussed in the literature review, but there are significant possibilities for future work including but not limited to:
    - A detailed techno-economic comparison of combined coal & biomass gasification systems versus coal alone and/or biomass alone.
    - Investigation of thermally-integrated coal power plant and biomass/bio-oil gasification systems.
    - A comparison of the economic and welfare effects of the various ways of biomass and coal mixing including

gasification, co-feed into steam boiler, bio-oil co-feed into steam boilers, separate combustion, etc.

- The study focused specifically on near-term hybridization possibilities. A similar literature review and brainstorming process involving cutting edge technologies could yield interesting results. Some ideas include:
    - thermochemical hydrogen production cycles from water.
    - oceanic biomass potential.
    - advanced ion transport membrane technologies in place of an ASU.
    - more advanced direct gasifier concepts involving experimental catalysts.
- This study assumed biomass gasification plants that were optimized for hydrogen production only producing electricity intermittently. Study of an inverse system in which a hydrogen slipstream is produced from a base load power generation plant would yield significantly different results.
- Price duration curves were needed for each region studied. Predicting the effect that increased renewables on the grid would have on these curves could not be done with existing data. A study to create an "ideal" price duration curve that could be used for similar studies would be extremely useful.
- Further refinement is needed on the ICC electricity production estimates and design since the ICC ASPEN model was not a detailed plant model.
- Determination of how the system would behave dynamically when switched between modes of operation. Would the switch between hydrogen production and power production or electric heating and syngas recycling cause overall system instability? If the system were stable, how long would switching take and what are the thermal cycling effects?

- Development of dynamic models and control strategies for each system. This is especially significant for the proposed integrated combined cycle peaking concept.

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APPENDIX A
NUMERICAL IDEA MATRIX

Analysis of Hybrid Hydrogen Production System Brainstorming

Matrix Description	**Initial decision matrix used to identify possible synergies between biomass processes and other existing and/or renewable processes. Zero represents no or limited correlations seen; 9 represents significant input/output correlation between the two technologies**		
Created	**7/24/08**	**Researcher**	**Jered Dean**
Updated	**11/22/08**	**Principle Investigator**	**Michael Penev**

INPUT VALUES FOR RANKING:	PAF	RPF	GHE	TRF	**Biomass Gasification**	**Biomass Pyrolysis**	**Bio-oil Reforming**	**Biomass Combustion**	**Biomass Fermentation**
PAF	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
PV Solar	[illegible]	[illegible]	[illegible]	[illegible]			**1**	**0**	**0**
Concentrated Solar	[illegible]	[illegible]	[illegible]	[illegible]					
Thermal Photovoltaic	[illegible]	[illegible]	[illegible]	[illegible]			**0**		**0**
Photo-electrochemical Solar	[illegible]	[illegible]	[illegible]	[illegible]	**1**	**1**	**0**	**0**	**0**
Photo-bio Electrolysis	[illegible]	[illegible]	[illegible]	[illegible]		**1**	**0**	**1**	
Wind Turbine (electrical)	[illegible]	[illegible]	[illegible]	[illegible]			**1**	**0**	**0**
Wind Turbine (mechanical)	[illegible]	[illegible]	[illegible]	[illegible]	**1**	**1**	**0**	**1**	
Thermal Integration w/ Nuclear	[illegible]	[illegible]	[illegible]	[illegible]		**1**		**0**	
Coal/Biomass Co-feed	[illegible]	[illegible]	[illegible]	[illegible]					**1**
Natural Gas Turbine	[illegible]	[illegible]	[illegible]	[illegible]					**1**
Steam Turbine	[illegible]	[illegible]	[illegible]	[illegible]					**0**
Thermal Integration w/ Coal	[illegible]	[illegible]	[illegible]	[illegible]				**0**	
Alkaline Fuel Cell	[illegible]	[illegible]	[illegible]	[illegible]	**0**	**0**	**0**	**0**	**0**
PEM Fuel Cell	[illegible]	[illegible]	[illegible]	[illegible]	**0**	**0**	**0**	**0**	**0**
Phosphoric Acid Fuel Cell	[illegible]	[illegible]	[illegible]	[illegible]	**0**	**0**	**0**	**0**	**0**
SOFC/MCFC	[illegible]	[illegible]	[illegible]	[illegible]				**0**	
Hydro-electric	[illegible]	[illegible]	[illegible]	[illegible]	**0**	**0**	**0**	**0**	**0**
PEM Electrolysis	[illegible]	[illegible]	[illegible]	[illegible]		**0**		**0**	**0**
Alkaline KOH Electrolysis	[illegible]	[illegible]	[illegible]	[illegible]		**0**		**0**	**0**
Hybrid SOFC/SOFC Electrolysis	[illegible]	[illegible]	[illegible]	[illegible]				**1**	**0**
Solid Oxide Electrolysis	[illegible]	[illegible]	[illegible]	[illegible]		**0**		**0**	**0**
Thermo-chemical cycles	[illegible]	[illegible]	[illegible]	[illegible]		**1**	**1**		**0**
Waste-water Treatment	[illegible]	[illegible]	[illegible]	[illegible]					

Notes: Weighting Factors are Potential Availability Factor (PAF) and Renewable Potential Factor (RPF) and Greenhouse Gas Emisions (GHE). The higher the PAF the more readily available plant and fuel are for the process. The RPF was divided into non-renewable (0), partially renewable or dependant on operation parameters (5) and 100% renewable (9). GHE was scored as no direct emissions (9), emissions that could be sequestered or reprocessed (7), and low emissions (5). Technology Readiness Factor (TRF)

