Integrated Pyrolysis Combined Cycle Biomass Power System Concept Definition Final Report

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Abstract

Advanced power systems based on integrated gasification/combined cycles (IGCC) are often presented as a solution to the present shortcomings of biomass as fuel. Although IGCC has been technically demonstrated at full scale, it has not been adopted for commercial power generation. Part of the reason for this situation is the continuing low price for coal. However, another significant barrier to IGCC is the high level of integration of this technology: the gas output from the gasifier must be perfectly matched to the energy demand of the gas turbine cycle.

We are developing an alternative to IGCC for biomass power: the integrated (fast) pyrolysis/ combined cycle (IPCC). In this system solid biomass is converted into liquid rather than gaseous fuel. This liquid fuel, called bio-oil, is a mixture of oxygenated organic compounds and water that serves as fuel for a gas turbine topping cycle. Waste heat from the gas turbine provides thermal energy to the steam turbine bottoming cycle.

Advantages of the biomass-fueled IPCC system include: combined cycle efficiency exceeding 37 percent efficiency for a system as small as 7.6 MW_e; absence of high pressure thermal reactors; decoupling of fuel processing and power generation; and opportunities for recovering value-added products from the bio-oil.

This report provides a technical overview of the system including pyrolyzer design, fuel clean-up strategies, pyrolysate condenser design, opportunities for recovering pyrolysis byproducts, gas turbine cycle design, and Rankine steam cycle. The report also reviews the potential biomass fuel supply in Iowa, provide and economic analysis, and present a summery of benefits from the proposed system.

Table of Contents

Lists of Graphical Materials

Tables

Figures

1.0 Introduction

Biomass would appear to be a logical fuel blend in existing coal-fired power plants. Both coal and biomass are solid fuels requiring more difficult storage, handling, and firing provisions compared to liquid or gaseous fuels. However, biomass has distinct physical and chemical characteristics that complicate the simple substitution of biomass for coal in existing boilers.

Advanced power systems based on integrated gasification combined cycles (IGCC) are often presented as a solution to the present shortcomings of biomass as fuel. Although IGCC has been technically demonstrated at full scale, it has not been widely adopted for power generation. Part of the reason for this situation is the continuing low price for coal. However, another significant barrier to IGCC is the high level of integration of this technology: the gas output from the gasifier must be perfectly matched to the energy demand of the gas turbine cycle.

This project evaluated an alternative to IGCC for biomass power: integrated (fast) pyrolysis combined cycle (IPCC). In this system, solid biomass is converted into liquid rather than gaseous fuel. This liquid fuel, called pyrolytic oil, is a mixture of oxygenated hydrocarbons and water that serves as fuel for a gas turbine topping cycle. Waste heat from the gas turbine provides thermal energy to a steam turbine bottoming cycle.

Advantages of the biomass-fueled IPCC system include: cycle efficiency exceeding that of biomass-fired Rankine cycles; does not require high pressure thermal reactors; reduces the strong coupling between fuel processing and power generation typical of integrated power systems; provides opportunities for recovering value-added products.

The overall objective of this project was to evaluate the feasibility of a biomass power system based on IPCC. Specific objectives included:

- Design a pyrolyzer to convert solid biomass to liquid fuel.
- Evaluate strategies for pretreatment of biomass feedstock and post-treatment of pyrolysate vapors to yield fuel clean enough for gas turbine applications.
- Design a combined gas turbine cycle and Rankine bottoming cycle that utilizes pyrolysate as fuel.
- Evaluate the characteristics of two pyrolysate recovery strategies.
- Optimize a Rankine bottoming cycle.
- Evaluate opportunities for separating high-value byproducts from pyrolytic products.
- Evaluate fuel supply availability and cost.
- Determine economic costs of the proposed biomass power system.

2.0 Executive Summary

The IPCC concept combines a biomass pyrolysis plant with a conventional combined cycle power plant. In this system, solid biomass is converted into a liquid "bio-oil" that is a mixture of oxygenated hydrocarbons and water. This liquid serves as fuel for a gas turbine topping cycle. Waste heat from the gas turbine provides thermal energy to a steam turbine bottoming cycle.

Numerous configurations were examined in the conceptual design process. The designs included various levels of integration, heat recovery and efficiency, value-added chemical production, and other configurations. The overall design minimizes the integration among the pyrolysis and combined cycle plants, allowing the two to operate independently if necessary. In some cases, less efficient equipment was selected when significant cost savings could be realized.

Two basic configurations are presented in this report: Maximum Power Production Design in which the entire bio-oil output of the pyrolysis plant is used to fuel the combined power plant and Value-Added Chemicals Design where significant operational modifications and added costs allow recovery of value-added chemicals from the bio-oil resulting in lower flow of bio-oil to the power cycle. Oat hulls and corn stover were examined as potential fuels.

The proposed cycle could be commercially offered in the near term with minimal research and development needs. In contrast, implementation of the value-added products design will require the development of new technologies and will entail some technical risk.

The proposed IPCC system is estimated to have a total project cost of $18.4 million, which, based on a net combined cycle output of 7,655 kW, is equal to $2,400/kW. For a small amount of additional capital, recovery of valuable chemicals may also be performed. The IPCC capital cost compares favorably to conventional biomass power systems in this size range, which cost between $2,000 and $3,000/kW.

The economic analysis evaluated a number of sensitivity cases to assess the impacts of key variables on the project economic viability. Key variables that were considered in the sensitivity analysis include: Capital cost, Operating and maintenance cost, Fuel cost and Anhydrosugar value. The analysis has shown that the maximum power option would not be competitive with traditional utility assets; however, this configuration appears to be competitive with other biomass options. Value-added chemical production greatly improves the economics of the project. To obtain the minimum equity return (12%), a renewable energy credit of only $18/MWh (equal to the current production tax credit) would be needed for oat hulls.

3.0 Technical

This section provides an overview of the technical design of the integrated pyrolysis combined cycle (IPCC) power plant.

In simple terms, the IPCC concept combines a biomass pyrolysis plant with a conventional combined cycle power plant. In this system, solid biomass is converted into a liquid "bio-oil" that is a mixture of oxygenated hydrocarbons and water. This liquid serves as fuel for a gas turbine topping cycle. Waste heat from the gas turbine provides thermal energy to a steam turbine bottoming cycle.

Numerous configurations were examined in the conceptual design process. The designs included various levels of integration, heat recovery and efficiency, value-added chemical production, and other configurations. Two basic configurations are presented in this report:

- **Maximum Power Production Design** – In this design the entire bio-oil output of the pyrolysis plant is used to fuel the combined power plant. A simple condenser is used to separate the oil from the non-condensable gases, and biomass pretreatment is minimal.

- **Value-Added Chemicals Design** – In this design, value-added chemicals are extracted from the bio-oil resulting in lower flow of bio-oil to the power cycle. Recovery of the chemicals requires additional biomass pre-treatment equipment, a fractionating condenser, and significant operational modifications and added costs.

The chosen design was developed by consensus among project team members as the best approach balancing technical risk, complexity, efficiency, and cost. In general, the overall design minimizes the integration among the pyrolysis and combined cycle plants, allowing the two to operate independently if necessary. In some cases, less efficient equipment was selected when significant cost savings could be realized. This was particularly true in the steam cycle design, as high efficiency equipment in this small size range demands a substantial cost premium. The maximum power production design includes largely standard equipment that involves minimal technical risk. Because of this approach, it is believed that the proposed cycle could be commercially offered in the near term with minimal research and development needs. In contrast, implementation of the value-added products design will require the development of new technologies and will entail some technical risk.

The conceptual cycle process flow is shown in Figure 3-1, and a general arrangement drawing is shown in Figure 3-2.

Figure 3-1. IPCC Power Plant Conceptual Design.

Figure 3-2. IPCC Power Plant General Arrangement.
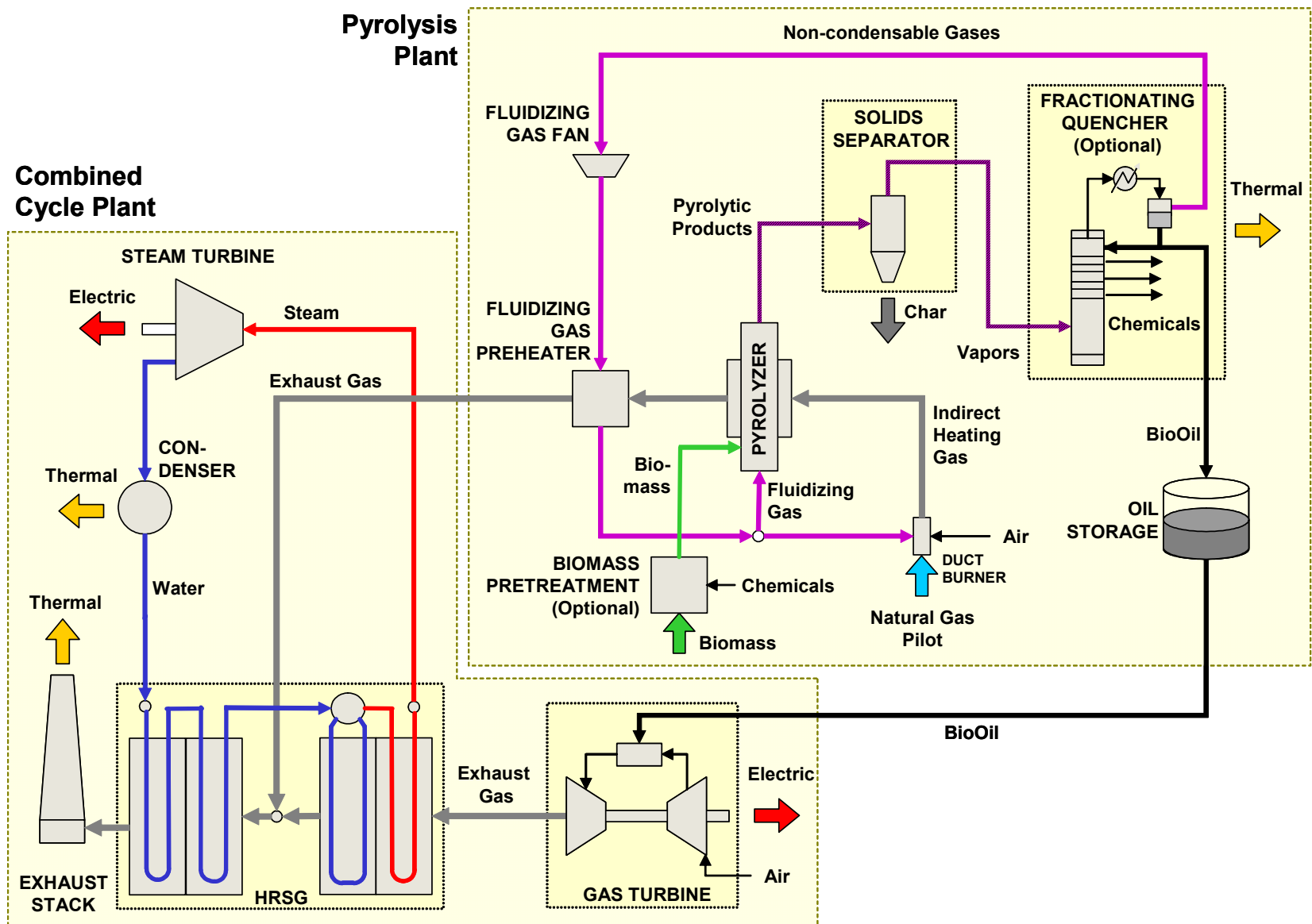
As mentioned previously, the cycle design consists of a pyrolysis plant and a combined cycle plant. The major components of each of these plants include the following:

- Pyrolysis Plant
  - Biomass storage and handling system
  - Biomass pretreatment (*optional for recovery of value-added products*)
  - Biomass pyrolyzer
  - Solids separator for recovery of char
  - Condenser to separate bio-oil and non-condensable gases *(optional replacement with a fractionating condenser for recovery of value-added products)*
  - Bio-oil storage
  - Fluidizing gas and indirect heating gas systems
- Combined Cycle Plant
  - Combustion turbine generator
  - Heat recovery steam generator
  - Steam turbine generator
  - Cycle heat rejection equipment

Table 3-1 highlights the major design criteria for the project. Note that for the purposes of this level of evaluation, the properties of the two biomass fuel types were assumed to be similar. The only major relevant difference between the two at this stage is the fuel price. Combined with several other logistical issues described in Section 4 (Fuel Supply Availability and Cost), the lower fuel price of oat hulls makes it a promising candidate for the first application of the technology.

The remainder of this section describes critical technical design features of the various plant systems in detail. These sections include:

- Pyrolyzer design
- Fuel clean-up strategies
- Pyrolysate condenser design
- Opportunities for recovering pyrolysis byproducts (optional value-added products)
- Gas turbine cycle design
- Rankine steam cycle design

Table 3-1. Process Design Criteria.		
	Metric Units	**English Units**
Biomass Information		
Biomass fuel type	Oat hulls/ corn stover	Oat hulls/ corn stover
Biomass moisture content (as fired)	6.1 percent	6.1 percent
Biomass higher heating value (LHV, as fired)	15,665 kJ/kg	6,732 Btu/lb
Pyrolysis Plant		
Biomass fuel feed rate (wet)	5.39 tonne/hr	5.94 ton/hr
Pyrolysis plant capacity factor	90.4 percent	90.4 percent
Annual biomass consumption	42,700 tonne/yr	47,100 ton/yr
Annual biomass consumption (energy, LHV)	669,000 GJ/yr	633,800 MBtu/yr
Pyrolysis yield (bio-oil / char)	68% / 20%	68% / 20%
Total annual bio-oil production	29,000 tonne/yr	32,000 ton/yr
Combined Cycle Plant		
Total annual bio-oil consumption	29,000 tonne/yr	32,000 ton/yr
Combined cycle plant capacity factor	80 percent	80 percent
Bio-oil fuel feed rate	4.14 tonne/hr	4.57 ton/hr

3.1 Pyrolyzer Design

This section describes in detail a fast pyrolysis process that converts a solid biomass waste, such as oat hulls and corn stover, into a liquid biofuel that can be used as a substitute for diesel oil or other hydrocarbon fuels to produce power in gas turbine engines and boilers. This technology has been developed and successfully demonstrated in DynaMotive Energy Systems pilot plants. The process utilizes a deep bubbling fluidized bed design combined with a novel bio-oil recovery system. The properties of bio-oil produced from agriculture biomass waste feedstock including oat hulls and forest residues such as bark and wood are given in detail, particularly in reference to their application as a fuel for gas turbine engines.

3.1.1 Pyrolysis Introduction

In the agricultural production process a substantial amount of the crop becomes a fibrous residue such as oat hulls, corn stover, and rice hulls. Traditionally, much of this solid waste product has been left in the field or incinerated in stationary boilers to produce steam for the process. As agricultural biomass often contains significant quantities of silica, its application as a fuel creates many operational problems due to

both the glazing (fouling) of spreader stoker equipment and high temperature heat transfer surfaces and the accelerated erosion of steel tubing exposed to the abrasive particles in the flue gas. In addition, the fouling and high ash characteristics of combusted biomass prevent the direct use of the combustion products in direct fired gas turbines for efficient electrical generation. As a consequence, disposal of this residue has been problematic, inefficient and often expensive to the industry.

It is also a well-established fact that combustion of fossil fuels such as coal, oil and natural gas for power generation is a significant contributor to global warming (UNDP, 1999; IEA, 1996; Environment Canada, 1997). On the other hand biomass has long been identified as an alternate sustainable source of renewable energy (Yan et al., 1997). However, power generation using a solid fuel has had significant limitations with respect to materials handling requirements and efficient energy conversion. Converting biomass fuel into a liquid addresses these issues and makes possible the use of higher efficiency combined cycle systems for power generation. "Fast pyrolysis" technology is a unique process that converts solid biomass waste materials, such as oat hulls, into a relatively clean burning liquid fuel called bio-oil that is also carbon dioxide and greenhouse gas (GHG) neutral. The DynaMotive Energy Systems Corp. fast pyrolysis technology has proven to be a reliable and efficient technology for converting a wide range of biomass to a high energy density liquid fuel that handles like a heavy fuel oil.

The nearest term commercial application for bio-oil is as clean fuel for generating power and heat from small stationary diesel engines, gas turbines and boilers. DynaMotive Energy Systems Corp is working with Orenda Aerospace Corporation to test and develop bio-oil fuels for power generation (Fuleki et al., 2000) using their OGT 2500 gas turbine package, with an ISO rating of 2.85 MW_e.

3.1.2 Fast Pyrolysis of Biomass

Fast pyrolysis (more accurately defined as thermolysis) is a process in which biomass material, such as oat hulls, is rapidly heated to high temperatures in the absence of oxygen. The oat hulls decompose into a combination of solid char, gas, vapors and aerosols. When cooled, most volatiles condense to a liquid referred to as "bio-oil". The remaining gases comprise a medium calorific value non-condensable gas.

Bio-oil is a liquid mixture of oxygenated compounds containing various chemical functional groups (Bridgewater et al, 1999; Gabra and Kjellstrom, 2002), such as carbonyl, carboxyl and phenolic. Bio-oil is made up of the following constituents: 20-25 percent water, 25-30 percent water-insoluble pyrolytic lignin, 5-12 percent organic acids, 5-10 percent non-polar hydrocarbons, 5-10 percent anhydrosugars and 10-25 percent other oxygenated compounds.

DynaMotive Energy Systems Corp owns a specific fast pyrolysis technology (Morris, 1999) patented by its partner Resource Transforms International (RTI). This fast pyrolysis process is described in the patent titled, "Energy Efficient Liquefaction of Biomaterials by Thermolysis" (Piskorz et al.).

In this particular fast pyrolysis process, biomass feedstock is introduced into a thermolysis reactor having a bed of inert material, such as sand, with a height to width ratio greater than one. The biomass is shredded to sufficiently small dimensions so that its size does not limit significantly the production of the liquid product fraction. Simultaneous introduction of pre-heated, non-oxidizing gas at sufficient linear velocity performs two principal functions: firstly, as a medium for fluidizing the hot sand bed and secondly, to cause automatic elutriation of the product char from the fluidized bed reactor. The process includes removing the elutriated char particles from the effluent reactor stream and rapidly quenching the gas, aerosols and vapors to produce a high conversion yield of liquid bio-oil. For maximum yield of liquid, the thermolysis reaction must take place within a period of a few seconds at temperatures ranging from 450°C to 500°C. The products must then be quenched as soon as possible to prevent cracking of the newly produced bio-oil.

Figure 3-3. BioTherm^{TM} 15 TPD Pilot Plant, Vancouver, BC.
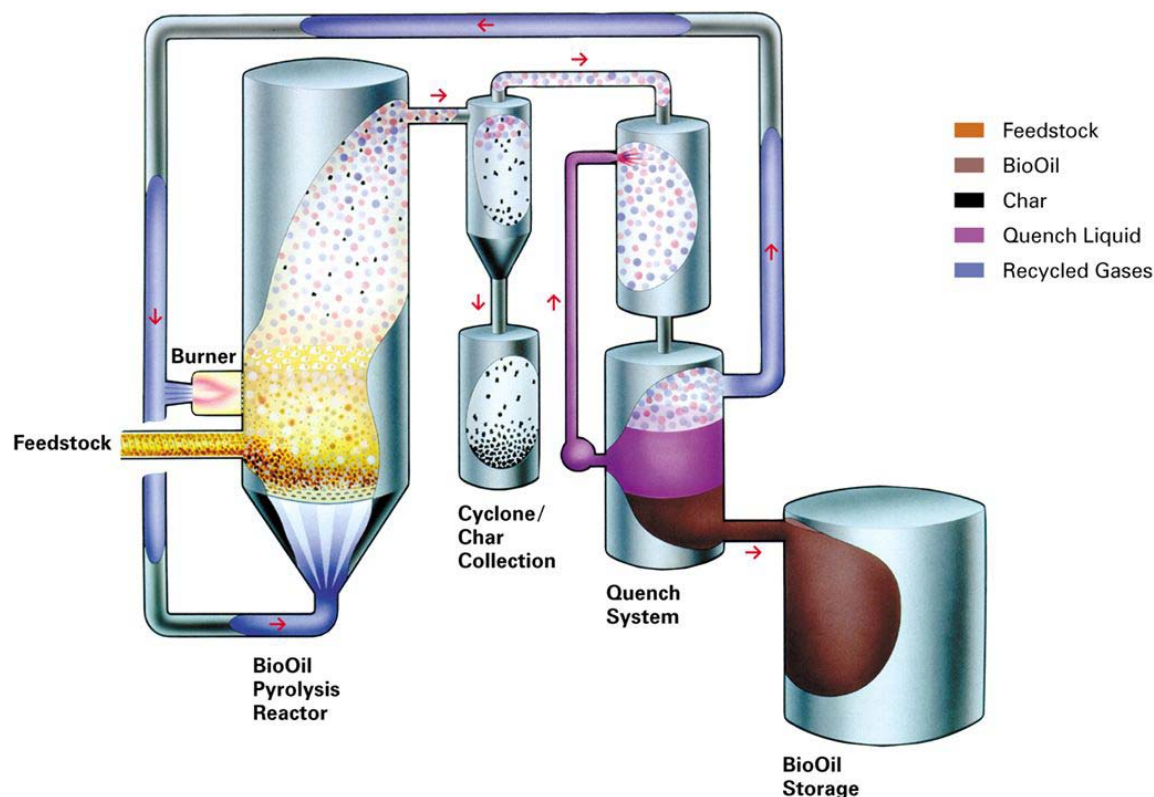
Figure 3-4. BioTherm^{TM} Flow Sheet.

3.1.3 Fast Pyrolysis Heat and Mass Balance

Feedstock for the fast pyrolysis process can be any biomass waste material including agricultural wastes and wood by-products. To maximize yield and minimize process development risk, DynaMotive has focused on near term bio-oil production from oat hulls, corn stover, sugarcane bagasse and wood waste (white wood and bark) feedstock. For the DOE Alliant Energy study the key biomass fuels under consideration are oat hulls and corn stover. The chemical properties of these two materials are sufficiently similar to enable them to be treated as the same for this level of analysis.

Preparation includes drying the feedstock to less than 10 percent moisture content to minimize the water content in the bio-oil and then grinding the feed to small particles to ensure rapid heat transfer rates in the reactor.

When processing bagasse feedstock, the conversion yield to liquid bio-oil, solid char and non-condensable gas is approximately 68 percent, 20 percent and 12 percent by weight, respectively, on an as fed basis. These yield rates were identical to those determined previously in laboratory size apparatus using the same operating conditions (Piskorz, 1999). For the oat hull and corn stover biomass the anticipated commercial yields are the conversion yield to liquid bio-oil, solid char and non-condensable gas is approximately 68 percent, 20 percent and 12 percent by weight, respectively, on an as fed

basis. The dry ground oat hull material appears to be suitable for direct use as a pyrolysis feedstock for 68 percent bio-oil yield after additional hammer mill preparation. Depending on the moisture content of the as-received corn stover, additional drying may be necessary. However, for the purposes of this investigation, it is assumed that in-field and on site drying are sufficient to result in an "as-fired" moisture content of 10 percent or less. Further discussion of biomass properties are provided in Section 4, Fuel Supply Availability and Cost.

The heat required for thermolysis is the total heat that must be delivered to the reactor to provide all the sensible, radiation and reaction heat for the process to proceed to completion. The heat of reaction for the fast pyrolysis process is marginally endothermic. When operating the pilot plant using prepared feedstock, the total heat requirement to produce bio-oil at a 68 percent yield rate (including radiation and exhaust gas losses) is approximately 2.5 MJ per kilogram of bio-oil produced. The net heat required from an external fuel source, such as natural gas, is only 1.0 MJ per kilogram of bio-oil produced. This applies when the non-condensable gases produced in the process are directly injected into the reactor burner. This represents approximately 5 percent of the total calorific value of the bio-oil being produced.

To date the operation and scale-up of the DynaMotive BioTherm fast pyrolysis technology has shown very reliable, stable operation with predicable fuel characteristics and energy yield. The consistency of the BioTherm bio-oil products has been critical in long term stable handling characteristics of the bio-oil produced with this technology.

3.1.4 Bio-Oil Analysis

Bio-oil is a dark brown liquid that is free flowing. It has a pungent smoky odor. Bio-oil contains several hundred different chemicals with a wide-ranging molecular weight distribution.

The following Table 3-2 lists the properties of bio-oil produced by the BioTherm™ pilot plant, derived from three different biomass feedstocks.

The density of bio-oil is high, approximately 1.2 kg/liter. On a volumetric basis bio-oil has 55 percent of the energy content of diesel oil and 40 percent on a weight basis.

The solids entrained in the bio-oil principally contain fine char particles that are not removed by the cyclones. As can be seen, the solids in the bio-oil have been reduced significantly to levels of approximately 0.1 percent by weight. The ash content in these solids ranges from 2 percent to 20 percent, depending on the ash content in the feedstock.

Table 3-3 lists the chemical composition of major constituents in the bio-oil produced in the BioTherm™ process derived from different feedstock, namely bagasse, wood/bark, and wood.

Table 3-2. DynaMotive Bio-Oil Properties.			
Biomass Feedstock	**Bagasse**	**Pine/Spruce (53% Wood + 47% Bark)**	**Pine/Spruce (100% Wood)**
Moisture wt%	2.1	3.5	2.4
Ash Content wt%	2.9	2.6	0.42
Bio-Oil Properties			
pH	2.6	2.4	2.3
Water Content wt%	20.8	23.4	23.3
Lignin wt%	23.5	24.9	24.7
Solids Content wt%	< 0.10	< 0.10	< 0.10
Ash Content wt%	< 0.02	< 0.02	< 0.02
Density kg/L	1.20	1.19	1.20
Calorific Value MJ/kg	15.4	16.4	16.6
Kinematic Viscosity, cSt @20°C	57	78	73
Kinematic Viscosity, cSt @80°C	4.0	4.4	4.3

Table 3-3. Bio-Oil Composition (Wt%).			
Biomass Feedstock	**Bagasse**	**Pine/Spruce (53% Wood + 47% Bark)**	**Pine/Spruce (100% Wood)**
Water	20.8	24.3	23.3
Lignin	23.5	24.9	24.7
Cellobiosan	-	1.9	2.3
Glyoxal	2.2	1.9	2.3
Hydroxyacetaldehyde	10.2	10.2	9.4
Levoglucosan	3.0	6.3	7.3
Formaldehyde	3.4	3.0	3.4
Formic Acid	5.7	3.7	4.6
Acetic Acid	6.6	4.2	4.5
Acetol	5.8	4.8	6.6

3.1.5 Market Competitiveness

In the proprietary DynaMotive process, bio-oil and char are commercial products and the non-condensable gases are recycled back into the process. No waste is produced in the DynaMotive process. The overall simplicity of the technology gives DynaMotive significant competitive advantages over other pyrolysis technologies in terms of capital and operating costs. DynaMotive's process also produces higher quality fuel and higher yields of bio-oil compared to competing technologies. This section provides an overview of the DynaMotive process production economics. A complete economic analysis of the overall system including the integrated pyrolysis portion is provided in Section 5, Economic Analysis.

DynaMotive has set a fully burdened production cost target of $3 per gigajoule (GJ) of net plant output (US$18.36 /BOE). This target was developed based on historic and current cost projections and an understanding of the potential optimizations yet to be realized in the current process.

Over the short term, these targets are expected to be met by concentrating on process optimization and minimizing expenses (e.g. low cost feedstocks, such as surplus agricultural biomass; heat recovery) and over the longer term by innovation and new technologies to balance increasing expenses (e.g. self generation, cutting edge drying technologies).

DynaMotive's target cost for bio-oil production is $3 per gigajoule (GJ) of net plant output (US$18.36 /BOE) based on a 400 tonne per day (tpd) commercial plant. The earliest and most appropriate market application for bio-oil is as a clean fuel to produce "green power" and heat in boilers, kilns, gas turbines and diesel engines. Over the past year, the company has established partnerships with gas turbine manufacturers (Orenda Aerospace, Solar/Caterpillar) to develop packaged bio-oil fueled gas turbine power systems and is collaborating with NRCan's CANMET combustion laboratory to develop burner nozzles for industrial boiler/kiln applications.

DynaMotive is also developing higher value fuel and chemicals applications for bio-oil including ethanol blended fuels, diesel/bio-oil emulsions and catalytic upgrading of bio-oil to synthesis gas, which can be converted to synthetic transport fuels and biomethanol for use in hydrogen fuel cells.

3.1.6 Conclusions

Converting biomass wastes produced from agriculture and forestry operations to a liquid bio-oil, using DynaMotive fast pyrolysis technology, has been demonstrated at the pilot plant level as a reliable and repeatable process. Test programs to demonstrate bio-oil application as a fuel in gas turbine engines, diesel engines and boilers are underway with

a host of engine manufacturers. To date the results have been most encouraging for small scale, high efficiency conversion of biomass to electric power opportunities. For example, Figure 3-5 compares the efficiency of three biomass electricity generation options for a Tanzanian sugar mill. Use of gas turbine combined cycle as proposed in this study appears particularly promising.
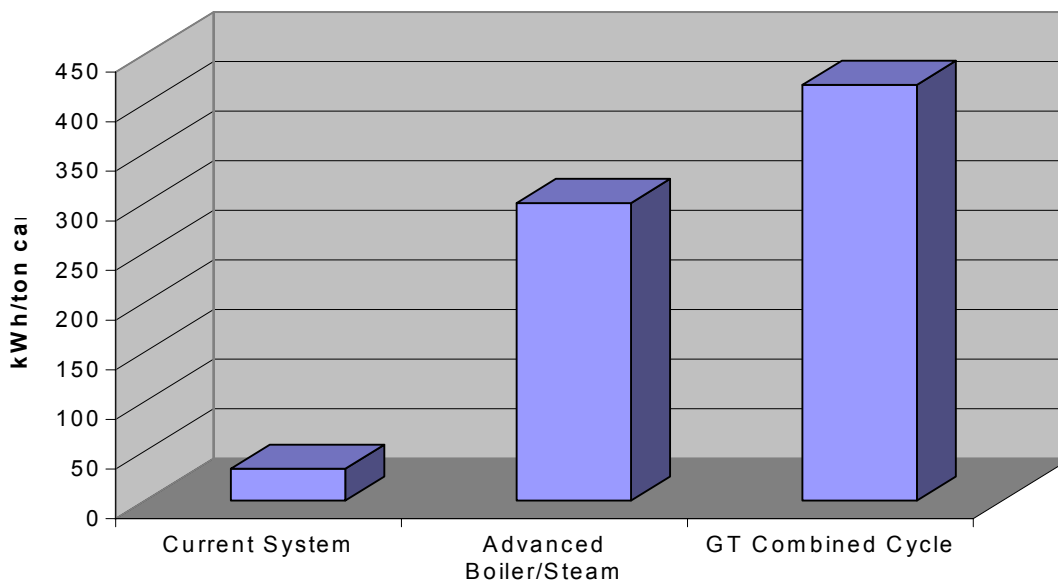
Figure 3-5. Comparison of Electrical Generating Options for a Tanzanian Sugar Mill (ref. Gabra and Kjellstrom).

3.2 Fuel Clean-up Strategies

Both pre and post-pyrolysis strategies were investigated to produce a suitable bio-oil feedstock for production of power and / or value-added chemicals. This section describes the technical and economic considerations of these systems.

3.2.1 Technology

Current systems for production of bio-oil yield products of only marginal quality as gas turbine fuel. Although pyrolytic products transported out of a pyrolyzer pass through a solids separator to remove char and ash from the gas flow, a significant quantity of particulate matter is often found in the condensed bio-oil. Most fast pyrolysis systems employ a gas cyclone to remove char and ash, which is inadequate for particulate matter finer than 10 μm (33 μft). The result is bio-oil containing fine char, which gives the liquid a characteristic dark brown or black color (Bridgwater et al., 1999).

The fine char in bio-oil represents an undesirable contaminant. Char is reported to catalyze chemical reactions that degrade pyrolysis liquids and promotes polymerization (Agblevor et al., 1995; Diebold et al., 1996). This char also interferes with atomization of bio-oil when injected into gas turbine combustors, which extends vaporization and burnout times of the fuel. Furthermore, alkali in biomass becomes concentrated in char upon pyrolysis (Diebold et al., 1996). Alkali can produce corrosion of gas turbine blades at even very low concentrations.

Both pre-pyrolysis and post-pyrolysis strategies for removing contaminants from the bio-oil were investigated. The pre-pyrolysis strategy consists of removing alkali by washing the biomass. Even rinsing biomass with water has proved effective in removing alkali, which exists in water-soluble form (Dayton et al., 1999; Turn et al., 1997). However, since removal of char also removes most of the alkali, the washing strategy was not pursued as a means of contaminant control. On the other hand, it is an integral element in the production of certain high-value products, as described in the section on recovery of chemicals from bio-oil. The rest of this section focuses on filtration of char from bio-oil as a dual-contaminant control strategy.

Although removal of char from bio-oil will simultaneously solve the particulate and alkali contamination problems, bio-oil is not readily filtered because the organic compounds in the liquid tend to clump around the char particles (Elliott, 1994). More effective is hot gas clean up of gases and vapors exiting the pyrolyzer.

DynaMotive includes a high efficiency cyclone in its standard pyrolysis plant design, and this system was included as part of the base case conceptual design of the IPCC plant. If further removal of char is necessary, alternate methods are available. For example, experiments have shown that a ceramic candle filter operating at temperatures above the condensation points of the pyrolysis vapors effectively removes fine char. Ash content as low as 0.01 percent and alkali content as low as 3 ppm have been obtained by vapor filtration of bio-oil at 400º C (752° F) (Diebold, 1996). However, long-term operation of candle filters for this purpose is problematic. Pyrolytic char is cohesive and the resulting dust cake is not readily removed from surface of a ceramic candle filter.

A more promising approach to hot vapor filtration may be the use of a moving bed granular filter. These filters pass hot, dirty gas through a bed of refractory granules, where fine particulate matter impacts and adheres to the granules. The granules continuously flow downward by gravity through the bed, removing the dirty granules and replacing them with fresh media (Saxena et al., 1985). Thus, even moderately cohesive dust will flow out of the filter without fouling the filter.

Iowa State University has developed a moving bed granular filter (MBGF) for dry scrubbing of product gas from biomass gasifiers, which may prove suitable for hot

filtration of pyrolytic vapors. In this filter, illustrated in Figure 3-6, granular material moving downward by gravity spills from a centrally located dipleg to form an interfacial region where dust cake forms and most particulate removal occurs. The lower edge of the filter cake is dispersed by the downward flow of granular material while the upper interface is continuously covered by a fresh layer of granular material cascading from the dipleg above the interface. In this fashion, the interface establishes a dust cake of quasi-steady thickness, which is controlled to give high collection efficiency and acceptable pressure drop. Details of the filter construction and operation are found in Brown et al. (2000). Recent cold-flow evaluations of the filter yielded particulate removal efficiencies exceeding 99 percent (Colver et al., 2002).   Iowa State University is presently constructing a MBGF to operate on the hot vapors from a fast pyrolysis pilot plant.

3.2.2 Economics

The base conceptual design includes a high efficiency cyclone as a standard component for char removal. Its cost is nominal and is included in the overall pyrolysis plant cost estimate described in Section 5.

Moving bed granular filters are not commercially available at this time. However, Energy Products of Idaho (EPI), which is evaluating the commercial potential for this kind of filter, estimates that a unit designed for 606 m^3/min (21,400 acfm) of gas will cost $400,000.   The pyrolysis power system proposed in the present study, processing 5,390 kg of biomass/h, would be considerably smaller than the unit being evaluated by EPI. The flow rate of pyrolytic vapors and gases is estimated as follows:

Approximately 15 percent of the feedstock is converted into non-condensable gases; application of the ideal gas law at the pyrolysis temperature of 450º C (840º F) and atmospheric pressure indicates at non-condensable gas flow of 28 m^3/min (1003 acfm). Using the rule of thumb that the flow rate of condensable vapors is approximately three times that of non-condensable gases, yields an approximate volumetric flow rate through the hot filter of 84 m^3/min (3009 acfm).
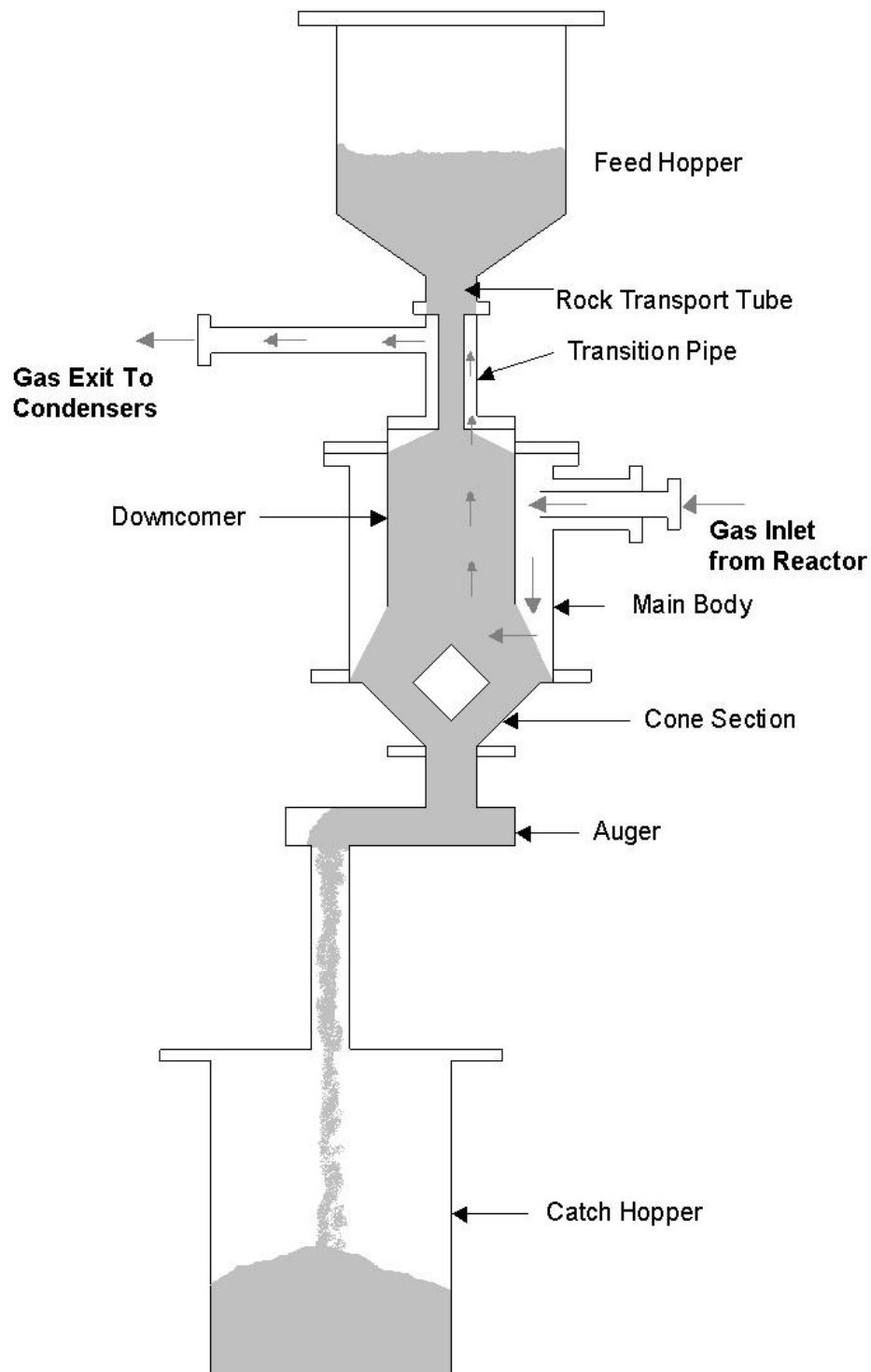
Figure 3-6. Moving Bed Granular Filter for Removing Particulate Matter from Pyrolysis Vapors.

Applying the "sixth-tenths" scaling rule suggests that the cost of a moving bed granular filter for a 5,390 kg/h (12,077 lb/h) plant will be:

$$C = \$400,000 \ (84/606)^{0.6} = \$122,220 \text{ (Free on Board or f.o.b)}$$

This does not include installation costs, which are conservatively estimated by the methodology outlined in Table 3-4 (Ulrich, 1984). Inclusion of these costs results in a total module cost of $294,000 for the moving bed granular filter.

3.3 Pyrolysate Condenser Design

Many commercial and pilot-plant pyrolysis systems employ a wet scrubber or quench system to recover pyrolysis liquid from the gas stream exiting the solids separator. In this case, cold pyrolysis liquid is sprayed into the hot gas stream entering the scrubber/quencher, cooling the gas and forcing pyrolysis vapors to condense on droplets of cold liquid. Essentially all the heat added to the pyrolyzer at 400 – 500º C (752 - 932° F) is recovered here as low quality heat at temperatures below 100º C (212° F).   This approach is wasteful from the perspective of integrated energy management and does not allow separation of different compounds from the bio-oil since they are all condensed in a single chamber. In addition to the standard quench condenser included in the base case conceptual design, two alternative approaches to collecting condensate were explored: one proposed by Iowa State University and the other by DynaMotive. These are each discussed in the following paragraphs.

3.3.1 Staged Condenser

Iowa State is investigating the feasibility of a pyrolysate condenser, which employs several water-cooled condensers in series to gradually cool the hot vapors exiting the fast pyrolysis reactor. The concept offers the possibility of recovering thermal energy at temperatures well above 100º C (212° F) as well as selectively condensing compounds in different sections of the condenser. This work, supported by the Iowa Energy Center, has not yet been published. However, preliminary results are reported here.

Table 3-4. Summary of Methodology for Capital Cost Calculations.

Cost	Calculation	Description
Direct Project Expenses		
Equipment (f.o.b.)	C_p	"Free on board" or f.o.b – cost when placed aboard the shipping carrier. Obtained from suppliers or estimated from tabulations of purchased equipment costs or installed bare module costs (Blazej et al., 1993) Adjusted for size and inflation.
Materials for installation	$C_M = \text{MMF} \times C_p$	Based on tabulated materials module factors (MMF) found in References 1 and 2; expressed as fraction of C_p
Direct Labor	$C_L = \text{LMF} \times (C_p + C_M)$	Based on tabulated labor module factors (LMF) in (Agblevor et al., 1995) and (Bridgewater et al., 1999); expressed as fraction of $C_P + C_M$
TOTAL DIRECT	$C_D = C_P + C_M + C_L$	
Indirect Project Expenses		
Freight, insurance, taxes	$C_{FIT} = 0.08 C_P$	Estimated as 8 percent of total purchased equipment costs
Construction overhead	$C_O = 0.7 C_L$	Estimated as 70 percent of total labor installation costs
Engineering expenses	$C_E = 0.15(C_P + C_M)$	Estimated as 15 percent of total equipment and materials costs
TOTAL INDIRECT	$C_{ID} = C_{FIT} + C_O + C_E$	
BARE MODULE COST	$C_{BM} = C_D + C_{ID}$	Defined as combined direct and indirect costs
Contingency & fee	$C_{CF} = 0.18 C_{BM}$	Estimated at 18 percent of bare module cost
TOTAL MODULE COST	$C_{TM} = C_{BM} + C_{CF}$	Defined as bare module cost plus contingency and fee

The four-stage condenser is illustrated in Figure 3-7. Typical gas temperatures and catches of bio-oil for each stage are indicated on the drawing. Although quantitative analyses have not been performed on the bio-oils collected with this condenser system, qualitative differences among the four stages were evident. The first stage, which operates above the condensation temperature for water, is very viscous, indicating that it contains very little liquid water. In some trials, the bio-oil from the first stage of condensation is so viscous that it cools to a solid in the collection jar. Only about 5 percent of the bio-oil condenses in this stage. Careful design of this first stage of condensation, which employs a single, large diameter tube rather than several small diameter tubes, can prevent plugging of the condenser. The third stage, on the other hand, yields bio-oil of very low viscosity, which suggests that much of the water condenses from the gas flow at this point. Future work will quantify the distribution of water and organic compounds collected from each stage of condensation. The overall condenser was able to collect 75 percent of the total bio-oil yield, which is better than many spray quenchers used commercially.
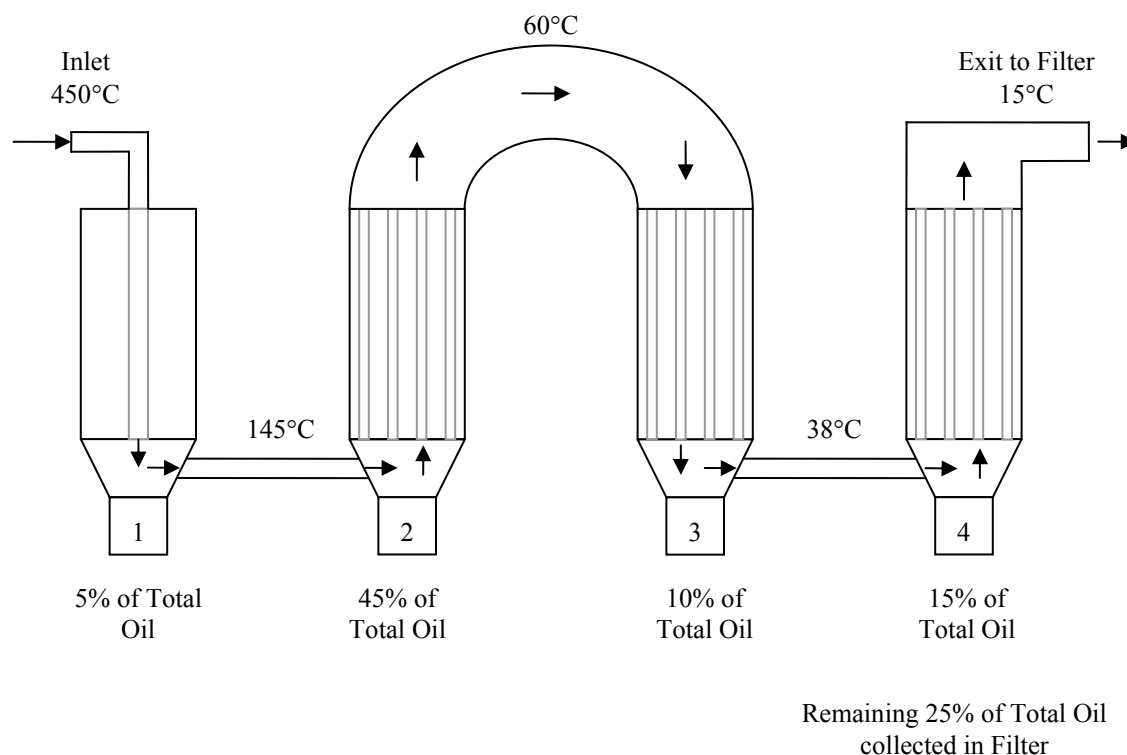
Figure 3-7. Pyrolysis Condensers: Water Cooled Shell-in-Tube Design.

This research showed that levoglucosan, of which the potential as a value-added product is discussed in the next section, was concentrated to 46 weight-percent in the first stage of a four-stage condenser while the second stage contained only 18 weight-percent of levoglucosan. Enrichment occurred despite the lack of careful temperature control at condensation surfaces. This approach to separation of pyrolysis products takes its inspiration from "integral reactive distillation" in which products are removed in-situ before side reactions can occur that produce undesirable products (Garside, 1994).

The original reason for investigating the staged condenser was to recover waste heat for use in feedwater heaters in the bottoming steam cycle. However, analysis of the power cycle dismissed the value of this waste heat for this purpose and the focus of the pyrolysate condenser turned to recovery of chemicals from the bio-oil. DynaMotive suggested the use of a fractionating quencher rather than the staged condenser envisioned by Iowa State University. The fractionating quencher, described in the following paragraphs, allows some recovery of organic chemicals and a relatively modest amount of low temperature waste heat.

3.3.2 Fractionating Quencher

The change in character on demineralization of biomass can be appreciated from Table 3-5, which compares the bio-crude oil (BCO) from untreated and demineralized softwood. The latter has a substantially lower content of low molecular weight carbonyls resulting in a BCO of greatly diminished volatility (a fact which also has implications for pyrolyzer performance). Gas yields are also dramatically lower. Similar changes upon demineralizing corn stover are evident in Table 3-6.

Some additional points of note:

- When the BCO is hydrolyzed to convert anhydrosugars to free sugars, the glucose equivalent yield of fermentable sugars is appreciably greater than that calculated on the basis of monomers (levoglucosan/anhydromannose) and dimers (cellobiosan) listed. Further, this is apart from the automatic 11 percent weight gain of glucose over levoglucosan. Thus the sugars listed in the table underestimate the total fermentable sugar yield.
- Although acetic acid is low for this feedstock, in some cases in can be as high as 6 - 10 percent.

Table 3-5. Pyrolysis Products for Woody Biomass (Bone-Dry Western Hemlock; Reaction Temperatures 450 – 490º C (842 - 914° F).

	None	De-mineralization
Pretreatment	None	De-mineralization
Additive	None	0.1 percent $(NH_4)_2SO_4$
Hexosan Content of Feed (weight percent)	~ 62	~ 62
Yields (weight percent of feed)		
Gas	12.0	5.1
Coke/Char	18.2	17.3
BCO	65.9	74.2
BCO Composition (weight percent):		
Water	14.3	12.8
Cellobiosan	0.5	2.6
Levoglucosan	2.4	16.8
Anhydromannose	0.0	5.3
Glycolaldehyde	10.6	2.2
Acetic Acid	1.4	1.6
Other C_i-C_3 Oxo Compounds	7.4	~ 0
Pyrolytic Lignin	19.9	21.6
Percent Hexosan Conversion to Sugars	3	33

Table 3-6. Pyrolysis Products for Herbaceous Biomass (Bone-Dry Corn Stover; Reaction Temperatures 450 – 490º C (842 - 914° F).

	None	De-mineralization
Pretreatment	None	De-mineralization
Additive	None	0.1 percent $(NH_4)_2SO_4$
Yields (weight percent of feed)		
Gas	22.6	7.9
Coke/Char	15.8	13.2
BCO	61.6	78.9
BCO Composition (weight percent)		
Water	~4.2	13.2
Cellobiosan	Trace	3.3
Levoglucosan	2.8	20.1
Glycolaldehyde	11.6	3.7
Acetic Acid	3.4	1.3
Pyrolytic Lignin	~33.4	17.7

The following is a first proposal for a scheme to recover these chemicals together with the potential to recover heat. It should be understood that this is hardly a definitive proposal as there are many associated uncertainties. Some specific examples are:

- Boiling points of various specific compounds and condensate fractions are highly uncertain. These will have to be established more definitively by experimentation.

- On account of the reactive nature of fresh bio-oil, subjecting it to repetitive heating and cooling sequences could cause irreversible changes in its physico-chemical properties that render some aspect of the scheme inoperable; for example by forming polymeric materials that cause fouling of surfaces.

- The viscosity of the various condensate fractions could play an important role in the possible designs for heat exchangers. While there are published studies of viscosity of whole bio-oil and an identification of some of the factors that influence it, the available information is insufficient to predict the rheological properties of specific fractions.

- The question is whether one can one build a column that is efficient and has low residence times (to minimize secondary reactions to which the BCO is prone). The high stability of the aerosols is problematic as are the low diffusion coefficients in viscous liquid fractions.

With these caveats, the proposed design, shown in Figure 3-8, is based on the following considerations:

- The irreversible nature of BCO vaporization suggest that we attempt to fractionate it during the cooling process rather than by re-boiling, using the sensible heat of the pyrolyzer effluent as the driving force for fractionation rather than through a reboiler.

- Unfortunately levoglucosan appears to have a boiling point that partially overlaps those of some of the higher MW phenolic monomers. This is not necessarily a serious problem as the phenols, unlike the anhydrosugars, have low water solubility.   However this aspect is neglected in the proposed scheme.

- The aqueous overhead condensate is re-mixed with the bottoms (largely pyrolytic lignin) to restore lower viscosity.