APPENDIX B

INDIRECT GASIFIER ASPEN PLUS

Simulation Details

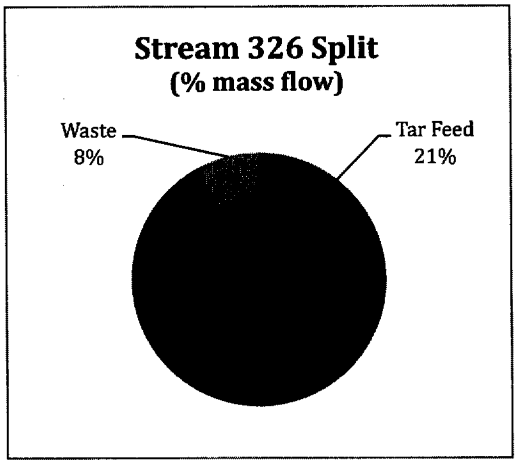
All ASPEN Plus simulations were modifications to the previous Wood to Hydrogen Using Indirectly-Heated Gasifier model available at http://devafdc.nrel.gov/biogeneral/Aspen_Models/. Specifically, the design for "Goal" tar reforming technology was used. Every attempt was made to maintain the base model thermal integration and minimize changes to the model. Details of the modifications made are given below.

GE 6FA Simple Cycle Gas Turbine

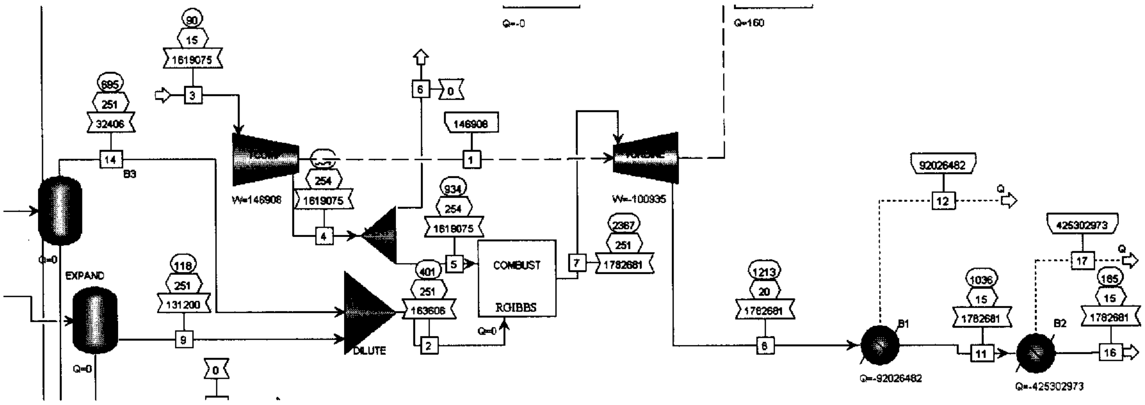
The following modifications were made to the existing Wood to Hydrogen ASPEN model:

1. The natural gas trim to the tar reformer catalyst regenerator (stream 427) flow was set to zero and the NGTRIM design spec was deactivated.
2. The PSA system and all downstream flows for hydrogen compression were set to zero and deactivated.
3. Stream 420 that would normal flow through the PSA system was instead routed to the tar reformer catalyst regenerator R-204.
4. The gas turbine model developed separately was inserted into the model.
5. Syngas stream 326 (directly after the LO-CAT reactor) was split to the water gas shift reactors and the turbine respectively. Any remaining syngas was considered waste. The following constraints were met with design specifications:

    a. TARFEED – adjusted the split fraction until the difference between the catalyst regenerator and the tar reformer was 200 °F.
    b. TBFEED – adjusted the split fraction until 6FA combustor temperature reached the specified 1288 °C.
6. A heat exchanger was added to the turbine exhaust so that lost heat to the plant steam turbine from H-405 and H-407 heat exchangers due to less flow through the WGS is exactly replaced by the turbine exhaust. This was controlled with calculator block "THEAT".



A screen capture of the gas turbine inputs and outputs can be seen below.



The final plant power specifications for this scenario are summarized in Table B.1.