- This mixture is adjusted to a proper temperature and injected into the column above the entry point of the BCO vapors at a rate sufficient to lower the temperature from about 480° C (896° F) to about 300° C (572° F).
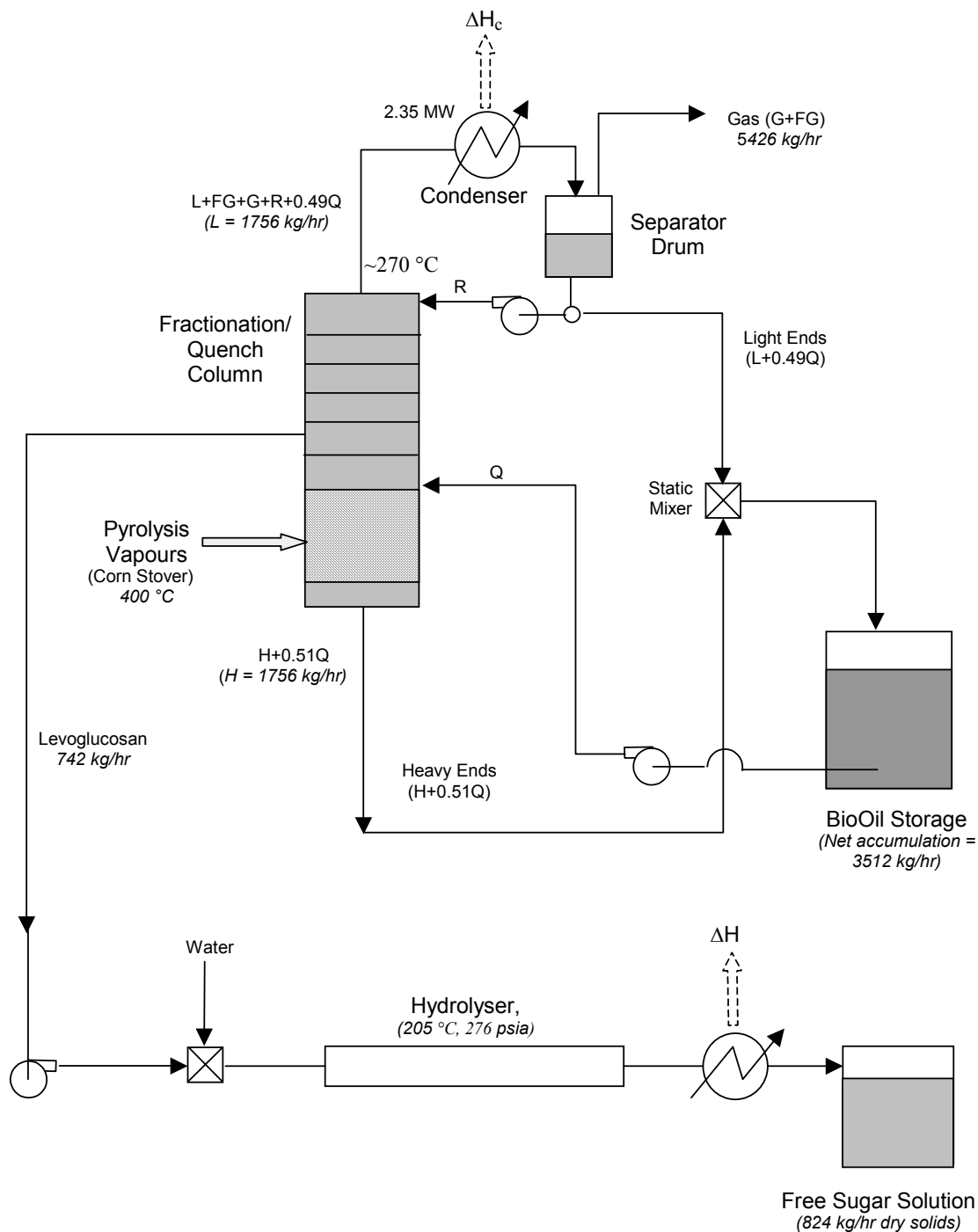
Figure 3-8. Concept for Fractionating Quencher.

- Optionally the difference in solubility of the pyrolytic lignin and anhydrosugars in water can be exploited by extracting the bottoms from the column with water to remove oligomeric sugars of low volatility.

Assuming the raw biomass material is de-mineralized corn stover, a rate of 5,390 kg/h (11,883 lb/h) of corn stover to the pyrolyzer will generate (0.789 + 0.079) x 5,390 = 4,679 kg/h (10,316 lb/h) of volatiles (V). The volatiles comprise gas (G) at 426 kg/h (938 lb/h), water (W) at 562 kg/h (1,240 lb/h) and organics (O) at 3,692 kg/h (8,139 lb/h).

$$V = W+G+O$$

In view of the uncertainties mentioned above, we shall apply a highly simplified scheme in which it will be assumed that the bio-oil is composed of the following components:

- The heavy fraction is composed mainly of pyrolytic lignin (PL)
- Levoglucosan (LG)
- A residual light fraction boiling below 300° C (572° F)
- A residual heavy fraction boiling above 300° C (572° F)

The residual organics exclusive of LG and PL will be arbitrarily assumed to be divided equally between light and heavy fractions, that is, we write for the total light (L) and heavy (H) fractions, respectively:

$$L = W + ½(O\text{-}LG\text{-}PL) = 1,711 \text{ kg/h } (3,751 \text{ lb/h})$$
$$H = PL + ½(O\text{-}LG\text{-}PL) = 1,824 \text{ kg/h } (3,978 \text{ lb/h})$$
$$LG = 742 \text{ kg/h } (1,639 \text{ lb/h}),$$
so that overall, the total bio-oil (B) is divided into:
$$B = LG + H + L.$$

After separation of LG, the stored part of the bio-oil is H + L. The boiling points of L and H will also be arbitrarily assumed to be 120° C (248° F) and greater than 300° C (572° F) respectively.

For simplicity, the numerical values of the material flows shown in the process diagram assume that the only chemical recovered is levoglucosan – no phenols, acetic acid or other light fractions; however, as pointed out earlier, we believe that a high valued phenolic fraction could be easily recovered.

Levoglucosan production rate is 742 kg/h (1,639 lb/h) assuming 100 percent recovery efficiency. Since hydrolysis to free sugars involves the addition of a water

molecule to each LG molecule, this would correspond to a production rate of 824 kg/h (1,811 lb/h) of free sugars.

Katritzky et al (1996) have described what appears to be a cheap and simple method that involves nearly stoichiometric hydrolysis of levoglucosan with high-pressure water. They report that a 0.30 Molar aqueous solution of LG was quantitatively converted to predominantly glucose and traces of another glucose isomer at 205° C and 1,906 kPa (401° F and 276 psi) so the required conditions are relatively mild and thus, such a process step would be expected to be rather inexpensive. We show this operation as a simple heat exchange.

Cellobiosan (C), which is a dimer of levoglucosan, is expected to appear in H. We show on the process diagram a water extraction of the cooled heavy fraction to remove C and accumulate it with LG. This is tentative since although PL is insoluble in water, it would be surprising if there were not other components of H that are. In that case a simple water extraction without additional clean up (lime, adsorption, etc.) to remove potentially bio-toxic compounds may not be practical. Recovery of C would provide an additional 122 kg/h (270 lb/h) of hydrolysable sugars or 136 kg/h (302 lb/h) of fermentable sugars.

A reflux stream R composed of light ends controls the upper temperature of the fractionator temperature while a quench stream Q composed of the components in the stored bio-oil controls the lower temperatures. The stored bio-oil is comprised of the primary oil less the anhydrosugars. Its composition is thus 49 percent L and 51 percent H.

The magnitudes of the various heat transfer operations in the indicated directions are denoted by ΔH_i. We have not shown any coupling between these various heat flows as there is too much uncertainty at the present time. However the net heat available from the process would be

$$\Delta H_1 + \Delta H_2 + \Delta H_3 - \Delta H_4 - \Delta H_5$$

Of course not all of this is available at the same temperature. For example, ΔH_1 and ΔH_2 are available at about 300° C (572° F) but ΔH_3 is available at only about 120° C (248° F). In view of the uncertainties in heats of evaporation of the various fractions it is perhaps not very meaningful to speculate on the total available heat at this time. However, using a published (Oasmaa and Peacocke, 2001) heat capacity for bio-oil of 3.2 kJ/kg-K (0.7 Btu/lb-F) (strictly applicable only to the range 26-61° C or 78-142° F), and so considering only the sensible heat, and also using a baseline reference temperature of 20° C (68° F), we find that 2,280 MJ/h (2.2 MBtu/h) (i.e. from LG + H) would be

available at about 300° C (572° F) and 683 MJ/h (0.65 MBtu/h) (i.e. from L + G) would be available at about 120° C (248° F).

These may be compared with the net heat required for pyrolysis which is, very roughly, just the sensible heat required to raise the biomass feed to the pyrolysis temperature, namely about 5,390 MJ/h (5.39 MBtu/h) at about 480° C (896° F). Thus, roughly half of the input heat could be recovered usefully under the conditions stated in the previous paragraph.

It is clear that substantial additional experimental work will be required to make more definitive statements.

3.4 Opportunities for Recovering Pyrolysis Byproducts

As discussed in the introduction to this section, the project team has proposed two conceptual designs for this project:

- Maximum Power Production Design
- Value-Added Chemicals Design

Previous sections have discussed the implementation of fuel clean-up strategies and advanced pyrolysate condenser designers as optional equipment to the base case IPCC cycle design (that is, maximum power production design). The intent of this additional equipment is to target recovery of value-added chemicals, primarily the organic components of the bio-oil (and in particular anhydrosugars). In addition, the carbonaceous solid char recovered from the hot vapor filtration system is a potential feedstock for production of activated carbon. This section explores in further detail the opportunities for recovery of these byproducts.

3.4.1 Recovery of Activated Carbon from Pyrolysis Char

The proposed pyrolyzer is estimated to generate char at the rate of 776 kg/h (1,711 lb/h). With a heating value of 23.3 GJ/t (20.0 MBtu/t), this could provide significant energy to the steam turbine bottoming cycle. If not used on site, the fuel might also make an attractive boiler fuel for co-firing at existing coal fired power plants. However, potentially more lucrative is activation of the char with steam to yield about half the weight of char as an inexpensive sorbent (Scott et al., 1999). This byproduct, which already sequesters the original alkali content of the biomass feedstock, could be marketed as an activated carbon sorbent for the control of mercury emissions from coal-fired power plants. As part of this project, chars produced from the pyrolysis of corn stover and oat hulls was characterized and activated, as described in the following paragraphs.

3.4.1.1 *Preparation and characterization of chars*

The biomass used to produce the char was pyrolyzed in the system shown in Figure 3-9. The tested biomass materials used in this research include oat hull and corn stover. The biomass was injected into a fluidized sand bed reactor at approximately 500º C (932º F) with inert nitrogen as the fluidizing agent. The typical biomass feed rate was 7 kg/h (15 lb/h). This process typifies fast pyrolysis because of the high heating rates of the rather small particles of biomass (less than 1 mm or 0.003 ft). Once the biomass has been injected into the sand bed and pyrolyzes, the resulting gas stream exits the freeboard of the reactor and enters into two heated cyclones in series as labeled in Figure 3-9. The cyclones are enclosed in a heating jacket at 450º C (842º F), and therefore they are not in open view. This is the point in the flow in which the char is collected. A vast majority of the char (more than 95 percent) is collected by the first cyclone. Typical char yields at the prescribe temperature in the system are approximately 12 percent on a mass basis of biomass injected. At the end of the run, the char is removed from the cyclone collection bins, weighed, and sealed in containers. The remaining hot gas flow continues through a set of condensers and exits the system.
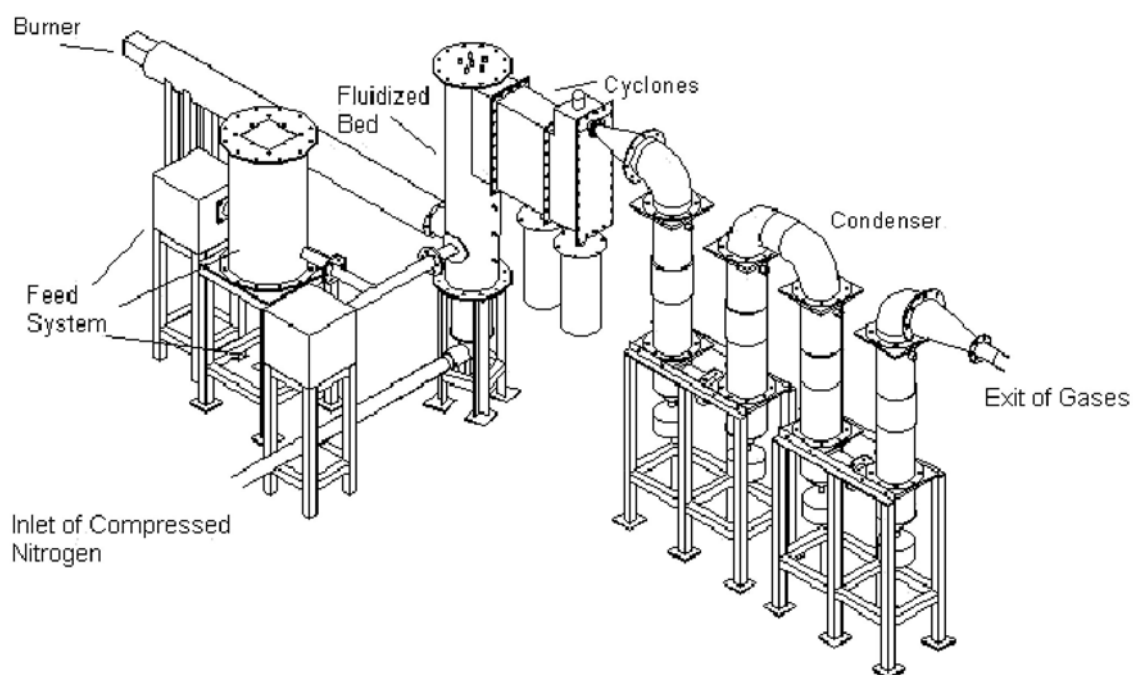
Figure 3-9. Schematic of Fast Pyrolysis Pilot Plant Used in the Production of Pyrolytic Chars from Oat Hulls and Corn Stover.

Contents of carbon in chars obtained from oat hull and corn stover were measured with thermogravimetric analyses, which were performed with a programmable Perkin Elmer TGA-7. Char samples were heated in an oven prior to the TGA tests to remove moisture in the same manner as used in the LOI tests. The first step of a TGA test was to load 25 ± 5 mg (60 ± 10 µlb)char sample into the TGA sample holder and then introduce nitrogen into the TGA furnace at a flow rate of 30 ml/min (0.001 ft^3/min) for approximately a half hour to purge the oxygen in the environment of the furnace. The loaded sample was then heated in the nitrogen atmosphere after the TGA apparatus was stabilized. The sample was first heated from room temperature to 750º C (1,382º F) at a heating rate of 20º C/min (68º F/min) the nitrogen environment to drive off volatile materials, including water from dehydration, carbon dioxide from calcinations of calcium carbonate and low molecular-weight hydrocarbons. Once 750º C (1,382º F) was reached, the temperature was held constant until the TGA curve became flat so that the volatile materials were completely released. Finally, nitrogen flow was stopped and the same flow rate (30 ml/min or 0.001 ft^3/min) of air was admitted. The oxygen in the introduced air was used to oxidize carbon in char into carbon dioxide. The whole TGA test was completed when the height of the weight loss curve stopped changing.

Concentrations of other elements in chars, including Na, Mg, Al, Si, P, S, K, Ca, Mn, F, and Zn, were also evaluated using a Philips PW 2404 X-ray fluorescence spectrometer (Rh X-ray tube operated at 3,600 watts). Several steps need to be taken to conduct these tests. First, char samples were homogenized by mixing in a Spex Mixer/Grinder for 5 minutes.   Then, samples were also diluted 1:1 with lithium tetraborate. Standards were made by mixing NIST or Reagent grade standards with graphite or lithium tetraborate.

Mixtures were again homogenized by mixing in a Spex mixer/grinder for 5 more minutes. Specimens were presented to the spectrometer as loose powders. The powders were contained in disposable sample cups covered with polypropylene film. Scans were made using both the raw samples and the samples that had been diluted. All scans were performed in a Helium atmosphere.   The measured intensities were converted to concentration using the Philips IQ+ program (a fundamental parameters package). The measured intensity of one of the standards (Std 1) was used to check the procedure. Agreement was reasonable for all elements except for Iron (Fe).

3.4.1.2 *Preparation and characterization of activated carbon*

One hundred grams of either oat hull char or corn stover char were placed in a porcelain evaporating dish and heated to 800° C in a Lindberg box furnace equipped with a gas-tight retort (Lindberg/Blue M, Watertown, WI). The char was heated in an inert

atmosphere of nitrogen gas supplied to the retort at a flow rate of 0.1 m^3/hr (3.5 ft^3/hr). Once the furnace temperature reached 800° C (1,472° F), water was pumped into the nitrogen gas stream at a rate of 3 mL/min (0.0001 ft^3/min) in order to achieve steam activation. The chars were steam activated for 30, 60, 90, and 120 min. At the conclusion of these time periods, the water flow was stopped and the heat source shut down. The chars were allowed to cool in the furnace overnight under an atmosphere of nitrogen gas.

The activated carbons were weighed to determine activation burn-off or mass loss due to activation. The following relationship was used:

Activation burn-off (percent) = 100 – {[mass after activation (g)/original mass (g)] x 100}

The activated carbons were washed with 0.1 M HCl to remove ash and then washed with water to remove residual acid. In each case, the carbon particles were retained on a 325 mesh sieve. The samples were then dried at 102° C (216° F). A final product yield or loss due to washing was recorded.

Surface area and pore size distribution of the carbons was determined by nitrogen adsorption at 77° K (-321° F) using a Micromeritics Gemini 2375 surface area analyzer (Micromeritics Inc., Norcross, GA) and a 15-points BET analysis. Micropore surface area and external surface area (mesopore + macropore surface area) were calculated by use of a t-plot.

3.4.1.3 Results: concentrations of elements in chars

TGA tests show that the carbon contents in oat hull char and corn stover char are 72.3 weight-percent and 68.0 weight-percent, respectively.

The results of other elements measured with Philips PW 2404 X-ray fluorescence spectrometer are listed in Table 3-7.

Table 3-7. Concentrations of Several Elements (as Oxide Forms) in Oat Hull Char and Corn Stover Char.

Elements (as oxides)	Oat Hull (weight-percent)	Corn Stover (weight-percent)
Na_2O	0.22	<0.02
MgO	1.11	0.17
Al_2O_3	0.06	0.06
SiO_2	1.50	0.89
P_2O_5	1.55	0.07
SO_3	0.44	0.06
K_2O	2.54	1.79
CaO	0.33	1.67
Mn_2O_3	<0.02	0.13
Fe_2O_3	0.47	0.17
ZnO	0.1	<0.02

3.4.1.4 Results: characterization of activated carbon

The characterization results of activated carbon obtained from oat hull and corn stover are shown in Table 3-8 and Table 3-9, respectively.

Both activation burn-off and BET surface area appear to exhibit a linear relationship with respect to activation time of oat hulls (Table 3-8). Activated carbons with good surface area (>500 m^2/g or 2.4 Mft^2/lb) can be obtained with steam activation times of 90 min or longer. Total yields of activated carbon from the original mass of oat hulls need to be calculated from the char yield and the activation burn-offs.

Table 3-9 shows that for corn stover activated carbons, there is no linear relationship between activation time and BET surface area. However, activation burn-off and activation time appear to relate in a linear manner. BET surface area reaches a maximum at an activation time of 60 min, and then declines at higher activation times. This may be due to the fragile structure of the corn stover char. As more surface area is generated there is a breakdown of the char particles. This explanation would explain the high mass loss due to washing at activation times of 90 and 120 min and the decline in BET surface area at those activation times. Corn stover is not likely to be a good candidate for activated carbon production. On the other hand, oat hulls seem to be a good source material for activated carbon production.

Table 3-8. Activation Burn-Off, Loss Due to Washing, BET Surface Area and Pore Size Distribution of Oat Hull Activated Carbons.

Activation Time (minute)	Activation Burn-Off (percent)	Mass Loss Due to Washing (percent)	BET Surface Area (m^2/g)	Percent Micropores; Percent (Meso + Macropores)
0 (char)	none	none	4.8	100
30	20.8	9.2	349	91.4; 8.6
60	27.6	13.4	431	88.4; 11.6
90	33.5	10.5	522	86.8; 13.2
120	39.7	16.0	625	84.2; 15.8

Table 3-9. Activation Burn-Off, Loss Due to Washing, BET Surface Area and Pore Size Distribution of Corn Stover Activated Carbons.

Activation Time (minute)	Activation Burn-Off (percent)	Mass Loss Due to Washing (percent)	BET Surface Area (m^2/g)	Percent Micropores; Percent (Meso + Macropores)
0 (char)	none	none	5.8	100
30	26.2	18.7	424	88.7; 11.3
60	30.5	22.0	442	86.2; 13.8
90	37.5	28.8	374	85.0; 15.0
120	40.2	27.1	311	84.2; 15.8

3.4.2 Recovery of Organic Compounds from Bio-oil

The bio-oil contains a wide variety of organic compounds, as illustrated in Table 3-10 for corn stover. These include alcohols, aldehydes, carboxylic acids, ketones, saccharides, anhydrosugars, furans, and pyrolytic lignin. Of particular interest is the anhydrosugars, as detailed below.

Table 3-10. Yields of Organic Compounds in Pyrolysis Syrups for Pretreatments of Corn Stover.				
Organics (Weight Percent)	No Pretreatment	Acid Hydrolysis	Demineraliza-tion	Demineralization with catalyst
Cellobiosan	Trace	4.97	4.55	3.34
Levoglucosan	2.75	23.10	17.69	20.12
Hydroxy-acetaldehyde	11.57	3.93	5.97	3.73
Formic acid	2.61	0.73	trace	Trace
Acetic acid	3.40	0.40	1.51	1.26
Acetol	4.53	trace	trace	Trace
Formaldehyde	2.75	0.70	1.63	Trace
Pyrolytic lignin	33.40	20.08	16.89	17.74

Among the various components in bio-oil (or BCO) those with a reasonable vapor pressure include:

- Low molecular weight carbonyl compounds
- Water
- Monomeric sugars
- Monomeric phenols

Other components are high molecular weight oligomeric phenolic fragments ("pyrolytic lignin") and oligomeric sugars and their partial decomposition products. Some researchers believe that this oligomeric material is not formed by vaporization but rather by direct ejection into the gas phase as aerosol particles and that this is partially why BCO shows irreversibility of distillation.

The volatility characteristics of the lower boiling components may be inferred from their known or estimated boiling points.

3.4.2.1 Monomeric phenols

Monomeric phenols (excluding phenol itself) derived from lignin include compounds with one or two methoxy groups ortho to the phenol group together with side chains of from 0 – 3 carbons, para to the phenol group. Their molecular weights range from 124 (guaiacol) up to about 210 (sinapyl alcohol). Boiling points are given in Table 3-11.

Table 3-11. Boiling Points of Some Monomeric Lignin-Related Phenols.

Monomeric Lignin-Related Phenols	Experimental ° C (° F)	Estimated[*] ° C (° F)
Phenol	182 (360)	-
Guaiacol	205 (401)	211 (412)
Methyl Guaiacol	-	230 (446)
Vinyl Guaiacol	-	247 (477)
Syringol	-	249 (480)
Eugenol	254 (489)	-
IsoEugenol	-	270 (518)
Vanillin	170 @ 15 mmHg (338 @ 0.3 psi)	274 (525)
AcetoGuaiacone	-	281 (538)
Syringaldehyde	193 @ 14 mmHg (379 @ 0.3 psi)	305 (581)
Coniferyl Alcohol	164 @ 3 mmHg (327 @ 0.06 psi)	325 (617)
Sinapyl Alcohol	-	350 (662)
Ferulic Acid	-	354 (669)

Note:
[*] Estimated using Stein and Brown method (Stein and Brown, 1994).

Note that the yields of individual compounds are small (0.5 - 2 percent each) and it would perhaps be simplest to collect impure fractions and to vacuum distil if higher purity was required.

Some of these phenols are quite valuable, ~ tens of dollars/kg, as flavors or flavor/pharmaceutical precursors and other specialty chemicals. For example, ferulic acid, coniferyl alcohol, 4-vinyl guaiacol and eugenol can all be fermented to valuable natural vanillin. On the other hand, phenol itself sells at about $1/kg ($0.5/lb). Given the present interest in BCO derived ("green") resin formulations, it seems likely that impure mixtures of monomeric phenols could command something in this price range.

3.4.2.2 Levoglucosan

Levoglucosan is an anhydrosugar with only limited commercial production at present but having broad potential as a platform molecule for green chemistry. A recent review by Radlein (2000) suggests its use in applications ranging from production of fermentation products (lactic acid, citric acid, etc.) to pharmaceuticals. However,

levoglucosan has been little studied for commercial applications because technical grade levoglucosan costs as much as \$50/g (\$22,680/lb). Pyrolysis of raw biomass typically yields only a few percent levoglucosan. However, studies by several researchers reveal that pyrolytic yields can be dramatically increased to as much as 30 percent by pretreating biomass to remove alkaline compounds (primarily potassium). Pretreatment is very simple, consisting of washing the biomass in weak acid to remove water-soluble alkali (this pretreatment also mitigates the previously described problem of alkali corrosion and erosion in gas turbines).

The principal monomeric anhydrosugar is levoglucosan (1,6 anhydro-β-D-glucose). Its yield is generally in the range 3 - 10 weight percent for untreated biomass and up to 17 percent for demineralized biomass. E. Suuberg et al (1996) has measured the vapor pressure of levoglucosan as:

$$\ln P \text{ [torr]} = 32.391 - 14452/T \text{ [K]}$$

indicating a boiling point of about 288° C (550º F). (Compare glycerol which also has three hydroxyl groups and which boils – with decomposition – at 290º C or 554º F.) The estimate therefore seems reasonable − especially in view of the fact that the hydroxyl groups of levoglucosan are rigidly held and presumably therefore less able to form strong intermolecular hydrogen bonds that decrease volatility. For comparison, the Stein and Brown method (Stein and Brown, 1994) gives a boiling point of 314° C (597º F). For the purposes of separation, it should also be noted that levoglucosan and other oligomeric sugars are water-soluble while most of the pyrolytic lignin is not.

3.4.2.3 C_1 - C_3 carbonyl compounds

The predominant low molecular weight carbonyls include formaldehyde, glycolaldehyde, glyoxal, acetol and acetic acid. The aldehydes are characterized by a high affinity for water with the formation of hydrates of much reduced volatility. (This fact is also a contributor to the irreversibility in the distillation of BCO.) Examples of boiling points are given in Table 3-12.

The aldehydes are also very reactive. For example, anhydrous formaldehyde is a gas at room temperature but in the presence of liquid water it will exist mainly in solution as a hydrate even at elevated temperatures.

Table 3-12. Boiling Points of Some Low Molecular Weight Carbonyls.		
Low Molecular Weight Carbonyls	**Experimental** $^\circ$ C ($^\circ$ F)	**Estimated**[*] $^\circ$ C ($^\circ$ F)
Formaldehyde	-19.5 (-3)	-
Methylene Glycol (Formaldehyde Hydrate)	145[**] (293)	131 (268)
Glycolaldehyde	-	126 (259)
Glycolaldehyde Hydrate	-	205 (401)
Acetic Acid	118 (244)	-
Acetol	146 (295)	-
Glyoxal	51 (124)	-
Glyoxal Dihydrate	-	252 (486)
Notes: [*] Estimated using Stein and Brown method (Stein and Brown, 1994). [**] Data taken from Albert et al, AIChE J., 42, 1741, 1996.		

It is therefore exceedingly difficult to predict the likely boiling points of condensates from this fraction. The kinetics of the various reactions that they can undergo will play a major role. Nevertheless, collectively these constitute the lowest boiling fraction in the bio-oil.

For untreated biomass feedstocks the total amount of these compounds is of the order of 10 - 20 weight percent of the bio-oil and they arise mainly from the thermal decomposition of the holocellulose fraction. On the other hand, the available data suggest that if the feedstock is demineralized to enhance levoglucosan yields, their total amount falls significantly to only a few percent. The decision as to whether to attempt to recover a low molecular carbonyls fraction would be determined by its economic value.

3.4.3 Utilization of Levoglucosan

3.4.3.1 Ethanol production

Radlein (2000) has discussed applications of levoglucosan as a fermentation feedstock recently in some detail. It was pointed out that naturally occurring microorganisms have been discovered that can utilize levoglucosan as a substrate. For example, Kitamura et al (1991) have studied the metabolism of levoglucosan. They refer to prior work indicating that many yeasts and fungi can use it as a carbon source and suggest that for eucaryotic microorganisms (i.e. yeasts and fungi) capable of utilizing

levoglucosan, the first step involves phosphorylation to glucose-6-phosphate. The responsible enzyme, termed "levoglucosan kinase" was isolated. In earlier work they also showed that Aspergillus terreus converts levoglucosan to itaconic acid with the same yield and rate as in the conversion of glucose. Subsequently (Nakahara et al, 1994) a new organism, Arthrobacter sp., isolated from soil was found to hydrolyze levoglucosan to glucose by a novel enzyme, levoglucosan dehydrogenase. Thus, somewhat surprisingly perhaps, levoglucosan specific enzymes do occur in nature; a fact that should no doubt have a strong bearing on the feasibility of genetic modification of organisms for levoglucosan specific activity.

Indeed an old German patent (Schuchardt, 1943) suggests that levoglucosan or wood carbonization liquors containing it can be used as a nutrient in the production of yeast, e.g. torula utilis. Blazej and Kosik (1993) reported that aureobasidium pollulans CCI-2771-14 showed good growth rates using anhydrosugars as a single carbon source on which basis they suggested that direct conversion of anhydrosugars to fodder yeast has great potential.

Prosen et al, (1993) have also identified several yeast species that grow on, albeit slowly, and produce ethanol from aqueous pyrolysates extract containing levoglucosan and other anhydrosugars.

Thus the available information is that there exist microorganisms that convert levoglucosan and other anhydrosugars directly to various products. However ethanol production seems to be the only existing large scale fermentative application that offers immediate markets. Unfortunately direct fermentation of anhydrosugars to ethanol is slow. No doubt specific development of organisms for this purpose would give dramatic improvements. Nevertheless it is clear that at the present state of the art, the anhydrosugars must first he hydrolysed to free sugars then fermented. It is therefore imperative that a cheap and efficient hydrolysis method be available.

The most obvious approach is acid hydrolysis that is well known to convert anhydrosugars to free sugars. The kinetics have been studied by Ménard et al (1984). Using ~3N H_2SO_4, hydrolysis of pure levoglucosan was essentially complete in less than 30 minutes at 100° C (212° F). At the same temperature, complete conversion required about 3 hour. using 0.3 N acid.

Katritzky et al (1996) have described an even cheaper and simpler method that requires no acid. They report that a 0.30 Molar aqueous solution of LG was quantitatively converted to predominantly glucose and traces of another glucose isomer at 205° C (401° F). This result is probably because liquid water intrinsically becomes both more acidic and more basic at elevated temperatures so besides the normal thermal acceleration there is an enhanced catalytic effect. At 205° C (401° F), the vapor pressure

of pure water is 1,906 kPa (276 psi) so the required conditions are relatively mild and thus, such a process step would be expected to be rather inexpensive. Unfortunately, kinetic data, which is required for equipment sizing, were not given so a precise cost estimate cannot be made. Nevertheless, it requires only the single extra unit operation of pumping an aqueous solution through a pressurized and heated tube.

3.4.3.2 Higher value uses

Although production of fermentable sugars is the most immediate application, there are other potential higher value uses for levoglucosan. These have also been discussed by Radlein (2000). We are interested here in possible large-scale usage so that it might be compatible with the scale of energy production. In the author's view, the only presently viable potential candidate is the production of biodegradable and "renewable" alkyl glycoside surfactants. These are premium products that find current application in soaps and cosmetics. They are presently made by condensation reactions of sugars with vegetable oil derived alcohols or acids.

A major issue with the current method is that the reactions are generally non-selective so that purification of the product is a difficult and expensive operation. On the other hand, preliminary experiments have indicated that levoglucosan can be combined by direct addition reaction with alkanols under non-aqueous conditions to give alkyl glucosides in almost stoichiometric yield (D. Radlein, unpublished data), as shown in Figure 3-10.
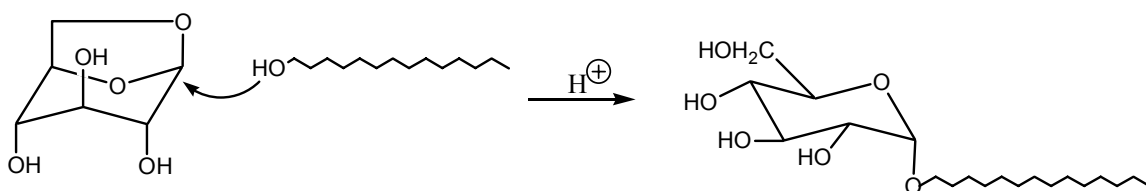
Figure 3-10. Addition Reaction of Levoglucosan with an Alkanol to give an Alkyl Glucoside.

Such an application would require pure levoglucosan. It is difficult to estimate a value for such an application. However, given that prices of alkyl glucosides appear to be in the range $1,500 - 2000/ton, and considering that pure sugars costing > $1/kg ($0.5/lb) are the normal starting materials for alkyl glycoside synthesis, while considering the advantages that a method giving a pure product might provide, it seems reasonable to

estimate that such an application could generate a price of at least $1 - 2/kg ($0.5 - 1/lb) for pure levoglucosan. The degree of product purity required would be higher though.

3.4.4 Economics of Levoglucosan Recovery from Bio-oil Production

Although activated carbon and several organic compounds are potential coproducts from the production of bio-oil, only the economics of recovering levoglucosan are reviewed here. Although economically attractive, entry into the activated carbon market is purportedly very difficult to achieve. Levoglucosan, among the organic compounds in bio-oil, occurs in the highest yield (when the raw biomass is demineralized) and has the most potential as a platform chemical for high-value product development.

No commercially available system for obtaining anhydrosugar from bio-oil is available at this time. Standard cost-estimating methods were used to estimate the cost of such a system. The system consists of two subsystems: a pretreatment system to demineralize the raw biomass and a fractionating quencher described in the previous section.

The pretreatment system, illustrated in Figure 3-11, consists of: a hammermill to reduce biomass to fibers of lengths of 1 – 3 mm (0.003- 0.01 ft); a washing unit to remove alkali from the biomass; a press to remove wash water from the biomass; and a dryer to bring moisture to 10 – 20 percent. The hammermill and press were designed to process 5,390 kg/h (11,883 lb/h) of biomass. The washing unit was sized to 10.6 m^3 (373 ft^3), which provides a residence time of 15 minutes for biomass with a bulk density of 130 kg/m^3 (8.1 lb/ft^3). The dryer was designed to give about 100 s residence time to the biomass and is estimated to require 2.0 m^2 of cross-sectional area.

The estimated f.o.b. costs for these various components and the assumptions made in deriving these estimates are detailed in Table 3-13. Bare-module costs for installed equipment, also given in Table 3-13, were calculated according to the methodology of Table 3-4. The total cost of the pretreatment system is estimated to be $0.72 million.

The major operating cost for the pretreatment system is acid to demineralize the biomass. Very little data is available on the optimum amount of acid for this purpose. Based on data on the use of steep water in the wet corn milling industry, it was assumed that 0.75 kg (1.6 lb) of acid solution (20 weight-percent nitric acid) was required for every kilogram of biomass. This equates to 29 million kg/yr of acid. At $77/Mg, the annual cost of acid is $2.3 million.
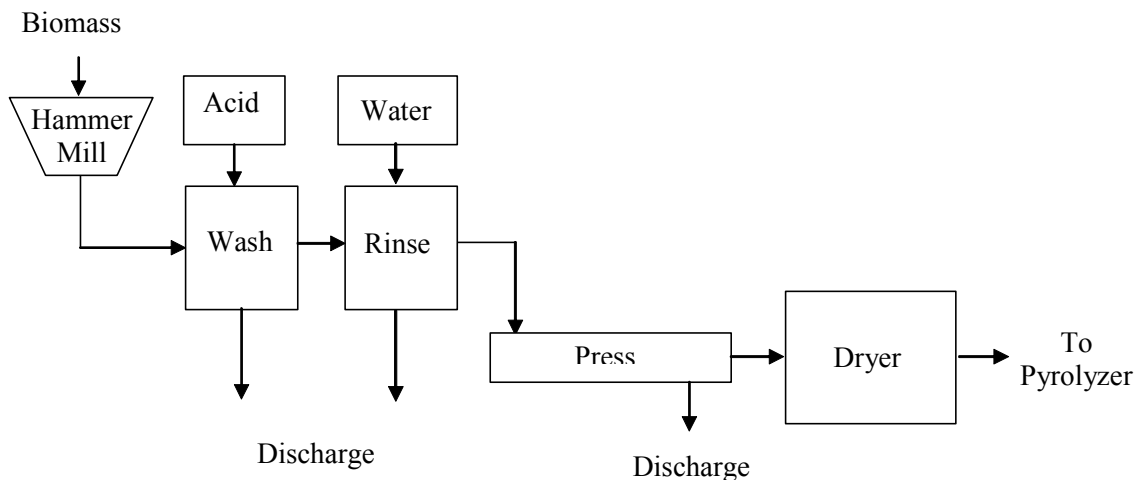
Figure 3-11. Biomass Washing and Drying System for Demineralizing Biomass.

Table 3-13. Cost Estimate for Biomass Pretreatment System.			
Component	**Cost f.o.b.**	**Scaling Relationship[*]**	**Bare Module Cost**
Hammermill	$11,720	$2,800(ton per hour)$^{0.85}$	$33,510
Wash vessel	$25,560	$6,200(m^3)$^{0.6}$	$92,640
Press	$224,000	Ulrich, 1984	$542,000
Dryer	$21,400	$16,000(m^3)$^{0.42}$	$59,100
		TOTAL	$727,250
Note: [*] Based on inflation-corrected correlations from Guthrie (1969) unless otherwise noted.			

A cost estimate was also prepared for the fractionating quencher and related separation equipment to extract levoglucosan from bio-oil. The proposed design is shown in Figure 3-8. The following major components are included in the design:

- Fractionating Column
- Heat Exchanger (Condenser)
- Separator Drum
- Hydrolyser
- Sugar Solution Storage Tank
- Pumps

The capital cost estimating approach for the fractionating condenser is detailed in Appendix A.   The estimated capital cost is $545,480.   Operating costs for chemical recovery were estimated at $1.0 million.

Table 3-14 summarizes the capital costs and operating costs estimated for the recovery of anhydrosugar from pyrolyzed biomass. The market price for levoglucosan is difficult to estimate because it is not currently produced in commercial quantities. However, other sugar derivatives, such as sorbitol, sell for as much as $1/kg ($0.5/lb). With an annual production of 5.8 x 10^6 kg/y (12.8 x 10^6 lb/y), this represents annual revenue of $5.8 million.

Table 3-14. Summary of Capital Costs and Operating Costs for Recovering Levoglucosan from Pyrolyzed Biomass.		
Subsystem	**Capital Cost (million)**	**Annual Operating Cost (million)**
Pretreatment	$0.73	$2.3
Anhydrosugar separation	$0.55	$1.0
TOTAL	$1.27	$3.3

3.4.5 *Comparative Economics of Levoglucosan for Ethanol Production*

In order to evaluate the economics of utilization of levoglucosan as a source for ethanol, the manner in which it could potentially integrate into the existing infrastructure must be examined.

Current large-scale (fuel) ethanol production in North America is largely from corn. Dry corn typically contains about 61 percent of starch that can be converted to fermentable sugars. Generally the corn is ground by either wet-milling or dry-milling following which the starch is hydrolysed in a first stage that reduces the starch to oligomeric fragments and in a second stage that converts the oligomers to free mono-molecular sugars.   Both acidic as well as enzymatic hydrolyses are used for these purposes. The solution of free sugars is then fermented to ethanol. In present practice, the overall efficiency of conversion of starch to ethanol is very high – well over 90 percent of the theoretical yield is attainable.   Indeed while typical yields are around 9.5 L ethanol/bushel of corn (2.5 gal/bushel of corn), as high as 11 L/bushel of corn (2.8 gal/bushel) have been achieved with recent technology.

Since the fermentable sugar solution is not normally separated out as a final product, it is difficult to estimate a precise cost for it, without a detailed process analysis. However, data in Table 3-15, obtained from Program Evaluation Division, Office of the

Legislative Auditor, Minnesota State Government: (http://www.auditor.leg.state.mn.us/ped/pedrep/9704toc.html), may provide some guidance.

Table 3-15. Costs of Dry Milling Corn in Minnesota	
	$/L of Ethanol ($/gal)
Costs	
Corn (1988 - 95 average)	0.23 (0.88)
Variable costs, except for corn	0.10 (0.37)
Fixed costs	0.08 (0.29)
Total	0.41 (1.54)
Revenues	
Ethanol (1988 – 95 average price)	0.34 (1.27)
Average DDGS[*] revenue	0.11 (0.42)
Minnesota state subsidy	0.05 (0.20)
Total	0.50 (1.89)
Note: [*] DDGS = Distillers Dry Grain with Solubles (fermentation residue principally used for animal feed).	

The current cost of corn is approximately $2/dry bushel. Since the starch content of the corn is about 61 percent, the effective cost of starch for fermentation works out at $0.116/kg ($0.005/lb) starch. As mentioned above it is not so easy to separate out the cost of the two hydrolytic steps involved in the conversion of the starch to fermentable sugar solution, but it must be less than the total non-corn variable costs of $0.10/L of ethanol ($0.37/gal) as shown in Table 3-15.

One gallon ethanol is equivalent to 6.2 kg starch at the conversion rate of 2.5 gal/bushel. Assuming hydrolysis to be a quarter of non-corn variable costs gives an estimate of only $0.012/kg starch. Thus we estimate the total production cost of fermentable sugar solution from corn at $0.128/kg of dry sugars. This is therefore the target price at which "thermo – saccharification" becomes competitive with cornstarch for ethanol production.

As a check against these numbers one may compare current prices for food grade glucose syrups and high fructose corn syrups (HFCS):

Glucose Syrup (19.7 percent moisture) $0.352/kg dry sugars
                                         (Wholesale list price, MidWest markets)

HFCS – 42 (29 percent moisture) $0.286/kg dry sugars
                                         (MidWest spot price, July '02)

They appear to be compatible with our estimate for cost estimate for fermentative sugars in solution in that they are more refined products.

In order to justify the production of such solutions from levoglucosan rich pyrolysates within the power production system under study, one must compare with the alternative product, namely electrical power. There is even more uncertainty here as electrical energy prices vary very considerably with location and season. However, as an example one might consider that residential the electrical price in Iowa was $0.081/kWh in 2001. (See: http://www.blueowltechnologies.com/pmcPricesUSA.asp)

A part of this price represents distribution and transmission charges. Assume that the actual energy commodity price is 60 percent of the total cost then we estimate the former at ~$0.050/kWh.

Now assume that the pyrolysis liquids have a heating value of 17 MJ/kg (7,309 Btu/lb) and that this is converted to electrical power at 30 percent efficiency. Then the value of the electrical energy product is about $71/ton ($0.03/lb) pyrolysis liquids. On the other hand, the value of that fraction of the pyrolysis liquids that can be converted to fermentable sugars is $128/ton ($0.06/lb) of the convertible fraction of pyrolysis liquid – approximately double the value per ton than power production.

More detailed comparison of the production of levogluasan versus the production of electrical power is provided in Section 5, Economic Analysis.

3.5 Gas Turbine Cycle Design

There are a variety of energy conversion options for the bio-oil produced by the pyrolysis process. These include boilers, internal combustion engines, combustion or "gas" turbines, and perhaps advanced devices including Stirling engines. Each of these devices has its own advantages and disadvantages. However, the low cost, high efficiency, established commercial maturity, and suitability for recovery of waste heat in a combined cycle makes combustion turbines the preferred alternative for this study.

This section provides an introduction to gas turbine technology, properties of bio-oil relevant to gas turbines, gas turbine options, and a description of the turbine proposed for this project, the Orenda Aerospace OGT2500.

3.5.1 Introduction to Gas Turbine Technology

The first successful gas turbine was completed in 1903. Over the next forty years, rapid advances were made to improve the technology to make it a viable means of aircraft propulsion. As the technology matured, gas turbines were adapted to land-based energy generation uses. With the deregulation of the power industry in the 1990s, gas

turbines became the generator of choice for a vast majority of all new power projects. Gas turbines currently have lower capital costs, shorter construction durations and lower operation and maintenance costs than any other large central plant available on the market.
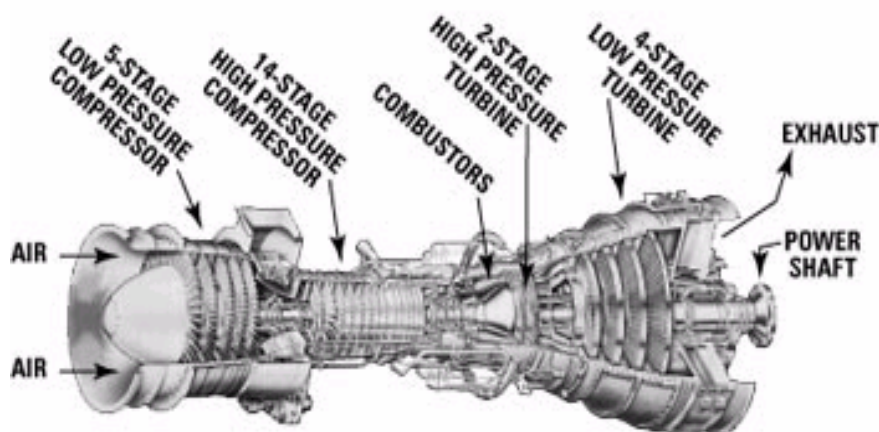
Figure 3-12. Gas Turbine Section (Source: Langston).

Power is generated when the combustion turbine compresses ambient air to approximately 12 to 16 atmospheres, heats the pressurized air to 1,100°C (2,000°F) or more by burning oil, natural gas or other fuels, and then expands those hot gases through a turbine. The turbine then drives both the air compressor and an electric generator. A typical combustion turbine would convert 30 to 35 percent of the fuel energy to electric power, with a substantial portion of the fuel energy exhausted in the form of hot gases exiting the turbine. When the combustion turbine is used to generate power and no energy is captured from the hot exhaust gasses, the power cycle is referred to as a "simple cycle" power plant.

Simple cycle combustion turbines are the power generation technology of choice for peaking service in the current domestic power industry. Simple cycle technology provides many of the same positive attributes as reciprocating engines, including rapid startup and modularity for ease of maintenance. In addition, combustion turbines have several advantages over reciprocating engines, including lower emissions and lower capital cost.

Simple cycle turbines can burn natural gas, diesel, propane, biogas and some bioderivative fuels such as bio-diesel and bio-oil. The next section describes some of the bio-oil properties relevant to gas turbines.

3.5.2 Bio-Oil Properties Relevant to Gas Turbine Engines

As a clean fuel, bio-oil has a number of environmental advantages over fossil fuels:

- CO2 / Greenhouse Gas Neutral – because bio-oil is derived from biomass (organic waste), it is considered to be greenhouse gas neutral and can generate carbon dioxide credits.

- No SOx Emissions – As biomass does not contain sulfur, bio-oil produces virtually no SOx emissions and, therefore, would not be subjected to SOx taxes.

- Low NOx – Bio-oil fuels generate more than 50 percent lower NOx emissions than diesel oil in gas turbines.

- Renewable and Locally Produced – Bio-oil can be produced in countries where there are large volumes of organic waste.

As bio-oil has unique properties as a fuel, it requires special consideration and design modifications. Some of these properties are presented in Table 3-16 and are compared to those of diesel fuel.

Table 3-16. Typical Properties of Bio-Oil Compared to Diesel Fuel.		
	Bio-Oil	**Diesel**
Calorific Value MJ/kg	15-20	42.0
Kinematic Viscosity cSt	3-9 @ 80°C	2-4 @ 20°C
Acidity pH	2.3-3.3	5
Water wt%	20-25	0.05 v% (combined)
Solids wt%	< 0.1	
Ash wt%	< 0.02	0.01
Alkali (Na + K) ppm	5-100	< 1

3.5.3 Combustion Turbine Options

A number of manufacturers offer combustion turbine systems in the size range appropriate for this study. Table 3-17 provides key characteristics for small combustion turbines. It must be noted that not all of these turbines are suitable for bio-oil use. The next section discusses criteria for selecting an appropriate combustion turbine.

Table 3-17. Small Gas Turbine Characteristics.						
Manufacturer	**CT Model**	**Base Load Output[2] kW**	**Heat Rate[2] Btu/kWh**	**Efficiency[2] Percent**	**Budget Price[1] US$**	**Budget Price[1] US$/kW**
Dresser-Rand	KG2-3E	1,830	21,070	16	1,200,000	656
Pratt & Whitney	ST18A	1,960	11,300	30	1,200,000	612
Orenda Aerospace	OGT2500	2,730	12,515	27	1,435,000	526
Mashproekt	UGT-2500	2,850	12,430	28	1,390,000	488
Kawasaki Heavy Industries	M1T13D	2,900	14,460	24	1,625,000	560
Vericor Power Systems	VPS3	3,105	12,775	27	1,520,000	490
Pratt & Whitney	ST30	3,340	10,660	32	1,600,000	479
Solar Turbine	Centaur 40	3,515	12,240	28	1,400,000	398
Vericor Power Systems	VPS4	3,570	11,800	29	1,601,000	448
Rolls-Royce	501-KB5S	3,950	11,765	29	1,600,000	405
Iskra Energetika	GTES-4	4,100	14,130	24	1,230,000	300
Pratt & Whitney	ST40	4,040	10,310	33	1,800,000	446
Solar Turbine	Centaur 50	4,600	11,630	29	1,600,000	348
Iskra Energetika	GTES-5	5,200	13,050	26	1,534,000	295
Solar Turbine	Taurus 60	5,200	11,225	30	1,800,000	346
GE Oil & Gas	PGT5	5,220	12,720	27	1,900,000	364
Alstom Power	Typhoon 5.25	5,250	11,200	31	1,850,000	352
Rolls-Royce	501-KB7	5,275	11,200	31	1,750,000	332
Kawasaki Heavy Industries	M7A-01	5,840	11,230	30	2,310,000	396
GE Oil & Gas	PGT5B	5,900	10,700	32	2,050,000	347
Iskra Energetika	GTES-6	6,200	12,780	27	1,705,000	275
Rolls-Royce	501-KH5 (Steam Injection)	6,420	8,560	40	2,300,000	358
Rolls-Royce	601-KB9	6,450	10,615	32	2,450,000	380

Source: Gas Turbine World 2001-2002 Handbook.

[1.] Budget prices in 2001US$ for basic electric power generator packages including a single-fuel gas turbine, air cooled electric generator, skid and enclosure, inlet and exhaust ducts with silencers, standard control and starting systems, and conventional combustion system.

[2.] Based on ISO conditions of 15º C (59º F) at sea level site conditions and 60 percent relative humidity with zero inlet and zero outlet duct losses.

3.5.4 OGT2500 Combustion Turbine

DynaMotive and Orenda Aerospace have cooperated on development of the 2.5 MW OGT2500 industrial gas turbine engine for applications with bio-oil. Based on promising results from testing completed so far, this turbine (see Figure 3-13) appears to be a viable candidate for bio-oil utilization. The project design includes two OGT2500 turbines. Table 3-18 shows the predicted OGT2500 performance on bio-oil.
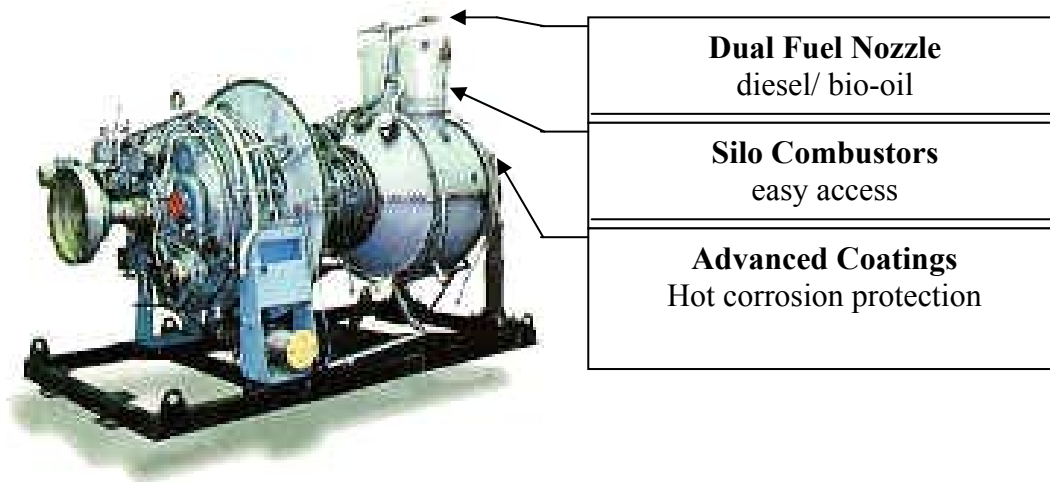
Figure 3-13. Application of Pyrolysis Oil to Gas Turbine Operation (OGT2500).

Table 3-18. OGT2500 Predicted Performance on Bio-oil.		
	Metric Units	**English Units**
Ambient Conditions		
Barometric Pressure	101.4 kPa	14.7 psia
Dry Bulb Temperature	15.0 °C	59 °F
Relative Humidity	60%	60%
Wet Bulb Temperature	10.81 °C	51.47 °F
Humidity Ratio	0.63%	0.63%
Fuel Parameters		
Fuel Flow	2,073 kG/hr	4,569 lb/hr
Heat Input (LHV)	35.5 GJ/hr	33.6 MBtu/hr
Exhaust Properties		
Mass Flow Rate (total)	53,649 kg/hr	118,275 lb/hr
Temperature	453 °C	848 °F
Generator Gross Output, (each)	2,642 kW	2,642 kW

The OGT2500 offers distinct technical advantages over other engines. Unlike aero-derivative engines, it has been designed as an industrial engine with durability being one of the main design criteria and not weight. In addition to its ruggedness, the distinct "silo" type combustion system allows for easy access and modifications to the entire combustion system, which is one of the critical systems for the adaptation of the engine to bio-oil.