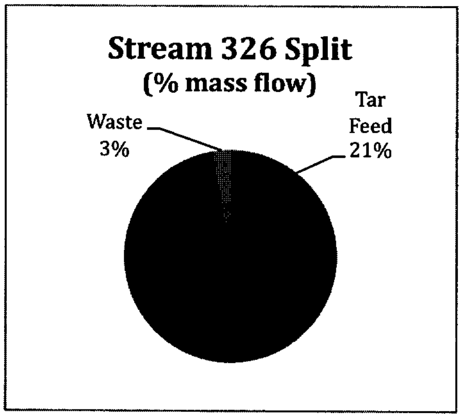
Table B.1– Plant Power Generation

	HP	*MW*
WPLANT	47442	35.4
WGEN	-43154	-32.2
WTURB	-100935	-75.3
WNET	**-96647**	**-72.1**

GE F-Class Simple Cycle Gas Turbine

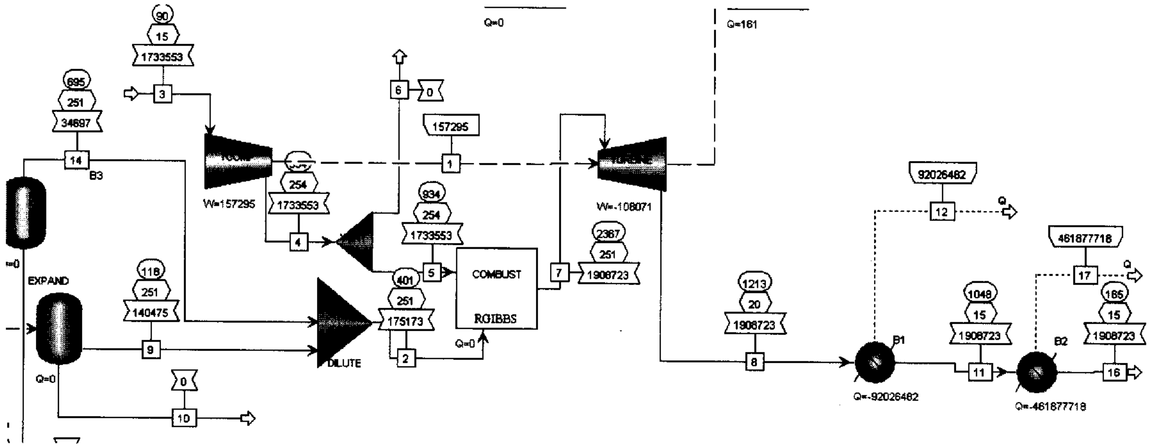
The following modifications were made to the case 6FA ASPEN model:

1. Syngas stream 326 (directly after the LO-CAT reactor) was split to the water gas shift reactors and the turbine respectively. The following constraints were met with design specifications:
    a. TARFEED – adjusted the split fraction until the difference between the catalyst regenerator and the tar reformer was 200 °F. Stated another way stream 107 was maintained at approximately 1791 °F
    b. TBFEED – All but available syngas remaining after the TARFEED design specification was met was sent to the gas turbine. Additional calculators were added so that the Air and Steam to Fuel ratios remained constant with the excess fuel flow.



Please note that 5000 lb/hr of the syngas stream was still sent to waste in this case. The Biomass to Hydrogen model was recently updated to include new yield correlations for the gasifier based on data collected in the Thermochemical Process Development Unit (Kinchin and Bain 2009). The updated model requires raw syngas to be diverted and combusted to supplement the heat delivered to the gasifier by the char combustor. The 5000 lb/hr waste stream effectively adjusts the energy flow in the current model to align with this soon to be released study.

A screen capture of the gas turbine inputs and outputs can be seen below.



The final plant power specifications for this scenario are summarized in Table B.2.

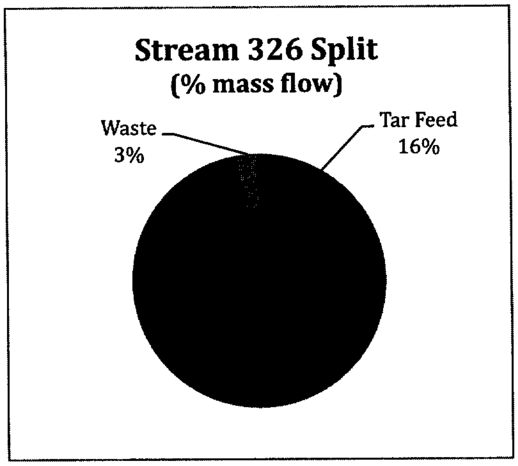
Table B.2 – Plant Power Generation

	HP	*MW*
WPLANT	47435	35.4
WGEN	-43163	-32.2
WTURB	-108071	-80.6
WNET	**-103799**	**-77.4**

Water Gas Shift Shutdown

The following modifications were made to the case F-Class ASPEN model:

2. Syngas stream 326 (directly after the LO-CAT reactor) was split to the tar reformer catalyst regenerator and the turbine respectively. The following constraints were met with design specifications:
    a. TARFEED – adjusted the split fraction until the difference between the catalyst regenerator and the tar reformer was 200 °F.
    b. TBFEED – All but available syngas remaining after the TARFEED design specification was met was sent to the gas turbine. Additional calculators were added so that the Air and Steam to Fuel ratios remained constant with the excess fuel flow.



Please note that 5000 lb/hr of the syngas stream was still sent to waste in this case. The Biomass to Hydrogen model was recently updated to include new yield correlations for the gasifier based on data collected in the Thermochemical Process Development Unit (Kinchin and Bain 2009). The updated model requires raw syngas to be diverted and combusted to supplement the heat delivered to the gasifier by the char combustor. The 5000 lb/hr waste stream effectively adjusts the energy flow in the current model to align with this soon to be released study.

The final plant power specifications for this scenario are summarized in Table B.3.