In addition to the engine design, important design modifications are necessary to compensate for the unique properties of bio-oil. These modifications will be discussed in order and based on the bio-oil properties listed in Table 3-16.

Bio-oil has an energy density approximately half that of diesel fuel. Therefore, to meet the same energy input requirement, the flow rate must be double. This requires design changes to the fuel system to be able to control higher flow rates and also modify the fuel nozzle to accommodate this larger flow. This lower energy density also can affect combustion since physically there must be twice as much fuel in the combustion chamber as with diesel. This, however, is another advantage of using an industrial engine in the fact that the combustion chambers are designed with a significantly longer residence time (and therefore a larger volume) for a given power output.

Higher viscosity of the fuel reduces the efficiency of atomization, which is critical to complete combustion. Large droplets take too long to burn. Proper atomization is addressed in three ways. Firstly, the fuel system is designed to deliver a high-pressure flow since atomization is improved with larger pressure drops across the fuel nozzle. Secondly, the fuel is pre-heated to lower the viscosity to acceptable levels. Thirdly and most importantly, the fuel nozzle has been redesigned to improve spray characteristics. These design improvements are important for complete combustion and effectively reducing CO emissions.

Due to its relatively low pH, material selection is also critical for all components wetted by BioOil. This does not require the use of exotic materials; however, it does eliminate some standard fuel system materials. Typically, 300 series stainless steels are acceptable metallic materials and high-density polyethylene (HDPE) or fluorinated HDPE for polymers.

Although looked at as a contaminant for diesel fuel, the water content in bio-oil has some advantages. Firstly, it is helpful in reducing the viscosity, since it is a relatively low viscosity fluid. As well, it is a factor in lowering thermal NO_x emissions.

The solids content is a combination of ash and char fines which have carried-over to the liquid part of the bio-oil. The effect of these solids is to cause sticking of close tolerance surfaces. They can result in particulate emissions because of the long residence

time required to fully combust. It is important that the solids level in the bio-oil is controlled to be less than 0.1 wt%.

The ash content in the fuel represents the material that cannot be combusted. Depending on the elements in the ash, it can result as a deposit on the hot gas path components that will reduce the turbine efficiency. This operational problem is a familiar one with the use of low-grade fuel oils that also have high ash content. The solution is a turbine wash system. This typically consists of two separate systems in which an abrasive medium is injected during operation to physically 'scrub' off the deposits. This allows turbine cleaning without any downtime. The second system is an offline process, which injects a cleaning fluid and allows a soak period to loosen the deposits that are then removed when the engine is started.

Within the ash are alkali elements that can result in hot corrosion of the hot gas path components with sodium and potassium being the most critical elements found in bio-oil. These elements form low melting temperature compounds that, as a liquid, will stick to the hot gas path components and then react and corrode the component. This effect can be mitigated through the use of fuel additives. As with the turbine wash systems, this technology was developed for the use of heavy fuel oils in gas turbines and has been in use for decades. The concept is to inject specific elements that preferentially react with the alkali metals such that they do not liquefy. This both reduces the propensity to stick to a surface and also reduces or completely eliminates its rate of attack. In combination with the additives, hot section coatings are being developed specifically for the type of attack that may be associated with bio-oil.

Due to the poor ignition characteristics of bio-oil, one other key design requirement is a bio-oil specific ignition system or process. To overcome this, the OGT2500 system starts on diesel fuel flowing through the primary channel in the fuel nozzle. Following a warm-up period, bio-oil is fed into the secondary channel at an increasing rate while the diesel fuel flow is reduced until 100 percent bio-oil flow is achieved.

Polymerization is also a key issue with bio-oil. This is the growing of molecular chains that can result in an increase in fuel viscosity. This process is highly dependent on the char level, time and temperature. For example, the equivalent change in properties can be achieved in 6 months at room temperature, compared to 8 hours at 90°C (R.G. Andrews et al., 1997). Therefore, as part of the fuel and combustion system design, maximum temperatures and fuel re-circulating are carefully controlled to ensure polymerization is maintained at a rate that is inconsequential to engine operation.

3.5.5 Combustion Turbine Testing

DynaMotive and Orenda Aerospace have been cooperating on development of the OGT2500 for applications with bio-oil. A first generation fuel system and combustion system were designed and tested, demonstrating the capability to operate a 2.5 MW industrial gas turbine on bio-oil (P. Gogolek et al., 1999). These tests not only revealed the feasibility of operation but also demonstrated that similar performance could be achieved for bio-oil and diesel. Although CO and particulate emissions were comparable to diesel, testing revealed that NO_x emissions were about half that from diesel fuel and the SO_2 emissions levels were so low as to be undetectable by the instrumentation.
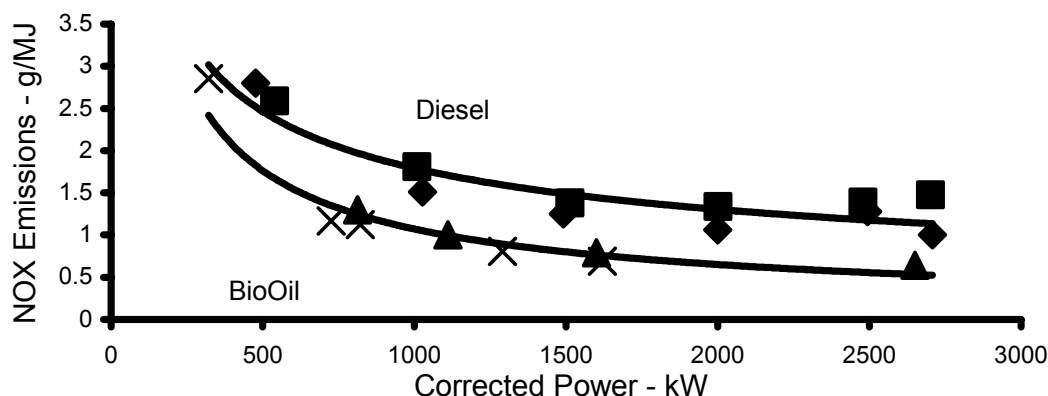
Figure 3-14. Pyrolysis Oil NO_x Emission Reduction.

3.6 Rankine Steam Cycle Design

By capturing the energy of the hot exhaust from the combustion turbine, additional power can be generated in a combined cycle configuration. This section discusses the steam or "Rankine" cycle portion of the combined cycle. This section also provides a heat and mass balance for the overall combined cycle and defines efficiency measures for the IPCC power plant.

3.6.1 Rankine Steam Cycle Overview

In comparison to other portions of the IPPC power plant, the Rankine steam cycle is relatively conventional technology with minimal technical risk. The steam cycle recovers the energy in the hot gases of the combustion turbine thereby raising steam in a heat recovery steam generator (HRSG). The steam from the HRSG is used to rotate a steam turbine generator and produce electrical power. The exhaust from the steam turbine is cooled until the water condenses, whereupon it is returned to the HRSG to be heated again. Heat is rejected from the condenser via a cooling tower. An alternative

approach would be to send the turbine exhaust steam to a district heating loop or an industrial steam user. In this manner the heat could be put to a productive use thereby raising cycle efficiency. However, such applications are site specific and were not evaluated in this general study.

3.6.2 Design Approach

Early configurations of the Rankine steam cycle were designed to maximize efficiency and power output. Items such as a dual pressure level HRSG, high steam pressures and temperatures, and mid-stage steam admission to the steam turbine were included. These features resulted in a combined cycle design of over 41 percent efficiency with a net power block output of 8.43 MW. However, at the small sizes contemplated here, these design choices are non-standard and often result in significant cost premiums versus more conventional designs. For example, indicative estimates from vendors showed that a steam turbine with a throttle pressure of 6,200 kPa (900 psia) costs nearly three times as much as a turbine with a 4,140 kPa (600 psia) throttle pressure ($2.5 million compared to $850,000). Further, this cost differential does not include the additional costs for the higher pressure HRSG, pumps, piping, and other balance of plant equipment. Based on these considerations, a compromise design was selected that balances efficiency with cost. These changes resulted in a cycle design of about 37 percent efficiency with a net combined cycle power block output of 7.66 MW. Even at this reduced efficiency, the design is much more efficient than conventional biomass power plants, which often struggle to meet 20 percent efficiency in this size range.

3.6.3 Combined Cycle Heat and Mass Balance

Based on the design choices selected above, a heat and mass balance of the combined cycle portion of the plant was prepared using the GateCycle software. A summary of this heat balance is presented in Table 3-19. Based on the GateCycle model, the net power output from the combined cycle portion of the plant is estimated to be 7,660 kW. The heat input to the gas turbines is 70.9 GJ/h (67.2 MBtu/h) on a lower heating value basis (LHV). A small amount of additional heat is added to the cycle from the exhaust of the pyrolysis plant (the pyrolyzer indirect heating gas stream). This amounts to 2.6 GJ/h (2.5 MBtu/h). With a total heat input of 73.6 GJ/hr (69.8 MBtu/h) (LHV), the estimated combined cycle net plant heat rate (LHV) is 9,610 kJ/kWh (9,110 Btu/kWh). This is equivalent to 37.5 percent efficiency. This efficiency measures the combined cycle portion of the project only and does not include the efficiency of the pyrolyzer in converting the solid biomass fuel into bio-oil. This is discussed further in the next section.

Table 3-19. Combined Cycle Heat and Mass Balance.

Fuel		Bio-Oil	Bio-Oil
Ambient Temperature		15 C	59 F
Number of CTG/HRSG Units Operating		2	2
STG Throttle Conditions		4140/340 kPa/C	600/650 psia/F
Condenser Pressure		103.429 mmHgA	4.072 in HgA
New & Clean Performance Summary			
Number of CTG Operating		2	2
Gross CTG Output (each)		2,642 kW	2,642 kW
Gross CTG Output (total)		5,284 kW	5,284 kW
Gross CTG Heat Rate (LHV)		13,426 kJ/kWh	12,727 Btu/kWh
Total CTG Heat Input (LHV)		70.9 GJ/h	67.2 MBtu/h
Heat Input - Pyrolyzer Indirect Heating Gas Exhaust		2.6 GJ/h	2.5 MBtu/h
Gross STG Output		2,566 kW	2,566 kW
STG Output, percent of CTG Output		48.6%	48.6%
Gross Plant Output		7,850 kW	7,850 kW
Gross Cycle Heat Rate (LHV)		9,037 kJ/kWh	8,567 Btu/kWh
Auxiliary Power/Losses		195 kW	195 kW
Auxiliary Power/Losses, percent of gross		2.48%	2.48%
Combined Cycle Plant Heat Input, MBtu/h (LHV)		73.6 GJ/h	69.8 MBtu/h
Net Combined Cycle Plant Output		7,655 kW	7,655 kW
Net Combined Cycle Plant Heat Rate (LHV)		9,613 kJ/kWh	9,112 Btu/kWh
Net Combined Cycle Plant Efficiency (LHV)		37.5%	37.5%
Combustion Turbine Generator (per unit)			
Ambient Conditions			
Barometric Pressure		101.4 kPa	14.7 psia
Dry Bulb Temperature		15.0 C	59 F
Relative Humidity		60.00%	60.00%
Wet Bulb Temperature		10.81 C	51.47 F
Humidity Ratio		0.63%	0.63%
Fuel Parameters			
Fuel Flow		2,073 kg/h	4,569 lb/h
Heat Input (LHV)		35.5 GJ/h	33.6 MBtu/h
Exhaust Properties			
Mass Flow rate		53,649. kg/h	118,275. lb/h
Temperature		453 C	848 F
Generator Gross Output, kW (each)		2,642 kW	2,642 kW
Heat Recovery Steam Generator (per unit)			
HP Superheater Gas Side Parameters			
Inlet	Mass Flow rate	107,299. kg/h	236,551. lb/h
	Pressure	101.4 kPa	14.7 psia
	Temperature	453.27 C	847.89 F
	Enthalpy	478.0 kJ/kg	205.5 Btu/lb

Table 3-19. Combined Cycle Heat and Mass Balance.

HP Superheater Steam Side Parameters					
Outlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	343.33	C	649.99	F
	Enthalpy	3,074.4	kJ/kg	1,322.0	Btu/lb
HP Steam Desuperheater Spray					
	Mass Flow rate	0.	kg/h	0.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	145.00	C	293.00	F
% Of HP Superheater #3 Steam Flow		0.00%		0.00%	
HP Evaporator Water Side Parameters					
Inlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	246.81	C	476.25	F
	Enthalpy	1,070.1	kJ/kg	460.2	Btu/lb
Blowdown	Mass Flow rate	0.	kg/h	0.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	246.81	C	476.25	F
	Enthalpy	1,070.1	kJ/kg	460.2	Btu/lb
HP Evaporator Steam Side Parameters					
Outlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	252.36	C	486.25	F
	Enthalpy	2,799.3	kJ/kg	1,203.7	Btu/lb
Heat Duty	Heat Duty	20.034	GJ/h	18.991	MBtu/h
HRSG Water Side Parameters					
Inlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	551.6	kPa	80.0	psia
	Temperature	52.31	C	126.15	F
	Enthalpy	219.3	kJ/kg	94.3	Btu/lb
Steam Turbine Generator (per unit)					
Main Steam					
Inlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	343.33	C	649.99	F
	Enthalpy	3,074	kJ/kg	1,322	Btu/lb
Shell	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	13.8	kPa	2.000	psia
	Pressure	103.429	mmHgA	4.072	in HgA
	Temperature	52.26	C	126.07	F
ELEP	Enthalpy	2,204	kJ/kg	948	Btu/lb
UEEP	Enthalpy	2,277	kJ/kg	979	Btu/lb
	Moisture	13.60%		14%	
LP Section Performance Parameters					
Exhaust Loss	Enthalpy	73.698	kJ/kg	31.690	Btu/lb
Annulus Area		2.56.	m^2	27.60.	ft^2
Annulus Velocity		11.37	m/sec	37.30	ft/sec
Efficiency		82.50%		82.50%	
Condenser/Cooling Tower					
Condenser Steam Side					
Inlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	13.8	kPa	2.0	psia
	Temperature	52.26	C	126.07	F
	Enthalpy	2,277	kJ/kg	979	Btu/lb
Outlet	Mass Flow rate	11,585.	kg/h	25,541.	lb/h

Table 3-19. Combined Cycle Heat and Mass Balance.

	Pressure	13.8	kPa	2.0	psia
	Temperature	52.26	C	126.07	F
	Enthalpy	218.7	kJ/kg	94.0	Btu/lb
Heat Duty	Heat Duty	23.849	GJ/h	22.608	Mbtu/h
Condenser Water Side					
Inlet	Mass Flow rate	512,828	kg/h	1,130,581	lb/h
	Pressure	103.4	kPa	15.0	psia
	Temperature	15.56	C	60.00	F
	Enthalpy	65.4	kJ/kg	28.1	Btu/lb
Outlet	Mass Flow rate	512,828	kg/h	1,130,581	lb/h
	Pressure	103.4	kPa	15.0	psia
	Temperature	26.67	C	80.00	F
Heat Duty	Heat Duty	23.853	GJ/h	22.612	Mbtu/h
Miscellaneous					
Boiler Feed Pump					
Suction	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	551.6	kPa	80.0	psia
	Temperature	144.44	C	292.00	F
	Enthalpy	608.3	kJ/kg	261.6	Btu/lb
Discharge	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	4,138.8	kPa	600.3	psia
	Temperature	145.00	C	293.00	F
	Enthalpy	612.9	kJ/kg	263.5	Btu/lb
HP BFP Performance Parameters					
Isentropic Efficiency		85.00%		85.00%	
Condensate Pump					
Suction	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	13.8	kPa	2.0	psia
	Temperature	52.26	C	126.07	F
	Enthalpy	218.7	kJ/kg	94.0	Btu/lb
Discharge	Mass Flow rate	11,585.	kg/h	25,541.	lb/h
	Pressure	551.6	kPa	80.0	psia
	Temperature	52.31	C	126.15	F
	Enthalpy	219.3	kJ/kg	94.3	Btu/lb
Condensate Pump Performance Parameters					
Isentropic Efficiency		85.00%		85.00%	

3.6.4 Overall Process Efficiency Calculations

Measurement of the overall process efficiency (for example, solid biomass to electric power) is not straightforward in this application for several reasons:

- The pyrolysis process and combined cycle operate at different capacity factors. The pyrolysis process is expected to operate slightly more often than the combined cycle (90.4 versus 80 percent capacity factor). For this reason, the hourly throughputs of the two plants are slightly different. A bio-oil storage tank allows for temporary storage. The storage component makes calculation of a "steady-state" efficiency difficult.

- Approximately twenty percent of the incoming biomass is converted to char. The char, which has value as a fuel or as a possible feedstock for activated carbon production, is not consumed in the pyrolysis process.

- Approximately twenty percent of the pyrolysis plant bio-oil production could be used to produce anhydrosugars. Anhydrosugars are a value-added product that is exported off-site. Anhydrosugar production will therefore significantly limit bio-oil fuel for the gas turbine and lower power output.

- A possible permutation of the process may involve several remote, distributed pyrolysis plants and power plants. These plants will be operated on different schedules to meet the needs of the different applications.

None of these represents limitations of the IPCC concept; rather, they all convey certain advantages. For this reason, the project team proposes three different measures of efficiency, as described below.

Net combined cycle efficiency, which measures the efficiency of the combined cycle in converting bio-oil (and a small amount of waste heat from the pyrolysis plant) to electrical power. The net combined cycle efficiency is calculated as:

$$\frac{\text{net combined cycle (CC) power output, kW}}{(\text{bio - oil flow, kg/hr})\,(\text{bio - oil HV, kJ/kg}) + (\text{waste heat input from pyrolyzer to CC, kj/hr})}$$

Table 3-20 shows the calculated net combined cycle efficiency to be 37.5 percent (lower heating value basis).

Table 3-20. Net Combined Cycle Efficiency Calculations (LHV basis).				
Heat Input	**Metric Units**		**English Units**	
Bio-oil mass flow	4,145	kg/hr	9,139	lb/hr
Bio-oil heating value	17,114	kJ/kg	7,359	Btu/lb
Bio-oil heat input	70.9	GJ/hr	67.2	MBtu/hr
Pyrolyzer waste heat input	2.6	GJ/hr	2.5	MBtu/hr
Total combined cycle heat input	73.6	GJ/hr	69.8	MBtu/hr
Power				
Gross combined cycle power output	7,850	kW	7,850	kW
Net combined cycle power output	7,655	kW	7,655	kW
Efficiency				
Net combined cycle heat rate	9,613	GJ/kWh	9,112	Btu/kWh
Net combined cycle efficiency	37.5%	percent	37.5%	percent

Annual average total fuel cycle efficiency, which measures the overall efficiency of the IPCC plant in converting solid biomass (and a small amount of supplemental natural gas) to electrical power. This ratio is calculated on an annual basis to compensate for the different capacity factors of the pyrolysis and combined cycle plants. It assumes that on an annual basis, all of the bio-oil produced by the pyrolysis plant is consumed by the combined cycle. It does not account for the energy in the char or anhydrosugar. The annual average total fuel cycle efficiency is calculated as:

$$\frac{(\text{annual net CC power production, kWh/yr}) - (\text{annual pyrolysis plant power consumption, kWh/yr})}{(\text{biomass consumption, kg/yr}) (\text{biomass HV, kJ/kg}) + (\text{natural gas heat input, kJ/yr})}$$

Table 3-21 shows the annual average total fuel cycle efficiency to be 22.2 percent (lower heating value basis).

Table 3-21. Annual Average Total Fuel Cycle Efficiency Calculations (LHV).

Heat Input	Metric Units		English Units	
Pyrolysis plant capacity factor	90.4	percent	90.4	percent
Biomass consumption (wet), design hourly	5.39	tonne/hr	5.94	ton/hr
Biomass consumption (wet), annual	42,705	tonne/yr	47,074	ton/yr
Biomass heating value	15,665	kJ/kg	6,732	Btu/lb
Biomass heat input, design hourly	84.5	GJ/hr	80	MBtu/hr
Natural gas heat input, design hourly	5.8	GJ/hr	5.5	MBtu/hr
Total IPCC heat input, design hourly	90.3	GJ/hr	85.5	MBtu/hr
Biomass heat input, annual	668,989	GJ/yr	633,801	MBtu/yr
Natural gas heat input, annual	46,150	GJ/yr	43,748	MBtu/yr
Total IPCC heat input, annual	715,139	GJ/yr	677,549	MBtu/yr
Power				
Combined cycle capacity factor	80	percent	80	percent
Net combined cycle power output	7,655	kW	7,655	kW
Annual combined cycle power production	53.6	GWh/yr	53.6	GWh/yr
Pyrolysis plant power consumption	1,197	kW	1,197	kW
Pyrolysis plant capacity factor	90.4	percent	90.4	percent
Annual pyrolysis plant power consumption	9.5	GWh/yr	9.5	GWh/yr
Total annual IPCC net power production	44.1	GWh/yr	44.1	GWh/yr
Efficiency				
Annual average total fuel cycle heat rate	16,198	GJ/kWh	15,347	Btu/kWh
Annual average total fuel cycle efficiency	22.2%	percent	22.2%	percent

Corrected annual average total fuel cycle efficiency, which measures the efficiency of the IPCC plant in converting a portion of the overall solid biomass consumption (and a small amount of supplemental natural gas fuel) to net electrical power. The value used for the biomass consumption excludes the biomass used to produce value-added byproducts (that is, char and anhydrosugar). Further, the electrical and natural gas consumption of the pyrolysis plant are pro-rated for their applicability to bio-oil production. An equation for the corrected annual average total fuel cycle efficiency has not been developed at this time.

4.0 Fuel Supply Availability and Cost

The state of Iowa was selected for establishing fuel supply availability and cost because of Alliant Energy's large service territory within the state and Iowa's desire to produce renewable energy from abundant indigenous fuel resources. Iowa is a state rich in agricultural resources that provide potential sources for biomass fuel. Biomass sources available in Iowa include herbaceous energy crops (HEC) like switchgrass, short-rotation woody crops (SRWC) such as poplar trees, forestry residues, wood by-products, animal manure, crop residues, municipal solid waste, and sewage sludge.

Iowa could become a world leader in the development of biomass fuels as a renewable energy resource because of its large quantity of agricultural products and a proficient agricultural knowledge base. The state has an estimated total biomass production of 4.3×10^5 TJ (4.06×10^{14} Btu) with corn crop residue making up the largest portion of about 42 percent while herbaceous and short-rotation energy crops contribute about 22 percent each. The remainder is made up of animal and other wastes. In addition, Iowa has about 2.66×10^6 acres of marginal land (Land Capability Classification or LCC IV to VII) that is currently used for crop production or considered to be non-Conservation Reserve Program (CRP)/Wetlands Reserve Program (WRP) pasturelands. If this marginal land were committed to dedicated HEC or woody biomass crops production, an additional 2.2×10^5 TJ/yr (2.1×10^{14} Btu/yr) of biomass energy could be obtained from this land (Brown, Hallam, et al., 2001). Based on the compatibility and availability, both corn stover and oat hulls were selected as the choice fuels for this study. This section explores the properties, availability, and cost of using these resources.

4.1 Corn Stover

Corn stover represents Iowa's largest biomass resource. On an annual basis, about 11.5 million tons of corn stover is theoretically available for energy production. Based on a heating value of about 17.4 MJ/kg (7,500 Btu/lb), this translates into approximately 2×10^5 TJ/yr (1.7×10^{14} Btu/yr), which is sufficient to provide heating and electricity requirements to slightly over 1 million homes annually (Iowa DNR, 2002). However, due to the current inefficient collection methods, the price of corn stover may be so high as to make it an infeasible energy source.

Recently, Iron Horse Custom Farms (IHCF) in Harlan, IA implemented the first large-scale corn stover collection operations, approaching 100,000 acres. The study team reviewed the IHCF 1999 Corn Stover Harvest Report, and used its findings as a basis to determine the feasibility of corn stover as potential biomass fuel in Iowa.

4.1.1 Corn Stover Fuel Properties

Fuel properties have significant impacts on the technical and financial viability of biomass fueled power plants. Discussions of corn stover properties are outlined in the following subsections.

4.1.1.1 Physical Properties

Corn stover is essentially the remains of the corn harvest, which is made up of the leaves, husks, stalks, and corncobs. Figure 4-1 illustrates the remnants of corn stover in a field after the corn has been harvested. Corn stover is typically baled in round or square bales; Figure 4-2 illustrates the square bale method.

Figure 4-1. Corn Stover Remnants.   **Figure 4-2. Square Baled Corn Stover.**
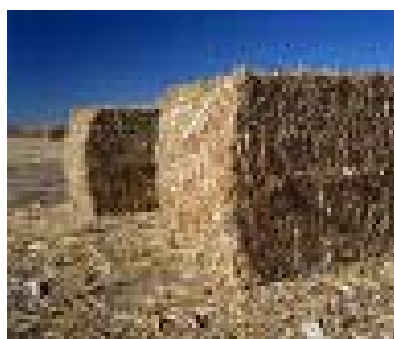
4.1.1.2 Moisture Content

One major concern with corn stover is its moisture content. With increased moisture content, deterioration of the stover becomes a major concern. In addition, moisture in the feedstock for the pyrolysis process is undesirable as it lowers the quality of the bio-oil. The moisture content of corn stover at harvest is generally at about 35 percent. After three days of dry weather the moisture content usually drops below 20 percent (Schechinger et al., 09/30/1999). However, desired moisture content upon delivery to the plant is not always guaranteed.

The main reason farmers cannot guarantee moisture content is due to the variability in weather conditions. In addition, farmers generally do not have the necessary moisture testing equipment to measure the moisture content of corn stover. As a general rule, the lower the moisture content required the higher the cost of the corn stover. By lowering the moisture content requirement of the corn stover there is a much smaller window during which the corn stover can be harvested thereby raising the cost of the fuel. Therefore, a moisture criterion must be set that balances the fuel costs with the desired output of the plant.

Another consideration to moisture content is the collection process. There are two common methods of collecting the corn stover, either by a windrow left by the combine or by raking the corn stover. Raking methods allow for better moisture control, given that evenly distributed corn stover will dry faster than the alternative windrow method should it rain.

4.1.1.3 Contaminants

The presence of contaminants in corn stover is also a major concern. These contaminants, which include dirt and rocks, make up about 2 to 5 percent of the total weight of corn stover. They can be very destructive to bailing equipment and power plant system components, and can increase operation and maintenance (O&M) costs substantially. Rocks and gravel can be avoided to a degree during collection by flailing the corn stover high and keeping the rake off of the surface of the soil. Nevertheless, current corn stover collection methods do not eliminate these contaminants. During the bailing process or when mobile collection equipment travels over corn stover, dirt and soil clods can easily be mixed with corn stover.

4.1.1.4 Chemical Properties

Table 4-1 provides the chemical properties of the different components of corn stover. For this report, the as-fired (6.1 percent moisture) higher heating value of corn stover is assumed to be 16,754 kJ/kg (7,200 Btu/lb). The lower heating value is 15,665 kJ/kg (6,730 Btu/lb).

Table 4-1. Corn Stover Analysis.						
	Seed Corn		**Corn Cobs**		**Corn Husks**	
	As Rec.	**Dry**	**As Rec.**	**Dry**	**As Rec.**	**Dry**
Proximate Analysis, percent:						
Moisture	9.0	0.0	4.2	0.0	6.5	0.0
Volatile Matter	77.9	85.6	77.2	80.6	75.1	80.3
Fixed Carbon	11.7	12.9	12.9	17.4	16.0	17.1
Ash	1.4	1.5	1.2	1.3	2.4	2.6
Ultimate Analysis, percent:						
C	41.7	45.82	44.92	46.89	42.46	45.41
N	6.43	5.96	5.66	5.42	5.89	5.52
H	1.10	1.21	0.95	0.99	0.61	0.65
S	0.13	0.14	0.07	0.07	0.06	0.06
O[*]	49.24	45.37	47.20	45.33	48.58	45.76
Trace Elements						
Cl (ppm)[**]	498		1640		1648	
K (percent)[**]	0.37		1.1		1.0	
Na (ppm)[**]	22		4		74	
Ca (percent)	0		0		0.14	
Notes: [*] Oxygen determined by difference [**] Determined by Instrumental Neutron Activation Analysis						

4.1.2 Corn Stover Fuel Availability and Requirements

Topics on corn stover availability and requirements for the IPCC facility are explored in the following subsections.

4.1.2.1 Harvest Potential

Based on 2001 Department of Agriculture Data, Iowa produced over 1.6 billion bushels of corn with an average yield of 361 bushels/ha (146 bushels/acre). It is estimated that the weight of corn stover produced during harvest is essentially equal to the weight of corn harvested (Schechinger et al., 1999). The weight of corn harvested also varies with its moisture content. The equivalent weight/bushel of grain at various moisture contents can be estimated using the following formula (University of Wisconsin, 2002):

$$EW_{Corn} = DW_{Corn} + [(M)(DW_{Corn})/(100 - M)]$$

Where:

EW_{Corn} = Estimated Weight of Corn per Bushel at Moisture Content M

DW_{Corn} = Dry Weight of Corn per Bushel= 21.5 kg/bushel (47.3 pounds/bushel)

M = Moisture Content of the Corn (percent)

Using the formula above and on the basis of 15 percent corn moisture content and 1.6 billion bushels annual production, the equivalent weight of corn available for harvest is about 40 million tonne/yr (44 million ton/yr). With 100 percent collection efficiency, this translates into about 40 million tonnes/yr of wet corn stover. By taking into account the inefficiencies in the collection methods and the restriction imposed by Soil Conservation Service on harvesting stover from farms with slopes that are 6 percent more, the corn stover harvest yield is approximately 3.36 dry tonnes/ha (1.5 dry tons/acre).

Figure 4-3 and Figure 4-4 illustrate the amount of corn stover available in Iowa on a tonne basis and on a tonne/km^2 basis, respectively, assuming soil conservation practices and a harvest yield of 3.36 dry tonnes/ha (1.5 tons/acre). In addition, Figure 4-3 and Figure 4-4 also show the electric transmission lines and locations of Alliant Energy generation facilities.

4.1.2.2 Stover Requirements

For this study it was assumed that the combined cycle biomass plant would consume biomass fuel at 5.39 tonnes/hr (5.94 tons/hr), with the pyrolysis portion of the plant operating 330 days per year on average. Table 4-2 summarizes the corn stover requirements for the plant operations.

Depending on the hours of operation, the power plant will need at peak consumption 129 tonnes/day (143 tons/day), which equates to approximately 220 round bails of stover/day. Industry contacts indicated that 220 bails/day was attainable. Since stover has such a small harvest window on-site storage or field storage will need to be considered.
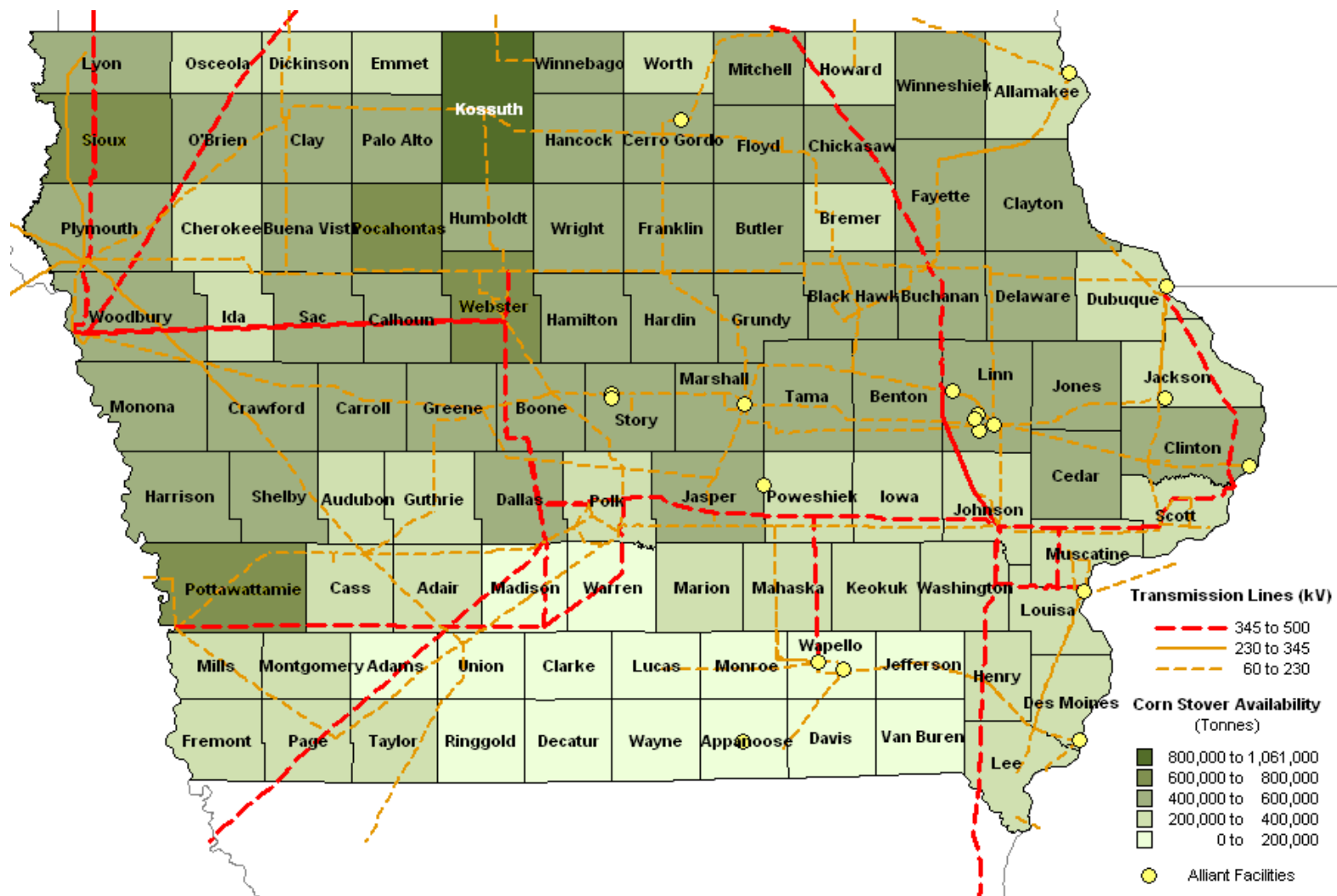
Figure 4-3. Annual Corn Stover Potential (tonnes).
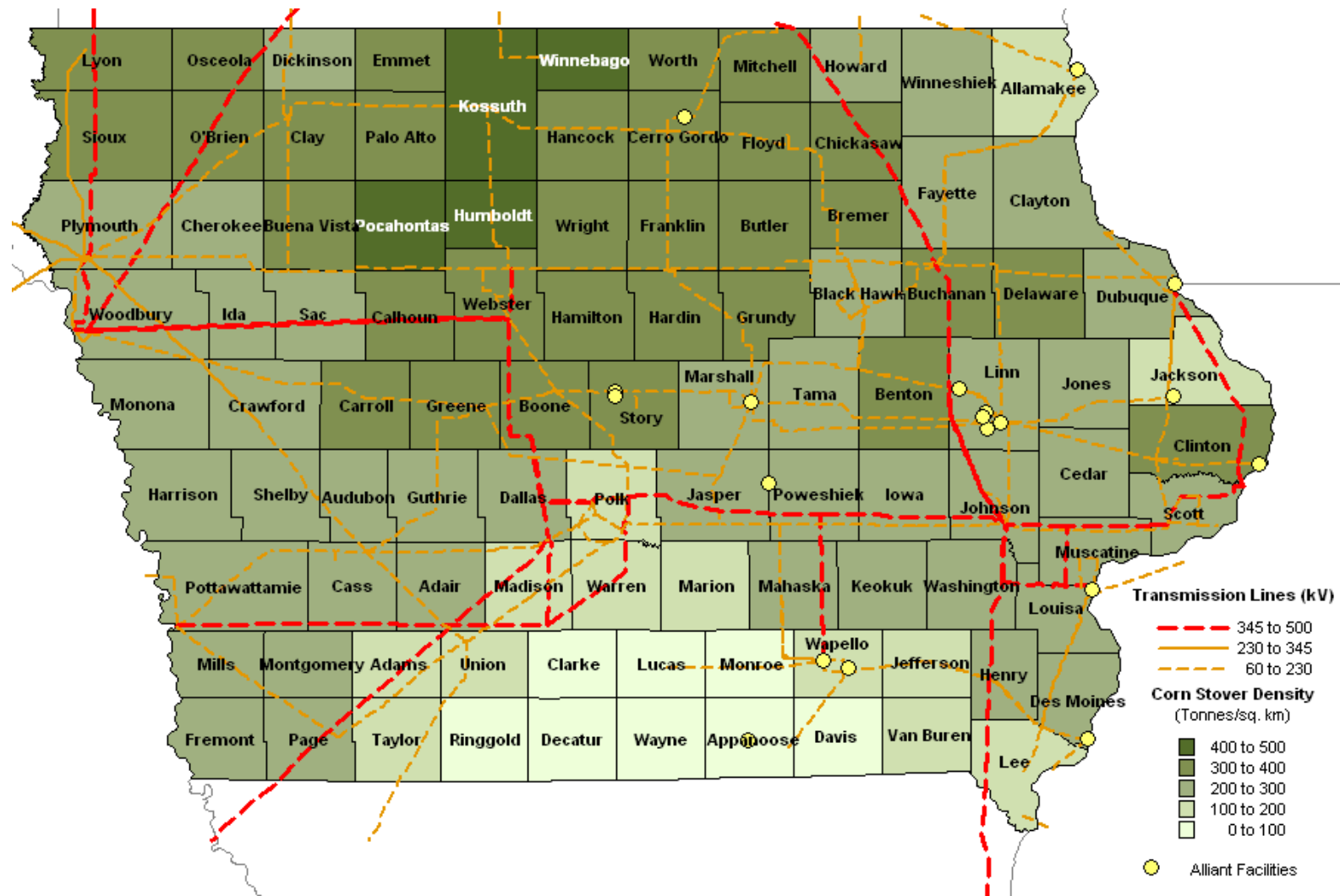
Figure 4-4. Annual Corn Stover Potential (tonnes/km^2).

Table 4-2. Corn Stover[a] Requirements for Pyrolysis Plant Operations.		
Item	**Metric Units**	**English Units**
Design Capacity		
Design Hourly Consumption	5.39 tonne/hr	5.94 ton/hr
Design Daily Consumption	129 tonne/day	143 ton/day
Projected Average Consumption		
Projected Pyrolysis Plant Capacity Factor[b]	90.4 percent	90.4 percent
Projected Hourly Consumption	4.88 tonne/hr	5.37 ton/hr
Projected Daily Consumption	117 tonne/day	129 ton/day
Projected Annual Consumption	42,700 tonne/yr	47,100 ton/yr
Projected Annual Consumption (bone dry)	40,100 tonne/yr	44,200 ton/yr
Farming Requirements		
Stover Harvest Density[c] (bone dry)	0.551 tonne/Ha	1.5 ton/acre
Required Harvest Acres (Annually)	11,931 Ha	29,500 acre
Average Iowa Farm Size[d]	139.6 Ha	345 acre
Average Number of Farms Required	85	85

Notes:
[a] Except as noted, refers to wet corn stover "as fired" at 6 percent moisture.
[b] Refers to pyrolysis portion of the cycle only. Combined cycle power generation may operate on different schedule.
[c] Conservative estimate of harvested stover in the state of Iowa.
[d] Source: National Agriculture Statistics Service for US Agriculture.

4.1.3 Storage

Storage is another key aspect to the feasibility of using corn stover as a potential fuel. Assuming that the plant operates with a fuel rate of 5.39 tonne/hr (5.94 ton/hr), the plant will require about 220 round bails per day. For this study it was assumed that the plant would implement just-in-time stover delivery with 2 days of on-site storage. Recommended storage for round bales is that the bales are stacked 5 high in a pyramid formation as illustrated in Figure 4-5 (Schechinger et al., 1999). Therefore, 2 days of storage (440 bales) would require a minimum of 255 m^2 (2,750 ft^2) of storage area stacked approximately 7 m (23 ft) high.

Moisture control and ventilation must also be considered to reduce the risk of loss due to deterioration and fire. Should corn stover be selected as a fuel, additional storage considerations must be explored such as covered or indoor storage, drainage, rodent

control, and moisture control. Alternative stover storage methods, such as ensiling in large concrete bunkers, are under development and may be considered in the future.
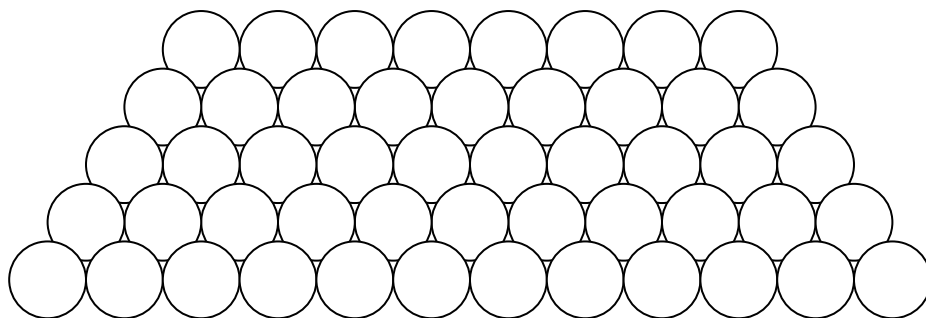
Figure 4-5. Recommended Round Bale Storage (Side View).

4.1.4 Harvesting Equipment

The majority of farmers that harvest corn do not have the necessary equipment for corn stover harvesting. Therefore, the power plant must either consider contracting with balers that have the necessary equipment or purchasing baling equipment and hiring operators to harvest the stover. Due to high equipment costs, operator training, and maintenance of baling equipment, it seems that contracting with balers may be the most economic alternative.

Presently, conventional harvesting systems do not treat the stover as a valuable commodity. It is left strewn on the ground and driven over by other equipment involved in the grain harvest. A whole stalk harvest system could improve the stover quality, reduce dirt contamination and lower the stover cost by collecting the stover in a one-step process and eliminating the need for wrapping or twining. New harvest systems are under development that should allow a substantial reduction in harvest cost of at least 50 percent per acre.

4.1.5 Corn Stover Delivered Cost

The study team conducted a market survey to determine the delivered cost of corn stover. Based on the literature (Schechinger et al., 1999) and information from current market sources, the study team estimates that facility might be expected to pay about $35/wet tonne ($31.8/ton) for delivered corn stover. This pricing is somewhat aggressive and will require diligence on the part of the project development team to ensure that it can be obtained. The economic model includes a high case sensitivity analysis at $50/wet tonne ($45.4/ton). To simulate the effect of increased harvesting efficiency, a case was also examined at $20/wet tonne ($18.1/ton). Note that the delivered price includes a transportation component in the range of $5 to $15/tonne.

4.2 Oat Hulls

Oat harvesting and processing in the Midwestern region make oat hulls an attractive alternative to corn stover as a candidate biomass fuel. Based on 2001 USDA agricultural data, Iowa produced about 9 million bushels of oats which is about 6 percent of US total. Not only is there an abundance of oats in the Midwestern region, but there are also numerous oat processing plants as well. These plants produce large quantities of oat hulls as a waste by-product. Unlike corn stover which is dispersed in fields and requires special harvesting, oat hulls are concentrated at oat processing plants. The study team contacted several Midwestern entities in the oat processing business to determine oat hull availability and market value.

4.2.1 Oat Hull Fuel Properties

Fuel properties have significant impacts on the technical and financial viability of biomass fueled power plants. Discussions of oat hull properties are outlined in the following subsections.

4.2.1.1 Physical Properties

When mature, the oat plant is about thirty six inches tall and has several stalks per plant as illustrated in Figure 4-6. There are normally up to four seeds per sheath, and the seeds hang on very fine, wiry stems. There are many seed sheaths per head, and the entire grain head can be up to eight or ten inches long. The oat seeds are covered with a hull, which is considered to be the oat hull as illustrated in Figure 4-7. Once the oats have been through the milling process the oat hull is removed as illustrated in Figure 4-8.

4.2.1.2 Moisture Content

Oat hulls have an obvious advantage over corn stover in that moisture content on an as received basis is approximately 9 percent, whereas corn stover moisture content ranges from 20 percent to 35 percent depending on stover harvesting conditions. To obtain lower moisture content for corn stover extended drying periods or thermal drying may be required. Since oat hulls are processed in a controlled environment, sheltered from rain and other environmental effects, moisture content of oat hulls is relatively simple to manage. The distribution of moisture in oat hulls is also relatively uniform, allowing the moisture content to be measured efficiently and accurately. Low moisture content in oat hulls ensures that there is lower tendency toward deterioration and better quality control.

Figure 4-6.	**Figure 4-7.**	**Figure 4-8.**
Oat Plant.	**Oat Seeds with Hull.**	**Oat Seeds with Hull**
		Removed.

4.2.1.3 Contaminants

Contaminants are of much less concern in oat hulls than they are in corn stover, since they have had the majority of contaminants removed at the oat processing plant. Therefore, additional processing and cleaning equipment may not be needed for an oat hull fuel source.

4.2.1.4 Chemical Properties

Table 4-3 provides the chemical properties of oat hulls. For this report, the as-fired (6.1 percent moisture) higher heating value of oat hulls is assumed to be 16,754 kJ/kg (7,200 Btu/lb). The lower heating value is 15,665 kJ/kg (6,730 Btu/lb). This is the same as the corn stover heating value.

Table 4-3 Ground Oat Hull Ultimate Analysis		
	As Received	**Dry**
Moisture, percent	9.56	0.00
C, percent	41.23	45.59
H, percent	5.35	5.914
N, percent	0.57	0.63
S, percent	0.05	0.05
O, percent	37.61	41.59
Ash, percent	5.63	6.23
Heating Value, Btu/lb	6,990	7,729
Source: Commercial Testing and Engineering CO.		

4.2.2 Oat Hulls Fuel Availability

Topics on oat hulls availability and requirements for the IPCC facility are explored in the following subsections.

4.2.2.1 Oat Hull Availability and Market Value

In order to determine the quantity of oat hulls available, several Midwestern oat processing plants and grain trading companies were contacted. The grain trading companies indicated that they could provide the facility with approximately 108 tonnes/week (15 tonnes/day) at an average of $50/tonne delivered. The mills produced between 12 and 136 tonnes/day, and most would not disclose a price at this time. However, one of the mills indicated that it would be interested in meeting the biomass facility requirements by signing a long-term contract at an undelivered price of $31/tonne. It may be possible to negotiate a lower rate.

Larger oat processing plants face greater disposal difficulties and were more likely to quote prices in the $25 - 35/tonne range (not delivered). Smaller plants generally had short term contracts with local businesses and disposal is not a significant issue; they quoted between $40 - 45/tonne. The study team also contacted LaBudde Group, a company out of Wisconsin that specializes in grain analysis and market pricing for grains and by-products. They suggested that the market price for oat hulls varied between $16 - $28/tonne. However, with a long term contract they believed that $22/tonne was reasonable. Based on the preceding and the assumption that the pyrolysis plant will be near if not adjacent to a large oat hull processing facility, $25/tonne was assumed for the base case economic model.

4.2.2.2 Oat Hull Requirements

At peak capacity the facility would need 129 tonnes (143 tons) of oat hulls per day. There are enough oat hulls available statewide to meet the facility's requirements; however, there are only two milling companies in the Midwestern region that can meet that demand entirely: Quaker Oats in Cedar Rapids and Con Agra in Sioux City. This demonstrates that the potential for oat hulls as a biomass resource is much smaller than corn stover. It may be advisable to locate the new power plant at one of these two facilities. Otherwise, a regional plant may need to be developed to collect the oat hulls from several sources.

4.2.2.3 Storage

Storage of oat hulls is more straightforward than corn stover. Oat hulls are typically stored in grain silos so they do not encounter rain and other environmental effects that may result in degradation.

Two common types of grain silos are concrete and steel grain silos. Bridging is common with the storage of oat hulls. This is essentially the congregation of oat hulls that results in a bridge preventing the further filling or emptying of the silo. There is less bridging with a concrete structure since the concrete structure minimizes condensation build up. Storing oat hulls may require either a grain bin vibration system to keep the oat hulls from adhering to the walls of a steel silo or an auger system to prevent bridging.

Locating the new power plant at an oat mill may substantially reduce or eliminate the need for additional storage.

4.2.2.4 Transportation

Transportation costs are not an issue if the plant is located at an existing mill. However, if this is not possible, local contractors may be retained to transport the hulls. A survey was conducted to acquire information on the capacity of the local agricultural trucks, number of daily truckloads required to meet the facility's requirements, and the price per truckload. The capacity of the trucks varied with the type of truck. Hopper bottom trucks hold between 20 to 23 tonnes of ground oat hulls per load while the vans hold approximately 15 to 18 tons/load. To maintain operations, 129 tonnes/day (143 tons/day) would have to be shipped to the site. This corresponds to approximately 6 hopper truck loads/day. Trucking companies typically charge between $350 to $500/day for each of their trucks.

4.3 Conclusions

Both corn stover and oat hulls are suitable biomass fuels that are widely available in Iowa to support the IPCC power plant. However, our analysis indicates that oat hulls are more economical and technically feasible biomass fuel for the IPCC power plant for the following reasons:

- Oat hulls can be collected at a single source and involve only one contract. While corn stover would likely require the facility to contract with at least 40 different farmers.
- The moisture level in oat hulls (9 percent) is much less than corn stover (18 – 30 percent) on an as received basis.
- Storage of oat hulls is much simpler utilizing grain silos, versus corn stover bales which require a much larger footprint and have a lower storage density.

- The handling of oat hulls is also much simpler than corn stover which involves awkward bale stacking procedures, difficult moisture control, fire risk, the disposal of baling twine or plastic, etc.

- Oat hulls are a significantly cleaner fuel than corn stover requiring virtually no processing before pyrolysis.

- The delivered cost of oat hulls ($25/tonne) is lower than the cost of corn stover ($35/tonne).

Nevertheless, the quantity of corn stover far exceeds oat hulls and stover collection and handling methods are being constantly improved. Though corn stover may not be the ideal source for biomass fuel at this time, it may be in the near future.

5.0 Economic Analysis

An economic analysis was performed using the Black & Veatch pro forma model to evaluate the economic viability of the proposed IPCC plant. The economic analysis considered two options for the IPCC project:

- **Maximum Power Production Option** – In this option the entire bio-oil output of the pyrolysis plant is used to fuel the combined power plant.
- **Value-Added Chemicals Option** – In this option, value-added chemicals are extracted from the bio-oil resulting in lower flow of bio-oil to the power cycle.

For each option, the pro forma evaluated a base case (most likely) scenario and a number of sensitivity scenarios based on the two types of biomass fuels; corn stover and oat hulls.

5.1 Pro Forma Overview

The pro forma determines the project return on investment, as measured by the project net present value (NPV) and the internal rate of return (IRR).

The NPV determines whether the financial returns will be high enough to justify the project. In the pro forma analysis, NPV is calculated by discounting the after-tax cash inflows to the beginning of the operating period and then subtracting the corresponding future value of the equity investment. If the difference is greater than zero, the project is acceptable, whereas a negative NPV would mean that the project is not acceptable at the chosen discount rate.

The IRR is the discount rate at which NPV is zero. An IRR higher than the chosen discount rate (or minimum equity return) indicates the project is acceptable.

5.2 Base Case Assumptions

The economic analysis for the base case maximum power production scenario is based on the study assumptions listed in Table 5-1.