Table B.3 – Plant Power Generation

	HP	*MW*
WPLANT	47435	35.4
WGEN	-43163	-32.2
WTURB	-114652	-85.5
WNET	**-110380**	**-82.3**

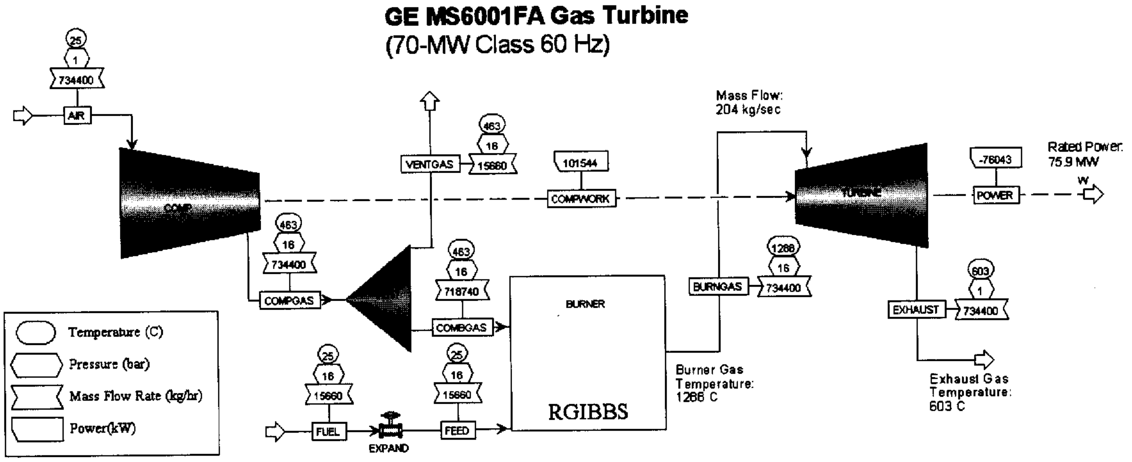
APPENDIX C

ASPEN PLUS GAS TURBINE MODEL DETAILS

Table C.1 – GE Gas Turbine Specifications

	MS6001FA
Output (MW)	75.9 (90)
Heat Rate (kJ/kWh)	10,332
Pressure Ratio	15.7:1
Mass Flow (kg/sec)	204
Turbine Speed (rpm)	5,254
Exhaust Temp (°C)	603
Turbine Inlet Temp (°C)	1288

Based on specifications from GE (summarized in Table C.1), a simple ASPEN Plus model was constructed of the 6FA simple cycle gas turbine. The flow sheet is shown below.


GE MS6001FA Gas Turbine
(70-MW Class 60 Hz)

The model was calibrated in the following manner:

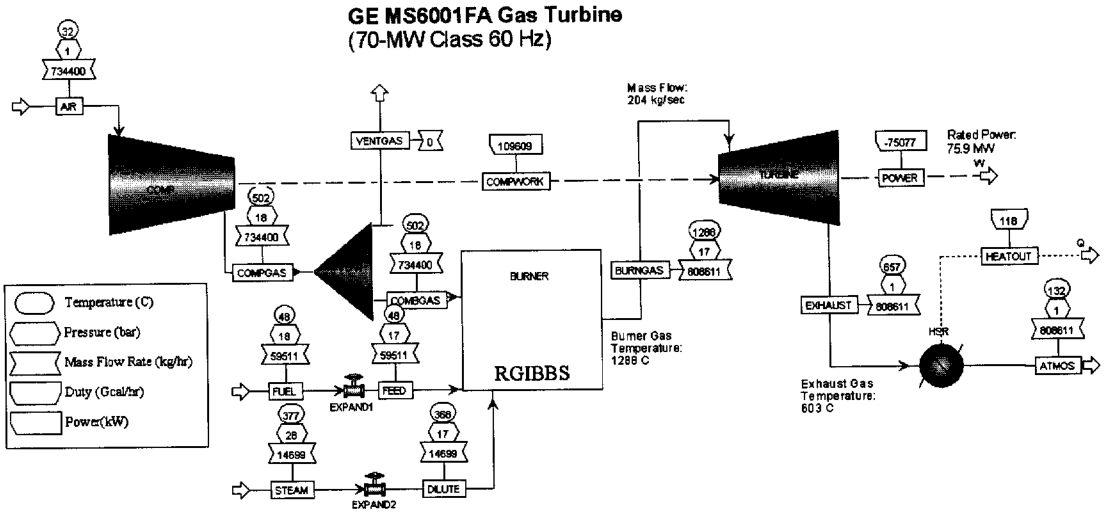
3. The heat rate was set so that 10,332 kJ/kWh of Methane (LHV) to the RGIBBS burner at ambient temperature and 15.7 bar pressure.
4. The mass flow through the compressor was set to 204 kg/sec of air with a pressure ratio of 15.7.
5. A calculator block was specified that sets the total "BURNGAS" flow to 204 kg/sec (vents compressor gas as needed).
6. The compressor isentropic efficiency was adjusted until burner temperature reached 1288 °C.
7. The turbine isentropic efficiency was adjusted until the outlet temperature was 603 °C.
8. Finally, the compressor mechanical efficiency was adjusted until the power output was approximately 75.9 MW

Table C.2 – ASPEN Plus NG Model Specifications

	Compressor	Turbine
Mech Efficiency (%)	93	100
Isentropic Efficiency (%)	77.8	92.5
Compression Ratio	15.7	-
Outlet Pressure (psia)	-	14.7

Table C.2 summarizes the natural gas compressor and turbine settings used for all simulations. In order to run the turbine on hydrated syngas, the following additional steps were taken:

1. The feed was adjusted until the burner temperature reached the target temperature of 1288 C°.
2. The calculator block was modified so that up to 14% extra flow (over the 204 kg/sec limit) could be passed through the turbine.
3. The compressor pressure ratio was adjusted so that the volumetric flow through the turbine remained the same as the base natural gas case.



In order to estimate the combined cycle outputs, the following additional steps were taken:

1. An additional heater block was inserted in the turbine exhaust stream that took the exhaust temperature to 132 °C.
2. The heat duty of this heater block was taken times the efficiency of .40 to estimate the ideal steam cycle power available.

APPENDIX D

ELECTRIC HEATER CAPITAL COSTS

Capital cost estimate for adding only electric, combustion-air preheating to the gasifier plant.

Cost of 2.2 MW Unit today	$	250,000.00
Size Priced	kW	2,200.00
Electricity to Heat Eff	-	0.90
Size Needed	kW	96,191.97
Scaling Factor	-	0.9
Cost of 86.5 MW Unit today	$	7,491,771.63
Cost of 86.5 MW Unit 2005	$	6,396,147.84
Installation Factor	-	2.47
Total '05 Installed Cost	$	15,798,485.16

Cost of Base Air Blower	$	34,860.00
Size Priced	kW	1,857.54
Size Needed	kW	3,885.10
Scaling Factor	-	0.6
Cost of Needed Unit today	$	54,275.74
Cost of Needed Unit 2005	$	46,338.26
Installation Factor	-	2.47
Total '05 Installed Cost	$	114,455.51
Base Case Cost	$	91,105.77
Incremental Increase	$	23,349.74

Total Capital	**15,821,834.90**

Capital cost estimate for completely replacing syngas recycle stream with electric heat.

Cost of 2.2 MW Unit today	$	250,000.00
Size Priced	kW	2,200.00
Size Needed	kW	86,572.77
Scaling Factor	-	0.9
Cost of 86.5 MW Unit '09	$	6,814,010.35
Cost of 86.5 MW Unit 2005	$	5,817,504.82
Installation Factor	-	2.47
Total '05 Installed Cost of Air Heater	$	14,369,236.91

Cost of Base Air Blower	$	34,860.00
Size Priced	kW	1,857.54
Size Needed	kW	5,827.64
Scaling Factor	-	0.6
Cost of Needed Unit today	$	69,224.61
Cost of Needed Unit 2005	$	59,100.96
Installation Factor	-	2.47
Total '05 Installed Cost	$	145,979.36
Base Case Cost	$	91,105.98
Incremental Increase	$	54,873.38

Inconel Heat Capacity	W/cm^2	13.00
Size Needed	kW	23,632.72
Area Needed	cm^2	1,817,901.56
Diameter of Rod	cm	1.59
Length of Rod Needed	m	3,639.45
Cost of Rod (2008)	$/m	175.43
Total Material Cost Today	$	638,476.92
Cost of Heater Factor	-	2.50
Total Cost of Heater Today	$	1,596,192.31
Cost of Heater 2005	$	1,362,759.37
Installation Factor	-	2.47
Total '05 Installed Cost	$	3,366,015.65

Total Capital Range for Modification **17,790,125.93**

APPENDIX E

DIRECT GASIFIER BASELINE CAPITAL COSTS

	Number Required	Number Spares	Equipment Name	Units	Scaling Stream Flow	Original Stream Flow	Size Ratio	Original Equipment Cost (per Unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2005$	Subtotals	Source
Feed Prep	2		Feed Prep	wet tonne/hr	83	65	1.29	$ 9,840,000	2003	$ 19,680,000	0.77	$ 23,935,511	1	$ 23,935,511	$ 27,897,950	$ 27,897,950	b
Gasifier	2		Gasifier	dry tonne/hr	73	42	1.76	$ 6,410,000	2003	$ 12,820,000	0.7	$ 19,026,792	1	$ 19,026,792	$ 22,176,609		b
	2		Primary Cyclone	actual m3/s gas	10	69	0.14	$ 910,000	2003	$ 1,820,000	0.7	$ 469,033	1	$ 469,033	$ 546,680	$ 22,723,289	b
ASU	1		ASU	tonne/hr pure O2	28	77	0.36	$ 22,700,000	2003	$ 22,700,000	0.5	$ 13,680,124	1	$ 13,680,124	$ 15,944,820		b
	1		N2 Compressor	MWe consumed	2	10	0.20	$ 4,140,000	2003	$ 4,140,000	0.67	$ 1,402,618	1	$ 1,402,618	$ 1,634,816		b
	1		O2 Compressor	MWe consumed	4	10	0.45	$ 5,540,000	2003	$ 5,540,000	0.67	$ 3,225,740	1	$ 3,225,740	$ 3,759,750	$ 21,339,385	b
Syngas Cleaning	1		Syngas Cooler	kW	60,000	77,000	0.78	$ 25,400,000	2003	$ 25,400,000	0.6	$ 21,869,056	1	$ 21,869,056	$ 25,489,400		b
	1	1	Sludge Pump	kg/hr	500	452	1.11	$ 3,911	2002	$ 7,822	0.33	$ 8,084	2.47	$ 19,968	$ 23,633		a
	1		Sludge Settling Tank	kg/hr	3,706	3,706	1.00	$ 11,677	2002	$ 11,677	0.6	$ 11,677	2.47	$ 28,842	$ 34,135		a
	1		LO-CAT Oxidizer Vessel	kg/hr	360	234	1.54	$ 1,000,000	2002	$ 1,000,000	0.65	$ 1,324,631	2.47	$ 3,271,837	$ 3,872,281		a
	1		ZnO Bed Preheater	kW	18,189	15,121	1.20	$ 71,389	2002	$ 71,389	0.44	$ 77,435	2.47	$ 191,263	$ 226,364		a
	2		ZnO Sulfur Removal Beds	kg/hr	117,088	83,771	1.40	$ 37,003	2002	$ 74,006	0.56	$ 89,269	2.47	$ 220,494	$ 260,959	$ 29,906,771	a
Fuel Synthesis	1		High Temperature Shift Reactor	kg/hr	161,102	146,450	1.10	$ 465,907	2002	$ 465,907	0.56	$ 491,461	2.47	$ 1,213,908	$ 1,436,683		a
	1		LT Shift Precooler/BFW Preheater #1	kW	17,523	25,035	0.70	$ 56,089	2002	$ 56,089	0.6	$ 45,281	2.47	$ 111,843	$ 132,369		a
	1		Low Temperature Shift Reactor	kg/hr	161,102	146,451	1.10	$ 323,464	2002	$ 323,464	0.56	$ 341,204	2.47	$ 842,774	$ 997,439		a
	1		PSA Precooler/Deaerator Water Preheater #2	kW	9,793	9,207	1.06	$ 21,089	2002	$ 21,089	0.6	$ 21,885	2.47	$ 54,055	$ 63,975		a
	1		PSA Air-cooled Precooler	kW	25,785	31,283	0.82	$ 388,064	2002	$ 388,064	0.6	$ 345,572	2.47	$ 853,563	$ 1,010,207		a
	1		Pre-PSA Knock-out #1	kg/hr	161,102	146,451	1.10	$ 129,979	2002	$ 129,979	0.6	$ 137,632	2.47	$ 339,950	$ 402,337		a
	1		Pre-PSA Knock-out #2	kg/hr	135,940	111,591	1.22	$ 55,291	2002	$ 55,291	0.6	$ 62,242	2.47	$ 153,738	$ 181,952		a
	1		Pressure Swing Adsorption Unit	kg/hr	7,120	6,950	1.02	$ 4,855,471	2002	$ 4,855,471	0.6	$ 4,926,334	2.47	$ 12,168,045	$ 14,401,109	$ 18,626,072	a
Steam Cycle	1		Based on Installed Cost of 474.34 $/kWh	kWh	36,380	29,973	1.21	$ 14,217,400	2002	$ 14,217,400	1	$ 17,256,489	1	$ 17,256,489	$ 20,423,378	$ 20,423,378	a