Table 5-1. Economic Analysis Assumptions (2003 $US)	
Maximum Power Production Option	
Financial assumptions	
Construction period, months	24
Start date of commercial operation	January 2005
Discount rate (minimum equity return), percent	12
Inflation rate, percent	2
Debt to equity ratio	50 / 50
Project term, years	28
Interest rate on long term debt, percent	8
Effective tax rate, percent	30
Depreciation method	7-year, double declining balance
Project revenues	
Average electric sales rate, $/MWh	30
Capacity credit, $/kW-yr	24
Char sales rate, $/GJ	1.00
Renewable energy credit, $/MWh	0
Anhydrosugar value, $/kg	1.00
Power production and consumption	
Net combined cycle power block output, kW	7,655
Annual combined cycle power production, GWh/yr	53.6
Pyrolysis plant power consumption, kW	1,197
Annual pyrolysis plant power consumption, GWh/yr	9.5
Fuel information	
Biomass fuel feed rate, tonne/hr	5.39
Oat hulls delivered cost, $/wet ton	25
Corn stover delivered cost, $/wet ton	35
Ash disposal cost, $/ton	0 (Land Applied)
Supplemental natural gas consumption, MBtu/yr	43,748
Supplemental natural gas price, $/MBtu	3.50
Operating profile	
Operating mode	Base Load
Pyrolysis plant capacity factor, percent	90.4
Combined cycle capacity factor, percent	80.0
Operating and maintenance cost	
Oat hulls O&M cost, $/yr	1,518,778
Corn stover O&M cost, $/yr	1,678,778
Total project cost, $	18,392,770

The following changes are made to the base case assumptions for the value-added chemicals option:

- Recover 20% of the bio-oil yield as anhydrosugar (5.8 million kg/yr)
- Add anhydrosugar sales revenue of $1/kg
- Reduce capacity factor of the combined cycle plant by 20% (from 80% to 64%)
- Raise the power sales price to $35/MWh (from $30/MWh) due to the few operating hours
- Increase capital and operating costs (see the following sections)

5.2.1 Capital Costs

A total project capital cost estimate was prepared based on inputs from the various team members, vendor quotes, and historical construction industry data. This estimate is indicative in nature (+/- 30 percent) and will vary based upon the final system design. The cost estimate includes direct and indirect costs for design, equipment procurement and construction. Table 5-2 shows the estimate summary. It was determined that at this level of accuracy, the cost estimate for the oat hull and corn stover plants could be assumed to be the same.

<table>
<tr><th colspan="2">Table 5-2. Capital Cost Estimate (2003 US$).</th></tr>
<tr><td>Maximum Power Production Option</td><td></td></tr>
<tr><td>Turnkey Pyrolysis Plant</td><td>4,631,000</td></tr>
<tr><td>Combined Cycle Plant</td><td></td></tr>
<tr><td>Combustion Turbine Generators</td><td>3,682,917</td></tr>
<tr><td>Steam Turbine Generator</td><td>850,000</td></tr>
<tr><td>Heat Recovery Steam Generator</td><td>520,000</td></tr>
<tr><td>Combined Cycle Balance of Plant</td><td>1,515,875</td></tr>
<tr><td>Combined Cycle Installation</td><td><u>2,627,517</u></td></tr>
<tr><td>Subtotal Combined Cycle</td><td>9,196,308</td></tr>
<tr><td>IPCC Balance of Plant Equipment and Construction</td><td>1,500,000</td></tr>
<tr><td>Indirect Expenses</td><td><u>3,065,462</u></td></tr>
<tr><td>Total Maximum Power Production Project Cost</td><td>18,392,770</td></tr>
<tr><td>Incremental Equipment for Value-added Chemicals</td><td></td></tr>
<tr><td>Biomass Pretreatment System</td><td>727,250</td></tr>
<tr><td>Fractionating Quencher</td><td><u>545,480</u></td></tr>
<tr><td>Subtotal Incremental Costs</td><td><u>1,272,730</u></td></tr>
<tr><td>Total Value-added Chemical Project Cost</td><td>19,665,500</td></tr>
<tr><td colspan="2">Note: Estimated capital cost is the same for both oat hull and corn stover fuels.</td></tr>
</table>

The proposed IPCC system is estimated to have a total project cost of $18.4 million, which, based on a net combined cycle output of 7,655 kW, is equal to

$2,400/kW. This cost does not include the additional equipment required for production of value-added chemicals. This additional equipment is estimated to cost $1.3 million. The IPCC capital cost compares favorably to conventional biomass power systems in this size range, which cost around $2,000 for "bare-bones" systems to over $3,000/kW for systems designed for higher efficiency and reliability.

Because this is a first-of-a-kind facility, a certain amount of uncertainty in capital costs should be reasonably expected. Based on the indicative accuracy level (+/- 30 percent), the capital cost for the maximum power production option may vary from $12.9 million to $23.9 million. An investigation of the sensitivity of the project to capital cost variation is included later in this section.

5.2.2 Operating and Maintenance Costs

Operating and maintenance (O&M) costs are defined as all expenses associated with the processing of the biomass feedstock and production of the char, bio-oil, and electricity. O&M costs typically include production and maintenance labor, chemical costs, water costs, waste disposal costs, maintenance parts and materials, and various other expenses associated with plant operation and maintenance. Not included in O&M costs are items such as fixed charges on capital investment which consist of return on investment, depreciation, and income tax.

Operation and maintenance costs are typically separated into two components: fixed costs and variable costs. The fixed component of O&M costs consists primarily of the wages and wage-related overheads for the permanent plant staff of operators, maintenance personnel, and supervisory personnel. These costs are classified as fixed because the expense for normal straight-time work is incurred whether or not the plant is operating. Conversely, the variable component of O&M costs includes expenses such as chemicals, electric power, maintenance parts and materials, and consumables, all of which are directly influenced by the amount of plant operation.

Even though the plant will be largely automated, by far the largest O&M cost for the maximum power production option is personnel. For oat hulls, it is estimated that a total of six people operating on four shifts could operate the plant continuously. Corn stover will likely require additional fuel handling personnel, and a staff size of eight is estimated. Including staff costs and other miscellaneous O&M expenses, the total estimated annual O&M cost is about $1.5 million and $1.7 million for oat hulls and corn stover, respectively.   These estimates are equivalent to about $30/MWh, which is comparable to O&M costs for other power plants in this size range. A breakdown of this estimate and estimates for the value-added chemicals option are provided in Table 5-3. The additional O&M costs for production of value-added chemicals are very large, $3.3

million extra per year. However, there is a proportionate increase in projected revenues from anhydrosugar sales.

Table 5-3. Operating and Maintenance Cost Estimate (2003 US$).	Oat Hulls	Corn Stover
Maximum Power Production Option		
Overall Plant Labor	480,000	640,000
Other Overall Fixed O&M Costs (non-labor)	288,000	288,000
Pyrolysis Plant		
Consumables & Services	67,847	67,847
Maintenance	261,406	261,406
Supplemental Natural Gas Consumption	153,118	153,118
Combined Cycle Plant		
Variable O&M Expenses	268,406	268,406
Fixed O&M Expenses	incl. above	incl. above
Total Maximum Power Production O&M Cost	**1,518,778**	**1,678,778**
Incremental O&M Costs for Value-added Chemicals		
Biomass Pretreatment System	2,300,000	2,300,000
Fractionating Quencher	1,000,000	1,000,000
Subtotal Incremental Costs	3,300,000	3,300,000
Total Value-added Chemical O&M Cost	**4,818,778**	**4,978,778**

5.3 Renewable Energy Credits

An additional and important consideration for this project is evaluation of renewable energy credits. The use of biomass for electricity generation may qualify for renewable energy incentives, green power sales, greenhouse gas emission credits, or other economic credits associated with renewable power. These are sometimes referred to as "Green Tags," "Renewable Energy Certificates," or "Tradable Renewable Credits". The emergence of renewable portfolio standards in various states (and perhaps on a national level) will support development of larger markets for trading these credits.

The base case assumptions do not assign any premium economic value for the renewable energy produced by the biomass plant. The green premium represents incremental costs above the indicated power sales prices ($30/MWh). This premium, the value of which is captured in "renewable energy credits," may be marketed to retail customers, wholesale customers, or used to offset purchases of renewable energy that might be made from other power/credit suppliers. The green premium might also reflect the value of tax credits for renewable power generation. Currently there is a $18/MWh federal production tax credit available for wind projects. The tax credit is also available

to biomass power produced from energy crops or poultry litter. There are current proposals in Congress to expand this credit to other biomass fuels, including those evalauted for this project.

As discussed in the next section, because there is no value assigned to the renewable attributes of the biomass power in the base case, most of the base case economic results are negative. The economic model was used to calculate the renewable energy credit value (breakeven value) for which the IRR equals the minimum equity return (the discount rate, 12 percent).

5.4 Base Case Results

Table 5-4 summarizes the results of the four scenarios investigated. These results are discussed further below. Appendix B includes an example pro forma model.

Table 5-4. Summary Economic Analysis Results.		
	Oat hulls	**Corn Stover**
Internal Rate of Return		
Maximum Power Production Option	NA	NA
Value-added Chemicals Option	5.46%	-1.58%
Renewable Energy Premium Required to Meet Minimum Equity Return (12%)		
Maximum Power Production Option	$55.13/MWh	$66.15/MWh
Value-added Chemicals Option	$17.99/MWh	$31.75/MWh

5.4.1 Maximum Power Production Option

For power production only, the base case economic model does not result in a positive internal rate of return (IRR). In other words, the project does not economically appear viable under the base case assumptions. These assumptions are based on dispatching the plant in a large utility system with relatively low power costs ($30/MWh energy, and $24/kW-yr capacity). This scenario demonstrates how the IPCC plant would compete against traditional open-market utility coal, hydro, and gas resources, without considering any of the external benefits of biomass power.

The base case scenario does not include any credits for the renewable energy aspects of the plant. The model was used to calculate the "breakeven" incremental renewable energy credit ($/MWh) for which the IRR equals the discount rate (12 percent). Based on the model results, the breakeven credit value is $55.13/MWh for oat hulls and $66.15/MWh for corn stover. These credits are in addition to the $30/MWh power sales rate; therefore the total breakeven power sales rate would be $85.13/MWh

for oat hulls and $96.15/MWh for corn stover. Power generated by most traditional biomass power plants is $70 to $100/MWh. Thus, although the maximum power option would not likely be competitive with traditional wholesale utility assets, it appears to be competitive with other biomass options. Considering that this study characterized a small first-of-a-kind facility and that future plants will likely have improved economics, additional investigation is recommended.

5.4.2 Value-Added Chemical Option

Value-added chemical production greatly improves the economics of the project. Including a value of $1/kg for anhydrosugar in the base case economic assumptions indicates that the project may be viable in the base case, with a positive IRR of 5.46 percent for oat hulls. To obtain the minimum equity return (12%), a renewable energy credit of $18/MWh (equal to the current production tax credit) would be needed for oat hulls, and $31/MWh would be needed for corn stover. Because of the potential benefit of value-added chemical production, it is recommended that this alternative be investigated further.

5.5 Sensitivity Cases

The economic analysis evaluated a number of sensitivity cases to assess the impacts of key variables on the project economic viability. Key variables that were considered in the sensitivity analysis include:

- Capital cost
- Operating and maintenance cost
- Fuel cost
- Anhydrosugar value

For ease of comparison, the incremental renewable energy credit necessary to obtain minimum equity return is used as the economic metric. The following charts show sample results of the sensitivity investigations.
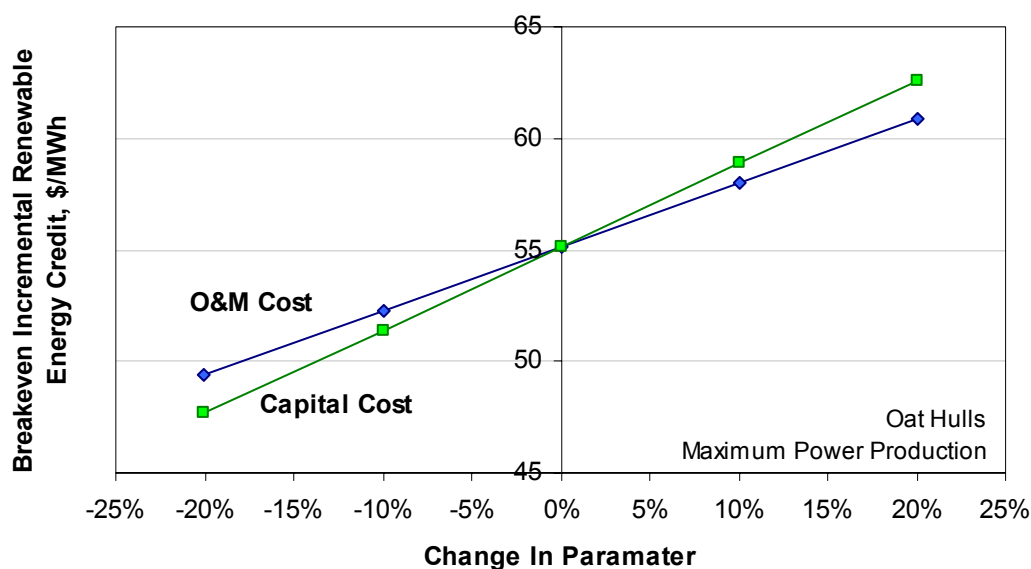
Figure 5-1. O&M and Capital Cost Sensitivity Investigations.
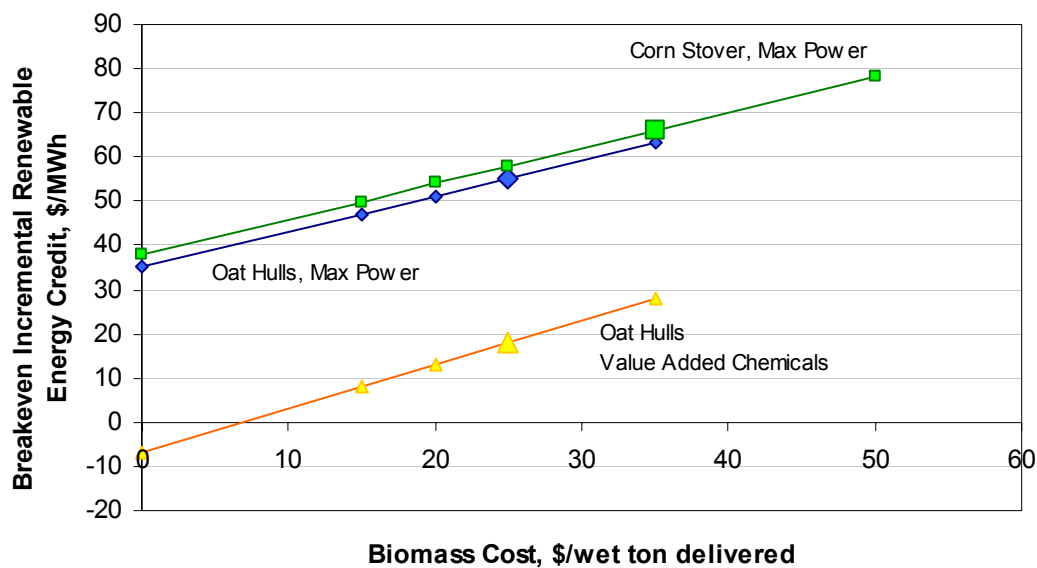
Figure 5-2. Biomass Cost Sensitivity Investigations.
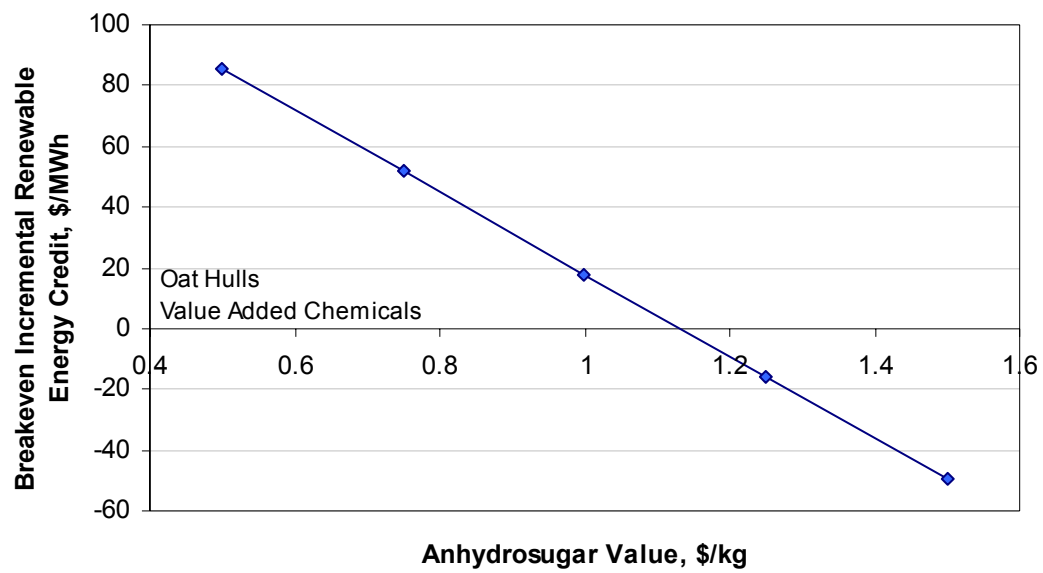
Figure 5-3. Anhydrosugar Sensitivity Investigation.

6.0 Benefits

The proposed project has the ability to address multiple national benefits (power, heat, byproducts; rural development; environmental quality; greenhouse gas emissions; economic benefits; infrastructure and fuel supply benefits and impacts). These are discussed further in this section.

6.1 Power, Heat, and Byproducts

The proposed integrated pyrolysis combined cycle converts 5.39 tonne/hr of biomass into 7.66 MW of electric power at 37.5 percent net combined cycle efficiency. Byproducts include 1.1 tonne/hr of char could be upgraded to activated carbon, and up to 742 kg/hr of highly valuable anhydrosugar levoglucosan.

The major differentiator between the proposed IPCC power system and conventional biomass power systems is efficiency. Efficiency is low for small conventional biomass power systems, typically around 15 percent, and rarely exceeding 20 percent. By comparison, the net combined cycle efficiency for this process is estimated to be 37.5 percent, and the average total fuel cycle efficiency is estimated to be 22.2 percent. The flexibility of not having to co-locate the pyrolysis plant with the power plant also potentially improves project economics. The bio-oil could be generated at numerous distributed pyrolysis plants located near biomass resource concentrations. The oil could then be transferred to a large central biopower facility (perhaps 100 MW or more) and utilized in state-of-the art combined cycle technologies with efficiencies of over 50 percent. The bio-oil would also be a suitable cofiring fuel for existing coal plants, with minimal retrofit costs necessary to cofire the fuel.

6.2 Rural Development Benefits

The proposed power system provides high thermodynamic efficiency at a scale suitable for many small communities (< 10 MWe). The fuel demand (129 tpd) can be provided locally. For example, a plant fueled with corn stover (assuming only one-half of the stover was removed from the field) would require collection of stover from about 12,000 hectares (30,000 acres) to support the annual fuel requirements. In Iowa, about 85 farms would be needed on average to support the project. Anecdotal evidence from a commercial corn stover harvesting and storage operation in western Iowa suggests that net income for farmers doubled on acreage for which they received $16/ton of corn stover collected by a custom harvester (Olson, 2001).

In the United States, most states are net importers of energy. For example, the state of Iowa imports over 90% of its energy, which represents billions of dollars

annually.   Many Midwestern states have sufficient agricultural base to generate a significant amount of their energy requirements without impacting their ability to provide food, feed, and fiber in the current U.S. economy. For example, the state of Iowa alone could provide 4.3 x 10^5 terajoules (4.06 x 10^{14} Btu) of biomass energy without appropriating any cropland presently used for food, feed, or fiber production (Brown, 1994). This is enough resource to fuel over 600 of the biomass power plants proposed here.

6.3 Environmental Quality and Greenhouse Gas Emissions

Environmental benefits are strongly dependent on the kind of fuel and conversion process the proposed power plant replaces. Compared to coal, the most common power generation fuel, the proposed plant would provide excellent environmental benefits. Reductions would be made in various pollutants. Biomass fuels contain little sulfur compared to coal, resulting in decreased production of sulfur dioxide. Toxic metals, such as mercury, cadmium, and lead, will also be reduced because of the substitution of biomass for coal. The proposed power cycle is designed to minimize emissions of particulates, and it will use combustion turbine power generation to substantially reduce nitrogen oxide emissions compared to coal.

Increasing energy efficiency and reducing dependence on fossil fuels with renewable energy are two of the leading options for reducing emissions of carbon dioxide (CO_2), the principal greenhouse gas. Fossil-based power generation accounts for thirty-five percent of the anthropogenic CO_2 emission in the U.S. Biomass power is viewed as a carbon-neutral power generation option.   While carbon dioxide is emitted during biomass combustion, an equal amount of carbon dioxide is absorbed from the atmosphere during the biomass growth phase. Thus, biomass fuels "recycle" atmospheric carbon, minimizing global warming impacts. The overall net effect of adding the IPCC project will be a net reduction of CO_2 emissions.

6.4 Project Sustainability and Opportunities for Replication

Sustainability of the proposed project is strongly dependent on fast changing political and regulatory factors in the United States. Without subsidy, dedicated energy crops, such as switchgrass, cost about $3/MMBtu. In contrast, Powder River Basin coal is delivered to Iowa at less than $1/MMBtu. Concerns about sulfur, mercury, and carbon dioxide emissions from coal-fired plants may eventually lead to regulatory changes that improve the attractiveness of biomass for power generation. Improvements in production and harvesting practices are also expected to improve the economics of this feedstock. However, at present, the sustainability of the proposed biomass power plant depends on

exploitation of niche markets associated with agricultural residues, such as corn stover, and agricultural processing wastes, such as oat hulls. Opportunities for early replication will be dependent on local factors of power demand and waste resource availability.

6.5 Economic Benefits

Properly configured, the proposed IPCC project represents a good opportunity for investors with attractive economic benefits. The proposed IPCC system is estimated to have a total project cost of $18.4 million, which, based on a net combined cycle output of 7,655 kW, is equal to $2,400/kW. For a small amount of additional capital, recovery of valuable chemicals may also be performed. The IPCC capital cost compares favorably to conventional biomass power systems in this size range, which cost around $2,000 for "bare-bones" systems to over $3,000/kW for systems designed for higher efficiency and reliability.

The economic analysis showed that at electricity sales prices of $30/MWh, the maximum power option would not be competitive with traditional utility assets; however, this configuration appears to be competitive with other biomass options. Value-added chemical production greatly improves the economics of the project. To obtain the minimum equity return (12%), a renewable energy credit of only $18/MWh (equal to the current production tax credit) would be needed for oat hulls. Considering that this study characterized a small first-of-a-kind facility and that future plants will likely have even better economics, additional investigation is recommended.

This technology may be an excellent match for on-site heat and power generation at facilities with a ready supply of waste biomass. The electricity produced would directly offset retail rates (as opposed to having to compete against low wholesale power costs). The combination of lower fuel cost, power cost savings, and recovery of process waste heat for steam generation would greatly enhance economics.

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List of Acronyms and Abbreviations

B	bio-oil
BCO	bio-crude oil
C	cellobiosan
CBM	bare module cost
CCF	contingency and fee
CC	combined cycle
CD	total direct cost
CE	engineering expenses
CFIT	total cost of freight, insurance, and taxes
CID	total indirect cost
CL	direct labor cost
CM	materials for installation cost
CO	construction overhead
CP	equipment (f.o.b) cost
CRP	Conservation Reserve Program
CTM	total module cost
ΔH	magnitude of heat transfer operations
DSCR	debt service coverage ratio
DWCorn	dry weight of corn per bushel
EPI	Energy Product of Idaho
EWCorn	estimated weight of corn per bushel at moisture content M
f.o.b	Free on Board
Fe	iron
G	volatiles comprise gas
H	heavy
HEC	herbaceous energy crops
HFCS	high fructose corn syrups
IGCC	integrated gasification combined cycle
IHCF	Iron Horse Custom Farms
IPCC	integrated pyrolysis combined cycle
IRR	internal rate of return
L	light
LCC	Land Capability Classification
LG	levoglucosan
LMF	labor module cost

M	moisture content of the corn
MBGF	moving bed granular filter
MMF	material module factors
NPV	net present value
O	organics
O&M	operation and maintenance
PL	pyrolytic lignin
SRWC	short-rotation woody crops
Std 1	measured intensity standard
V	volatiles
W	water
WE	water extraction
WRP	Wetlands Reserve Program

Appendix A. Fractionating Condenser Cost Estimation Approach

A simplified fractionating column will be used for cost estimation purposes as shown below:
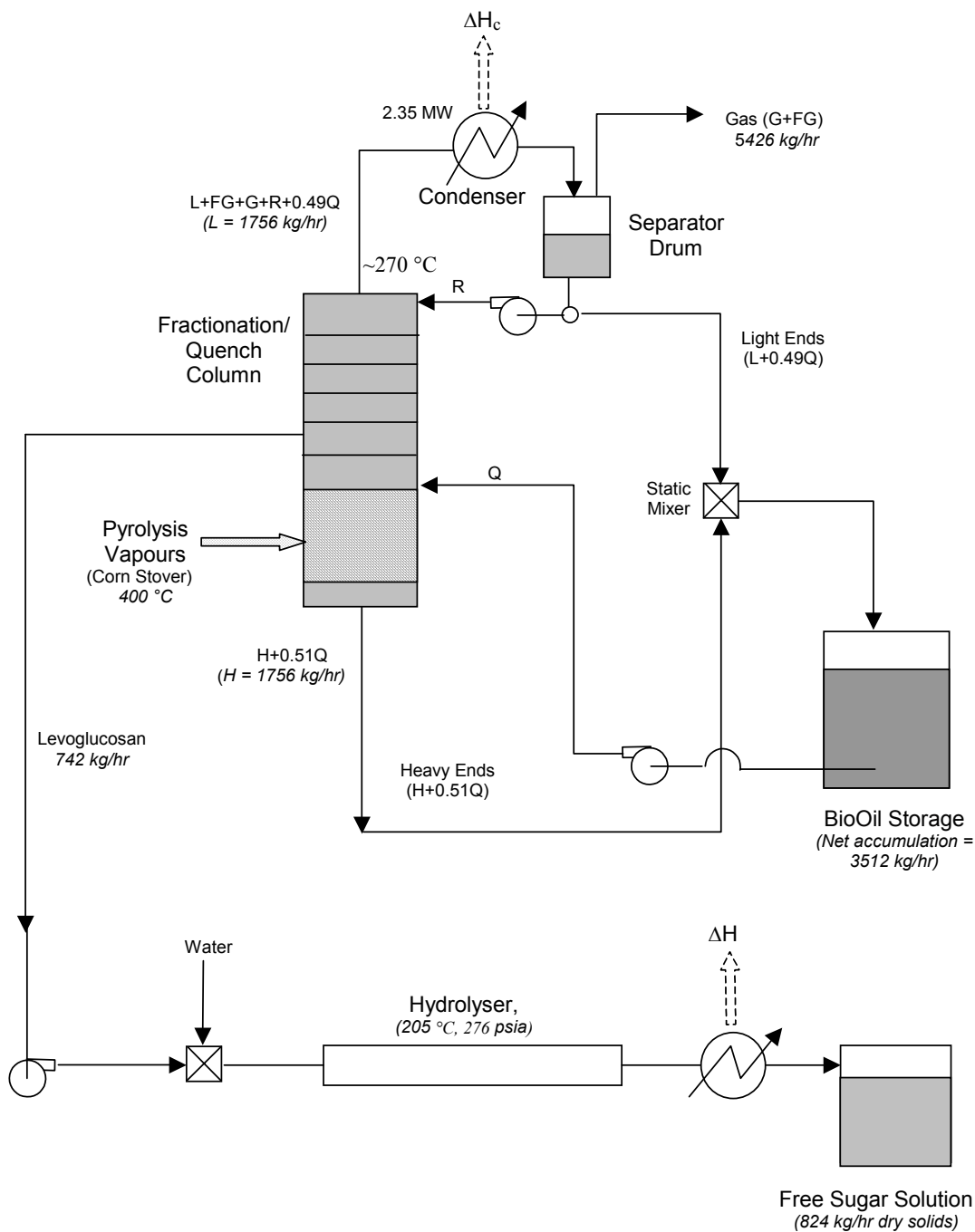
Assumptions:

- Only levoglucosan is recovered. (Though for a relatively small incremental cost other chemicals could be recovered as well, e.g. acetic acid, phenols.)
- Furthermore, no attempt is made to extract the higher molecular weight, oligomeric, anhydrosugars from the heavy ends.
- Levoglucosan is recovered solely for the purpose of conversion to fermentable sugars.
- The hot levoglucosan is hydrolysed by combining with fresh water and pumping the resulting solution through a tubular reactor at about 205 °C. The resulting sugar solution requires no further treatment before sending for fermentation.
- Heat is recovered only from the overhead condenser.
- The condensing heat exchanger on the exit stream can be a basic shell and tube type since removal of the bottoms containing the heavy tars (pyrolytic lignin) removes the fouling problems they normally cause.
- The system is intended to replace *all* of the current biooil quench and condensation train. (The biooil storage tank is not included.)
- We use methods from Guthrie (*Chemical Engineering*, March, 1969) and Happel and Jordan, (Chemical Process Economics, 2'nd ed., 1975) for preliminary estimation.