Subtotal	$ 140,916,844	
Contingency (1.5%)	$ 2,113,753	a
Structures	$ 6,368,900	a
Total	**$ 149,399,497**	

Sources:

	a	Spath et. al. "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier". NREL/TP-510-37408. May 2005.
	b	Jin, H. et al. "Performance and cost analysis of future, commercially mature gasification-based electric power generation from switchgrass". Biofpr Vol 3 pp 142-173. 2009.

APPENDIX F

DIRECT GASIFIER HYBRID SYSTEM CAPITAL COSTS

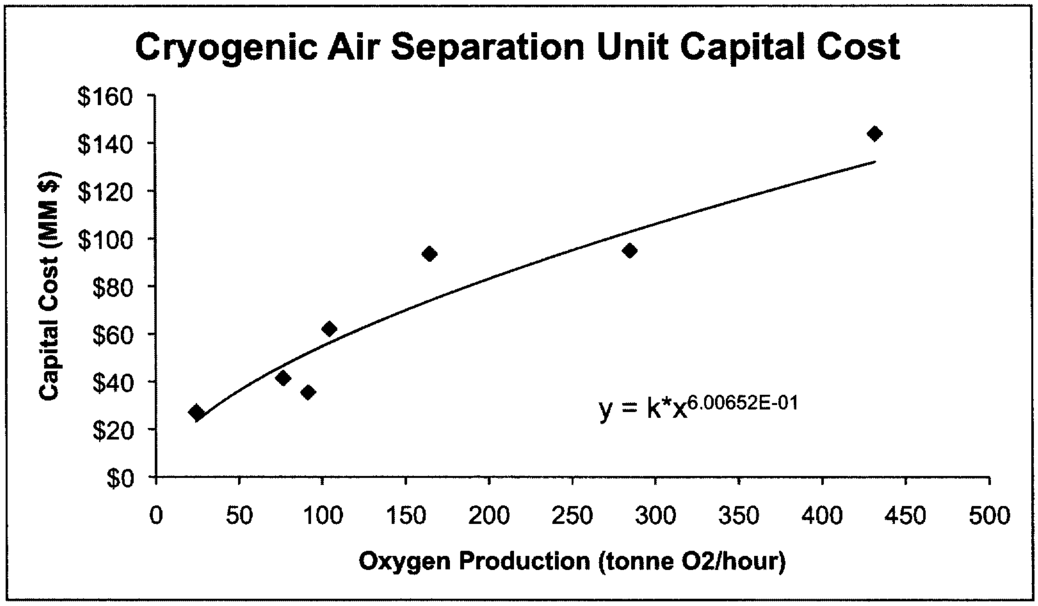
	Number Required	Number Spares	Equipment Name	Scaling Stream Units	Scaling Stream Flow	Original Stream Flow	Size Ratio	Original Equipment Cost (per Unit)	Base Year	Total Original Equip Cost (Req'd & Spare) in Base Year	Scaling Exponent	Scaled Cost in Base Year	Installation Factor	Installed Cost in Base Year	Installed Cost in 2005$	Subtotals	Source
Feed Prep	2		Feed Prep	wet tonne/hr	83	65	1.29	$ 9,840,000	2003	$ 19,680,000	0.77	$ 23,935,511	1	$ 23,935,511	$ 27,897,950	$ 27,897,950	b
Gasifier	2		Gasifier	dry tonne/hr	73	42	1.76	$ 6,410,000	2003	$ 12,820,000	0.7	$ 19,026,792	1	$ 19,026,792	$ 22,176,609		b
	2		Primary Cyclone	actual m3/s gas	10	69	0.14	$ 910,000	2003	$ 1,820,000	0.7	$ 469,033	1	$ 469,033	$ 546,680	$ 22,723,289	b
Electrolyzer	1		Electrolyzer Plant	Future_H2A											$ 95,402,426		c
	1		O2 Compressor	Mwe consumed	4	10	0.45	$ 5,540,000	2003	$ 5,540,000	0.67	$ 3,225,740	1	$ 3,225,740	$ 3,759,750	$ 99,162,176	b
Syngas Cleaning	1		Syngas Cooler	kW	60,000	77,000	0.78	$ 25,400,000	2003	$ 25,400,000	0.6	$ 21,869,056	1	$ 21,869,056	$ 25,489,400		b
	1	1	Sludge Pump	kg/hr	500	452	1.11	$ 3,911	2002	$ 7,822	0.33	$ 8,084	2.47	$ 19,968	$ 23,633		a
	1		Sludge Settling Tank	kg/hr	3,706	3,706	1.00	$ 11,677	2002	$ 11,677	0.6	$ 11,677	2.47	$ 28,842	$ 34,135		a
	1		N2 Compressor	Mwe consumed	2	10	0.20	$ 4,140,000	2003	$ 4,140,000	0.67	$ 1,402,618	1	$ 1,402,618	$ 1,634,816		b
	1		2-Stage Selexol Acid Gas Removal	kg/hr	117,280	310,258	0.38	$ 59,505,000	2006	$ 59,505,000	0.6	$ 33,193,515	1	$ 33,193,515	$ 31,107,293		d
	1		Elemental Sulfur Recovery	kg/hr	23	5,676	0.00	$ 11,997,000	2006	$ 11,997,000	0.6	$ 440,234	1	$ 440,234	$ 412,565	$ 58,701,843	d
Fuel Synthesis	1		High Temperature Shift Reactor	kg/hr	161,102	146,450	1.10	$ 465,907	2002	$ 465,907	0.56	$ 491,461	2.47	$ 1,213,908	$ 1,436,683		a
	1		LT Shift Precooler/BFW Preheater #1	kW	17,523	25,035	0.70	$ 56,089	2002	$ 56,089	0.6	$ 45,281	2.47	$ 111,843	$ 132,369		a
	1		Low Temperature Shift Reactor	kg/hr	161,102	146,451	1.10	$ 323,464	2002	$ 323,464	0.56	$ 341,204	2.47	$ 842,774	$ 997,439		a
	1		PSA Precooler / Deaerator Water Preheater #2	kW	9,793	9,207	1.06	$ 21,089	2002	$ 21,089	0.6	$ 21,885	2.47	$ 54,055	$ 63,975		a
	1		PSA Air-cooled Precooler	kW	25,785	31,283	0.82	$ 388,064	2002	$ 388,064	0.6	$ 345,572	2.47	$ 853,563	$ 1,010,207		a
	1		Pre-PSA Knock-out #1	kg/hr	161,102	146,451	1.10	$ 129,979	2002	$ 129,979	0.6	$ 137,632	2.47	$ 339,950	$ 402,337		a
	1		Pre-PSA Knock-out #2	kg/hr	135,940	111,591	1.22	$ 55,291	2002	$ 55,291	0.6	$ 62,242	2.47	$ 153,738	$ 181,952		a
	1		Pressure Swing Adsorption Unit	kg/hr	7,120	6,950	1.02	$ 4,855,471	2002	$ 4,855,471	0.6	$ 4,926,334	2.47	$ 12,168,045	$ 14,401,109	$ 18,626,072	a
Steam Cycle	1		Based on Installed Cost of 474.34 $/kWh	kWh	36,380	29,973	1.21	$ 14,217,400	2002	$ 14,217,400	1	$ 17,256,489	1	$ 17,256,489	$ 20,423,378	$ 20,423,378	a

Subtotal	$ 247,534,706	
Contingency (1.5%)	$ 3,713,021	a
Structures	$ 6,368,900	a
Total	**$ 257,616,627**	

Sources:		
	a	Spath et. al. "Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier". NREL/TP-510-37408. May 2005.
	b	Jin, H. et al. "Performance and cost analysis of future, commercially mature gasification-based electric power generation from switchgrass". Biofpr Vol 3 pp 142-173. 2009.
	c	Future Central Hydrogen Production from Grid Electrolysis version 2.1.1 Available at: http://www.hydrogen.energy.gov/h2a_prod_studies.html
	d	"Cost and Performance Baseline for Fossil Energy Plants Vol 1". DOE/NETL-2007/1281

APPENDIX G
AIR SEPARATION UNIT CAPITAL COST

Seven published capital cost estimates were found in a literature review and used to estimate the scaling factor for cryogenic air separation units. The data points and a fitted curve are shown in the figure below. The raw data is given in the corresponding table. Based on these values, a scaling factor of 0.6 is appropriate for air separation unit capital cost scaling.



Base Capacity	*Ratio*	*Base Cost*	*Scaling Factor*	*Source*
tonne O2/hr		**2005 $**		
77	1.000	$41,423,155	0.5	A
285	3.720	$95,022,376	-	B
92	1.197	$35,401,762	-	C
165	2.151	$93,500,000	-	D
104	1.364	$62,133,621	-	E
24	0.313	$27,310,677	0.75	F
432	5.642	$143,731,325	-	G

The 77 tonne/hour numbers were used as a baseline because the numbers were perceived as relatively reliable and were based on actual hardware quotes.

References Used:

A – Kreutz, T. Williams, R. Consonni, S. Chiesa, P. 2005. Co-production of hydrogen, electricity and CO2 from coal with commercially ready technology. Part B: Economic Analysis. International Journal of Hydrogen Energy. 30:769-784. (engineering and contingency removed)

B & G – Dillon, D.J. Panesar, R.S. Wall, R.A. Allam, R.J. White, V. Gibbins, J. Haines, M.R. Oxy-combustion processes for CO2 Capture from Advanced Supercritical PF and NGCC Power Plant. (less contingencies, fees, and owners costs)