Mass and Heat Flows

The raw biomass material is assumed to be de-mineralized corn stover. A rate of 11,883 lb/hr (5390 kg/hr) of corn stover to the pyrolyzer will therefore generate $(0.789+0.079) \times 5390 = 4680$ kg/hr of volatiles (V). The total volatile pyrolyzer effluent (VE) comprise pyrolysis gas (G) at 426 kg/hr, water (W) at 562 kg/hr, organics (O) at 3692 kg/hr, as well as the fluidizing gas (FG) at about 5000 kg/hr.

$$VE = W+G+O+FG$$

The combined water and organics, exclusive of levoglucosan (LG) are arbitrarily assumed to be divided equally between L and H. I.e., we write for the a light fraction (L), a heavy fraction (H).

$$L = 1756 \text{ kg/hr}$$

$$H = 1756 \text{ kg/hr}$$

$$LG = 742 \text{ kg/hr.,}$$

After separation of LG, the residual part of the biooil is H + L.

Levoglucosan production rate is 742 kg/hr assuming 100% recovery efficiency. Since hydrolysis to free sugars involves the addition of a water molecule to each LG molecule, this would correspond to a production rate of 824 kg/hr of free sugars.

A reflux stream R composed of light ends controls the upper temperature of the fractionator temperature while a quench stream Q composed of the components in the stored biooil controls the lower temperatures. The stored biooil is comprised of the primary oil less the anhydrosugars. Its composition is thus 49% L and 51% H.

However, using a heat capacity for biooil of ~3.2 kJ/kg-K and ~1 kJ/kg-K for FG, then considering only the sensible heat, we find that 2320 MJ/hr (i.e. from L+ FG) would be available at ~250 °C at a reference temperature of 20 °C.

These may be compared with the net heat required for pyrolysis which is, very roughly, just the sensible heat required to raise the biomass feed to the pyrolysis temperature, namely about 5000 MJ/hr at ~480 °C. Thus roughly half of the input heat could be recovered usefully under the conditions stated in the previous paragraph.

Unit Operations

Fractionating Column

LG must be separated from H in the lower end of the fractionator and from L in the upper end. In order to estimate the number of plates required for the lower end separation, we consider LG as the light key (LK) and consider Syringaldehyde (BP ~ 305 °C) as the heavy key (HK).

Using the vapour pressure of levoglucosan as the correlation given by Suuberg,

$$\ln P \text{ [torr]} = 32.391 - 14452/T \text{ [K]},$$

the relative volatility is $\alpha_{LK/HK} = 1616/760 = 2.13$. The Fenske equation gives an estimate or the number of trays at total reflux as

$$N_m = \ln\left[\frac{\xi_{LK}(1-\xi_{HK})}{\xi_{HK}(1-\xi_{LK})}\right] \bigg/ \ln\alpha_{LK/HK}$$

where ξ_i is the mass fraction of component i in the overhead product. We assume a target of $\xi_{LK} = 0.98$ and of $\xi_{HK} = 0.20$, thus $N_m = 7$.

In the upper section LG is separated from L. In this case LG is the heavy key and we take IsoEugenol (BP ~ 270 °C) as the light key. Thus $\alpha_{LK/HK} = 760/322 = 2.36$. Taking targets of $\xi_{LK} = 0.80$ and of $\xi_{HK} = 0.02$, then $N_m = 6.2$.

Assuming 0.6 m height per tray and a tray efficiency of 80% gives an overall height of 13*0.6 = 9.8 m for the tray section. We include and additional 2 m disengagement space at the top for an **overall column height of 12 m**.

Since there is a large overburden of non-condensable gas, we size the diameter on the basis that the total gas + vapour flow rate should be in the generally preferred range around 1 m/s. Assume that the entrance temperature is 400 °C and that the vapour density at this temperature is 1.18 kg/m^3, then the volumetric rate at the inlet is (5000+4340)/(3600×1.18) = 2.20 m^3/s, giving a **column diameter of 1.7 m**. (A more accurate analysis should take account of the flooding limits of the trays.)

We estimate the reflux ratio, r, for the upper section of the column using the following correlation:

$$ r \approx r_{LK} = 1.38 / \left[\left(\alpha_{LK/HK} - 1 \right)^{0.9} \left(1 - \xi_{LK} \right)^{0.1} \right]. $$

This gives $r \sim$ **1.6** so the total reflux flow is **R = 2539 kg/hr** (Specific Gravity = 1.1).

The flow Q is estimated from that required to reduce the temperature of the incoming vapour from 400 °C to say 350 °C. We find **Q ~ 503 kg/hr**. (Specific Gravity = 1.2

Heat Exchanger (Condenser)

It is assumed that the condensing heat exchanger on the exit stream can be a basic shell and tube type since removal of the bottoms containing the heavy tars (pyrolytic lignin) removes the fouling problems they normally cause.

There is some question as to whether the condenser should be air or water-cooled. We shall assume that an air heat exchanger is used. It could provide hot air for biomass drying for instance. Another design issue is the desirable level of heat recovery, i.e. the final air temperature. We assume 150 °C.

We assume that the heat of vapourization of the volatile organics is roughly the same as that of levoglucosan as reported by Oja and Suuberg (J. Chem. Eng. Data, (1999) 44, 26-29), namely ~0.62 MJ/kg. Taking account of the water ($\Delta H_{vap} \approx 2.26$ MJ/kg), we find an overall rough heat of vapourization of ~ 1 MJ/kg for the condensable light end stream. (Note that this is very close to the estimated net heat for pyrolysis.)

Using the reflux ratio estimated above, the latent heat to be removed per second is (2.6×1587 + 0.49×503) × 1/3600 = 1.21 MW. The sensible heat to be removed per second at an exit temperature of say 50 °C, which includes that of the gas as well as the vapour, is (5395×1 + 1587×2.6×3.2×(270-50)/3600 = 1.14 MW. Thus the overall condenser duty is **Q$_C$ = 2.35 MW**.

For finned 316 SS tubes, we assume the purchase cost in 1970 $ is .$144(A/ft^2)$^{0.75}$ while the installed cost is $324(A/ft^2)$^{0.750}$ and the overall heat transfer coefficient is U = 90 BTU/(hr.ft^2.°F) based on bare non-finned tube surface. (Happel and Jordan, p. 226.)

Separator Drum

On the basis of the total liquid flow, and assuming 5 mins. hold up time, the required volume is estimated at 0.66 m^3 (23.3 ft^3). Following standard ratios, the length is estimated at 8 ft and the diameter as 2 ft. The material is SS clad carbon steel.

Hydrolyser

The hydrolyser is presumed to be a plain tubular reactor under laminar flow. Very little information is available on the kinetics of the hydrolysis of levoglucosan in high pressure water; neither is the optimum concentration known. We arbitrarily assume a first order rate constant of 20 min^{-1} and a concentration of 50% (equivalent to 55% free sugars in the product) so that only ~ 688 kg/hr of water for dilution would be required. (A small heat input may be required to adjust the temperature here. We neglect this.) Thus the total volume required is estimated to be 420 L (110 gal) so only about 10 ft of 6" tube would suffice. The associated pump would have negligible cost.

Storage Tank (Sugar solution)

Assume a carbon steel tank (MPF = 2.1). A 20,000 gal tank would have more than three days production storage capacity.

Pumps

Pumps are required for both the efflux flow R and the quenching flow Q, which are ~2 and ~ 10 gal/min, respectively.

Summary

The overall installed cost is $545,480 in 2002 $ as detailed in the table below.

Equipment	BC ($ 1968)	MPF	MF	UF 2002	Uninstalled Cost ($ 2002)	Installed Cost ($ 2002)
Empty Column (316 SS clad)	10,350	2.25	4.23	3.39	78,975	192,270
Trays: sieve, 2 ft. spacing (SS)	3,274	1.7	1	3.39	18,866	18,870
Separator Drum (316 SS clad)	1,145	2.25	4.23	3.39	8,736	36,950
Heat exchanger (SS tubes/CS shell)	23,520	1	-	3.39	79,750	179,430
Hydrolyser, horizontal mount (CS, 276 psi)	330	1.35	3.18	3.39	1,100	3,950
Sugar Soln. Tank (CS; 20,000 gal)	7,540	1	-	3.39	25,560	102,280
2 Liquid Pumps (cast iron)	1,020	1	3.38	3.39	3,460	11,730
Total						**$545,480**

BC = Base cost
MPF = Material & pressure factor
MF = Module factor
UF = Update factor

Installed Cost = UF.BC(MPF+MF-1)
Uninstalled Cost = UF.BC.MPF

Appendix B. Example Pro Forma Model

- Oat Hulls
- Value-added Chemicals Option
- Renewable Energy Credit = $18/MWh

Summary

Project Description	
Name of Project	Alliant IPCC Project
Primary Fuel Type	Biomass
Number of Units	1
MW Output	7.655
Steam Production (lb/hr)	0

Pro Forma Summary			
	Total	Senior	Subordinate
Avg Debt Service Coverage Ratios	3.048	3.048	#N/A
Min Debt Service Coverage Ratios	2.304	2.304	0.000
	Project	Investor 1	Investor 2	Investor 3
NPV of A-T Cash Flows During Operation ($1000)	11,655	11,655	0	0
+ NPV of Year 0 A-T Cash Flows ($1000)	0	0	0	0
- FV of Equity Contributions ($1000)	11,655	11,655	0	0
= Project NPV ($1000)	(0)	(0)	0	0
Internal Rate of Return (%)	11.67%			
NPV Discount Rate (%)	12.00%			

Project Schedule		
	COD 1	COD 2
Start of Construction	Jan-2003	#N/A
Construction Period (months)	24	#N/A
Start of Commercial Operation	Jan-2005	#N/A
End of Commercial Operation	Dec-2032	#N/A

Project Sensitivity Scenarios		
Sensitivity	Change	IRR

Sources of Funds at Closing			Debt Term (Years)	Amount ($1000)
	Percentage	Interest Rate		
Tranche A	50.00%	8.00%	28	10,453
Tranche B	0.00%	0.00%	0	0
Tranche C	0.00%	0.00%	0	0
Investor 1	50.00%			10,453
Investor 2	0.00%			0
Investor 3	0.00%			0
Total Sources of Funds	100.00%			20,905

Capital Budget / Uses of Funds	Amount ($1000)
EPC	19,666
Owner's Costs	0
Finance and Legal Costs	0
Capitalized Reserve Funds	1,240
Interest During Construction	0
Contingency	0
Total Capital Budget/Use of Funds	20,905

Project Description

Name of Project	Alliant IPCC Project
Number of CODs	1
Operating Period (max 30 years)	28
Monetary Unit	United States Dollar ($)
English or Metric Units	English
Percent of Year On-peak/Full load operation	100.0%
Percent of Year Off-peak/part load operation	0.0%

Cash Flow

Interest Rate (Before-Tax) on Reserve Funds (%)	8.0%
No. Months of P&I for Debt Service Fund	6
No. Days of Oper. Exp. For Working Capital Fund	45
Discount Rate (%)	12.0%
General Inflation (%)	2.0%

Construction Period Information

	COD 1	COD 2
Construction Period (max 48 months)	24	0
Interest Rate on Construction Loan (%)	0.0%	0.0%
% Debt	50.0%	100.0%
% Equity	50.0%	0.0%
Start of Commerical Operation (month/year)	Jan-2005	#N/A

Debt/Equity Information

		COD 1 ($1000)	COD 2 ($1000)	Total ($1000)
Percent Debt:				
(Senior) % Tranche A	50.0%	10,453	0	10,453
(Senior) % Tranche B	0.0%	0	0	0
(Subordinate) % Tranche C	0.0%	0	0	0
Total Long Term Debt	50.0%	10,453	0	10,453
Percent Equity:				
% Investor 1	50.0%	10,453	0	10,453
% Investor 2	0.0%	0	0	0
% Investor 3	0.0%	0	0	0
Total Equity Contribution	50.0%	10,453	0	10,453
Total Sources of Funds	100.0%	20,905	0	20,905

Project Use of Funds

		Depr. %	COD 1 ($1000)	COD 2 ($1000)	Total ($1000)
EPC Cost		100%	19,666	0	19,666
Owner's Costs:					
Site Purchase		0%	0	0	0
Infrastructure		100%	0	0	0
Start Up		100%	0	0	0
Spare Parts		100%	0	0	0
Construction Management Expenses		100%	0	0	0
Construction Management Fees		100%	0	0	0
Consultants		100%	0	0	0
Subtotal Owner's Costs		100%	0	0	0
Financing Fees:					
Up Front (%)	0.00%	100%	0	0	0
Commitment (%)	0.00%	100%	0	0	0
Subtotal Financing Fees		0%	0	0	0
Legal Fees		100%	0	0	0
Contingency (Owner's Reserve)		100%	0	0	0
Initial Contribution to Funds:					
Debt Service		0%	473	0	473
Working Capital		0%	767	0	767
Major Maintenance		0%	0	0	0
Subtotal Initial Contributions to Funds		0	1,240	0	1,240
Interest During Construction		100%	0	0	0
Total Capital Cost			20,905	0	20,905
Total Depreciable Amount			19,666	0	19,666

Financing Terms

	Tranche A (Senior)	Tranche B (Senior)	Tranche C (Subordinate)
Term (year)	28	0	0
Payments per Year (1,2,4)	1	2	4
Interest Rate (%)	8.0%	0.0%	0.0%
First Payment (month/year)*	Dec-2005	#N/A	#N/A
Last Payment (month/year)*	Dec-2032	#N/A	#N/A
Payment ($1000)	946	0	0
Equal P&I Payments?	Equal P&I	Equal P&I	Equal P&I
Other Fees (%)	0.0%	0.0%	0.0%
* Debt payments occur at end of each payment period and last day of displayed month			

Depreciation & Tax Parameters

Taxes (%):	
Federal	0.0%
State	0.0%
Local	0.0%
Effective (Overridden)	30.0%
Depreciation Method:	
Double Declining Balance (DDB)	
Depreciation Asset Life (Years)	7
Depreciation Convention	
Normal	
Tax Loss Treatment:	
Carryforward	

Electric/Cogeneration (E/C) Performance

		COD 1/Plant	COD 1+2
Electrical Output:			
Customer 1 (MW)		7.7	0.0
Customer 2 (MW)		0.0	0.0
Customer 3 (MW)		0.0	0.0
Annual Capacity Factor (%)		64.0%	80.0%
Degradation (%)		1.0%	0.0%
Fuel 1:	Biomass		
Heat Rate, on-peak (Btu/kWh)	Input	12,000	#DIV/0!
Heat Input Rate, on-peak (MBtu/hr)	Calculate	0.0	0.0
Heat Rate, off-peak (Btu/kWh)	Input	12,000	10,000
Heat Input Rate, off-peak (MBtu/hr)	Calculate	91.9	0.0
Heat Rate Degradation (%)		1.0%	3.0%
Percentage of Operation (%)		100.0%	80.0%
Fuel 2:	Natural Gas		
Heat Rate, on-peak (Btu/kWh)	Input	0	10,000
Heat Input Rate, on-peak (MBtu/hr)	Calculate	0.0	0.0
Heat Rate, off-peak (Btu/kWh)	Input	0	10,000
Heat Input Rate, off-peak (MBtu/hr)	Calculate	0.0	0.0
Heat Rate Degradation (%)		0.0%	3.0%
Percentage of Operation (%)		0.0%	20.0%

Steam Delivery Data

Steam-Only Capacity Factor (%)	0.0%
Steam-Only Delivery Rate (lb/hr)	0
Cogeneration Steam Delivery Rate (lb/hr)	0
Steam Usage Fractions (%):	
Customer 1	0.0%
Customer 2	0.0%
Customer 3	0.0%
Total	0.0%

Steam-Only Fuel Performance

		All Units
Fuel Type		None
Heat Rate (Btu/kWh)	Input	0
Heat Input Rate (MBtu/hr)	Calculate	0.0
Heat Rate Degradation (%)		0.0%

Revenues

Item	Units	Price	Escalation
On-Peak Electric Rates:			
Customer 1	$/MWh	35.000	2.0%
Customer 2	$/MWh	0.000	2.0%
Customer 3	$/MWh	0.000	2.0%
Off-Peak Electric Rates:			
Customer 1	$/MWh	35.000	2.0%
Customer 2	$/MWh	0.000	2.0%
Customer 3	$/MWh	0.000	2.0%
Ancillary Services	$1000	0.000	2.0%
Capacity Rates:			
Capacity Uplift	$/kw-yr	24.000	2.0%
Customer 1	$/kw-yr	0.000	2.0%
Customer 2	$/kw-yr	0.000	2.0%
Customer 3	$/kw-yr	0.000	2.0%
Steam Sales Rates:			
Customer 1	$/klb	0.000	2.0%
Customer 2	$/klb	0.000	2.0%
Customer 3	$/klb	0.000	2.0%
Other Revenues:			
Renewable Energy Credit	$1000	771.994	2.0%
Anhydrosugar Sales	$1000	5800.000	2.0%
Char Sales	$1000	154.165	2.0%
Item 4	$1000	0.000	2.0%
Energy=Fuel + Variable O & M pass-through			No

Non-Fuel Operating Expenses

Item	Units	Price	Escalation	Amount	Units
Variable Expenses:					
Variable O&M	$/MWh	0.00	2.0%		
Water	$/mgy	0.00	2.0%	0	mill gals
Wastewater	$/mgy	0.00	2.0%	0	mill gals
Chemicals	$/ton	0.00	2.0%	0	tons
Ash Disposal	$/ton	0.00	2.0%	0	tons
Start Cost	$/start	0	2.0%	0	starts
Other Variable Expenses:					
Item 1	$1000	0.00	2.0%		
Item 2	$1000	0.00	2.0%		
Item 3	$1000	0.00	2.0%		
Fixed Expenses:					
Fixed O&M	$/kw-yr	0.00	2.0%		
Property Taxes	$1000	0.00	2.0%		
Insurance	$1000	0.00	2.0%		
General & Administrative	$1000	0.00	2.0%		
Other Fixed Expenses:					
Total Fixed & Variable O&M	$1000	4,818.78	2.0%		
Pyrolysis Plant Electricity	$1000	331.77	2.0%		
Biomass Fuel Cost	$1000	1,067.63	2.0%		

Fuel Operating Expenses

	Units	Price	Escalation
Elec/Cogen Fuel 1	$/MBtu	0.000	2.0%
Elec/Cogen Fuel 2	$/MBtu	0.000	2.0%
Steam Fuel	$/MBtu	0.000	2.0%

Operating Year		Year 0	1	2	3	4	5	6	7	8	9	10	11	12
Months of Operation in Year		0	12	12	12	12	12	12	12	12	12	12	12	12
Calendar Year			2005	2006	2007	2008	2009	2010	2011	2012	2013	2014	2015	2016
	Units													
Operating Revenues:														
Electric Sales	$1000	0	1,487	1,517	1,547	1,578	1,610	1,642	1,675	1,708	1,742	1,777	1,813	1,849
Capacity Revenues	$1000	0	184	187	191	195	199	203	207	211	215	220	224	228
Ancillary Services	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Steam Sales	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
RE Cred., Char, VA Chem	$1000	0	6,726	6,861	6,998	7,138	7,281	7,426	7,575	7,726	7,881	8,038	8,199	8,363
Total Operating Revenues	$1000	0	8,397	8,565	8,736	8,911	9,089	9,271	9,456	9,645	9,838	10,035	10,236	10,441
Operating Expenses:														
	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
O&M and Fuel Costs	$1000	0	6,218	6,343	6,469	6,599	6,731	6,865	7,003	7,143	7,286	7,431	7,580	7,732
	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Operating Expenses	$1000	0	6,218	6,343	6,469	6,599	6,731	6,865	7,003	7,143	7,286	7,431	7,580	7,732
Cash Available for Debt Service	$1000	0	2,179	2,222	2,267	2,312	2,358	2,406	2,454	2,503	2,553	2,604	2,656	2,709
Senior Debt Service:														
Interest Payment for Tranche A	$1000	0	836	827	818	808	797	785	772	758	743	727	709	690
Interest Payment for Tranche B	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Principal Payment for Tranche A	$1000	0	110	118	128	138	149	161	174	188	203	219	237	256
Principal Payment for Tranche B	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Other Fees for Tranche A	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Other Fees for Tranche B	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Senior Debt Service	$1000	0	946	946	946	946	946	946	946	946	946	946	946	946
Cash Available for Subordinate Debt Service	$1000	0	1,233	1,277	1,321	1,366	1,413	1,460	1,508	1,557	1,607	1,658	1,710	1,763
Subordinate Debt Service:														
Interest Payment for Tranche C	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Principal Payment for Tranche C	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Other Fees for Tranche C	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Total Subordinate Debt Service	$1000	0	0	0	0	0	0	0	0	0	0	0	0	0
Cash Flow Net Total Debt Service	$1000	0	1,233	1,277	1,321	1,366	1,413	1,460	1,508	1,557	1,607	1,658	1,710	1,763
Income Taxes	$1000	0	0	0	0	0	0	0	0	0	0	0	40	606
After-Tax Operating Cash Flow	$1000	0	1,233	1,277	1,321	1,366	1,413	1,460	1,508	1,557	1,607	1,658	1,670	1,158
After Tax Interest Income	$1000	0	69	70	71	72	72	73	74	75	76	77	78	79
Net Reserve Fund Transfers	$1000	0	0	(15)	(16)	(16)	(16)	(17)	(17)	(17)	(18)	(18)	(18)	(19)
After Tax Cash Flow / Funds Available for Distribution	$1000	0	1,302	1,331	1,376	1,422	1,469	1,517	1,565	1,615	1,666	1,717	1,730	1,218
Annual Debt Service Coverage Ratios:														
Senior Debt	Ratio		2.304	2.350	2.397	2.445	2.493	2.543	2.594	2.646	2.699	2.753	2.808	2.864
Subordinate Debt	Ratio		N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
Total Annual Debt Service Coverage Ratio	Ratio		2.304	2.350	2.397	2.445	2.493	2.543	2.594	2.646	2.699	2.753	2.808	2.864
After Tax Cash Flow Distribution:														
Investor 1	$1000	0	1,302	1,331	1,376	1,422	1,469	1,517	1,565	1,615	1,666	1,717	1,730	1,218

	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28
	12	12	12	12	12	12	12	12	12	12	12	12	12	12	12	12
	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032
	1,886	1,924	1,962	2,001	2,041	2,082	2,124	2,166	2,210	2,254	2,299	2,345	2,392	2,440	2,488	2,538
	233	238	242	247	252	257	262	268	273	278	284	290	296	301	307	314
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	8,530	8,701	8,875	9,053	9,234	9,418	9,607	9,799	9,995	10,195	10,399	10,606	10,819	11,035	11,256	11,481
	10,649	10,862	11,080	11,301	11,527	11,758	11,993	12,233	12,477	12,727	12,982	13,241	13,506	13,776	14,052	14,333
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	7,886	8,044	8,205	8,369	8,536	8,707	8,881	9,059	9,240	9,425	9,613	9,805	10,002	10,202	10,406	10,614
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	7,886	8,044	8,205	8,369	8,536	8,707	8,881	9,059	9,240	9,425	9,613	9,805	10,002	10,202	10,406	10,614
	2,763	2,818	2,875	2,932	2,991	3,051	3,112	3,174	3,238	3,302	3,368	3,436	3,504	3,575	3,646	3,719
	670	648	624	598	570	540	508	473	435	394	350	302	251	195	135	70
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	276	298	322	348	376	406	438	473	511	552	596	644	695	751	811	876
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	946	946	946	946	946	946	946	946	946	946	946	946	946	946	946	946
	1,817	1,873	1,929	1,987	2,045	2,105	2,166	2,228	2,292	2,356	2,423	2,490	2,559	2,629	2,700	2,773
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
	1,817	1,873	1,929	1,987	2,045	2,105	2,166	2,228	2,292	2,356	2,423	2,490	2,559	2,629	2,700	2,773
	628	651	675	700	726	753	781	810	841	873	906	940	976	1,014	1,053	1,095
	1,189	1,221	1,254	1,286	1,319	1,352	1,385	1,418	1,451	1,484	1,517	1,550	1,582	1,615	1,647	1,678
	80	81	83	84	85	86	87	88	90	91	92	94	95	96	98	49
	(19)	(19)	(20)	(20)	(21)	(21)	(21)	(22)	(22)	(23)	(23)	(24)	(24)	(25)	(25)	1,756
	1,251	1,283	1,316	1,350	1,383	1,417	1,450	1,484	1,518	1,552	1,586	1,620	1,653	1,686	1,719	3,483
	2.921	2.980	3.039	3.100	3.162	3.226	3.290	3.356	3.423	3.491	3.561	3.632	3.705	3.779	3.855	3.932
	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
	2.921	2.980	3.039	3.100	3.162	3.226	3.290	3.356	3.423	3.491	3.561	3.632	3.705	3.779	3.855	3.932
	1,251	1,283	1,316	1,350	1,383	1,417	1,450	1,484	1,518	1,552	1,586	1,620	1,653	1,686	1,719	3,483