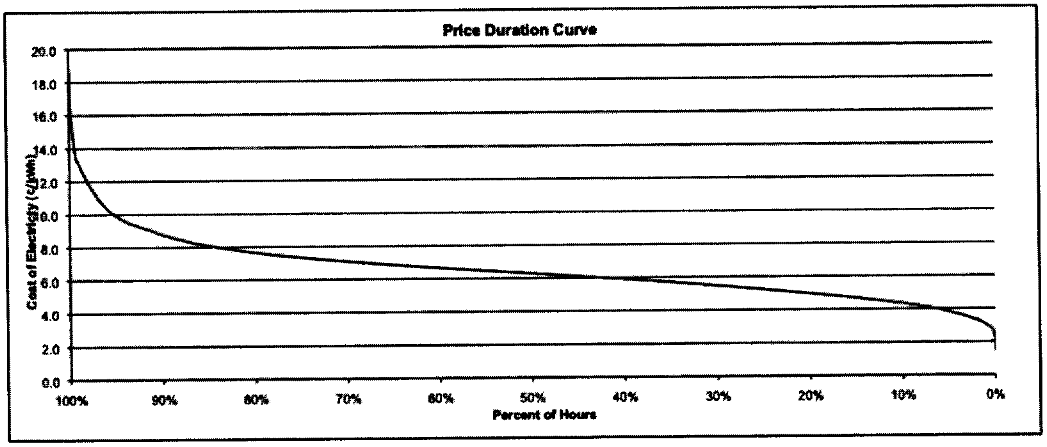
C & E – Parsons Corporation. 2003. Capital and Operating Cost of Hydrogen Production from Coal Gasification.

D – Apt, J. Newcomer, A. Lave, L. Douglas, S. Dunn, L.M. 2008. An Engineering-Economic Analysis of Syngas Storage. National Energy Technology Laboratory. NETL-2008/1331.

F – Tijmensen, M.J.A. Faaij, A.P.C. Hamelinck, C.N. vanHardeveld, M.R.M. 2002. Exploration of the possibilities for production of Fischer Tropsch liquids and power via biomass gasification. Biomass and Bioenergy. 23:129-152.

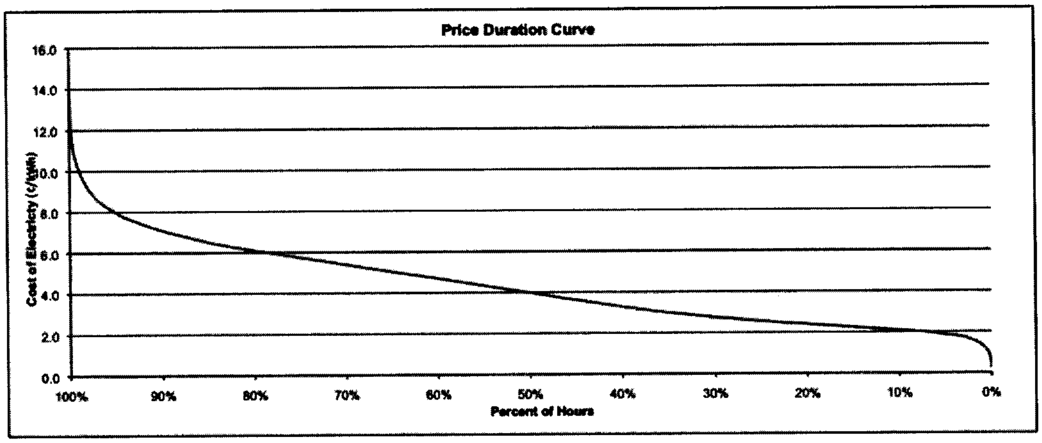
APPENDIX H
ELECTRICITY PROFILES

Northeast ISO



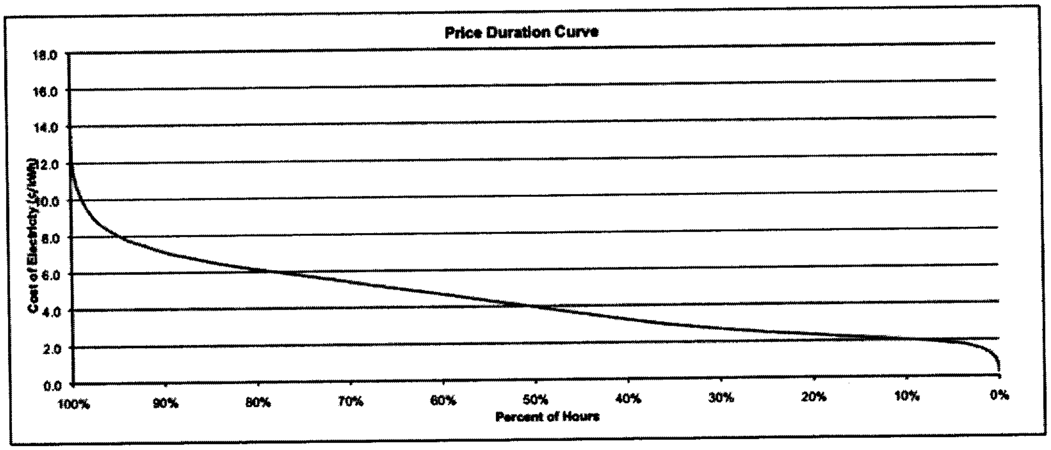
- High average cost because of reliance on natural gas turbines
- Hourly day ahead market data received directly from NE ISO website.
- 6.55 ¢/kWh average price of electricity

Midwest ISO



- Low cost of electricity because of large amount of coal on the grid
- Hourly day ahead market data received from MISO directly
- 4.33 ¢/kWh average price of electricity

Northwest Interface



- Low cost of electricity because of large amount of hydro on the grid
- No RTO/ISO in the area so data from the northwest interface used instead
- Hourly load lambda data received from DOE records
- 4.31 ¢/kWh average cost of electricity

APPENDIX I

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I approve.

Christopher Kinchin
Biomass Process Analysis Section
National Renewable Energy Laboratory
1617 Cole Blvd.
Golden, CO 80401-3393

Phone: (303) 384-7709
Fax: (303) 384-6363
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