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Techno-Economic Analysis of Biomass Fast Pyrolysis to Transportation Fuels

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Technical Report
NREL/TP-6A20-46586
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Prepared under Task No. BB07.7510

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Foreword

The purpose of this techno-economic analysis is to determine the economics of converting biomass to transportation fuel components via fast pyrolysis. Every effort has been made to place this analysis on an equivalent basis with other biomass conversion technologies analyzed in separate reports by using common assumptions. The process design and parameter value choices underlying this analysis are exclusively based on public domain literature. Accordingly, the results should not be interpreted as optimal performance of mature technology, but as the most likely performance given the current state of public knowledge.

Acknowledgments

This study was made possible by support from ConocoPhillips Company under project number 2007-P-01 and from the National Renewable Energy Laboratory. We would like to acknowledge the U.S. Department of Energy Office of Energy Efficiency and Renewable Energy's Biomass Program. We appreciate the helpful comments of Andy Aden, Christopher Kinchin, and Mary Biddy from NREL, and Richard Jefferys and Lou Burke from ConocoPhillips Company.

List of Acronyms

ASPEN	Advanced Simulator for Process Engineering (software)
bbl	barrel
BTU	British thermal unit
DB	declining balance
DOE	U.S. Department of Energy
ESP	electrostatic precipitator
GGE	gallons of gasoline equivalent
IC	indirect costs
IRR	internal rate of return
IRS	Internal Revenue Service
MACRS	(IRS) Modified Accelerated Cost Recovery System
MM	million
MT	metric ton
NCG	non-condensable gas
NREL	National Renewable Energy Laboratory
PV	product value
TEIC	total equipment installed cost
TPI	total project investment
USDA	U.S. Department of Agriculture
VOCs	volatile organic compounds

Executive Summary

The purpose of this study is to develop techno-economic models for assessment of the conversion of biomass to valuable fuel products via fast pyrolysis and bio-oil upgrading. Liquefaction of biomass by fast pyrolysis and subsequent upgrading of the resulting pyrolysis oil (bio-oil) by hydrotreating and hydrocracking—refinery processes that use hydrogen to remove impurities and break large molecules down to smaller ones—is a promising means for producing renewable transportation fuel. The upgrading process assessed in this study produces a mixture of naphtha-range (gasoline blend stock) and diesel-range (diesel blend stock) products. This study develops techno-economic models and uses them to analyze the economics of two scenarios. In one, hydrogen needed for the upgrade process is produced onsite by reforming bio-oil. In the other, the hydrogen is purchased from an outside source.

Both scenarios are based on a fast pyrolysis plant with bio-oil upgrading using 2,000 metric tons per day (MT/day) of corn stover feedstock. Major assumptions made for this analysis match those of companion analyses for producing transportation fuel from biomass via biochemical and gasification technologies. Product value—defined as the value of the product needed for a net present value of zero with a 10% internal rate of return—is first calculated for a mature industry or n^{th} plant and then adjusted for a pioneer plant or one of the first of its kind.

The study results indicate that petroleum fractions in the naphtha distillation range and in the diesel distillation range are produced from corn stover at a product value of $3.09/gal ($0.82/liter) with onsite hydrogen production or $2.11/gal ($0.56/liter) with hydrogen purchase. These values correspond to a $0.83/gal ($0.21/liter) cost to produce the bio-oil. Based on these n^{th} plant numbers, product value for a pioneer hydrogen-producing plant is about $6.55/gal ($1.73/liter) and for a pioneer hydrogen-purchasing plant is about $3.41/ gal ($0.92/liter). Although these results suggest that pyrolysis-derived biofuels are competitive with other alternative fuels, the technology is relatively immature, resulting in a high level of uncertainty in these estimates.

Capital costs for integrated hydrogen production are estimated at $287 million with a fuel yield of 35 million gallons (134 million liters) of naphtha-range and diesel-range products per year. Capital costs with purchased hydrogen are estimated at $200 million with a fuel yield of 58 million gallons (220 million liters) of naphtha-range and diesel-range products per year.

Sensitivity analysis identifies fuel yield as a key variable for the hydrogen-production scenario. A 5% decrease in the bio-oil to naphtha and diesel fuel yields increases fuel costs by $0.80/gal ($0.21/liter) for the hydrogen production scenario and $0.27/gal ($0.07/liter) for the hydrogen purchase scenarios. Biomass cost is important for both scenarios. Changing feedstock cost from $50–$100 per short ton changes the price of fuel in the hydrogen production scenario from $2.57–$3.62/gal ($0.68–$0.96/liter).

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Background

A number of studies analyze the techno-economics of biomass fast pyrolysis to produce bio-oil, but very few explore upgrading of bio-oil to transportation fuel. Also, key assumptions found in the literature for biomass cost, plant capacity, reactor technology, and other variables vary widely among different studies. It is therefore difficult to compare costs from various previous studies without taking into account differences in process assumptions.

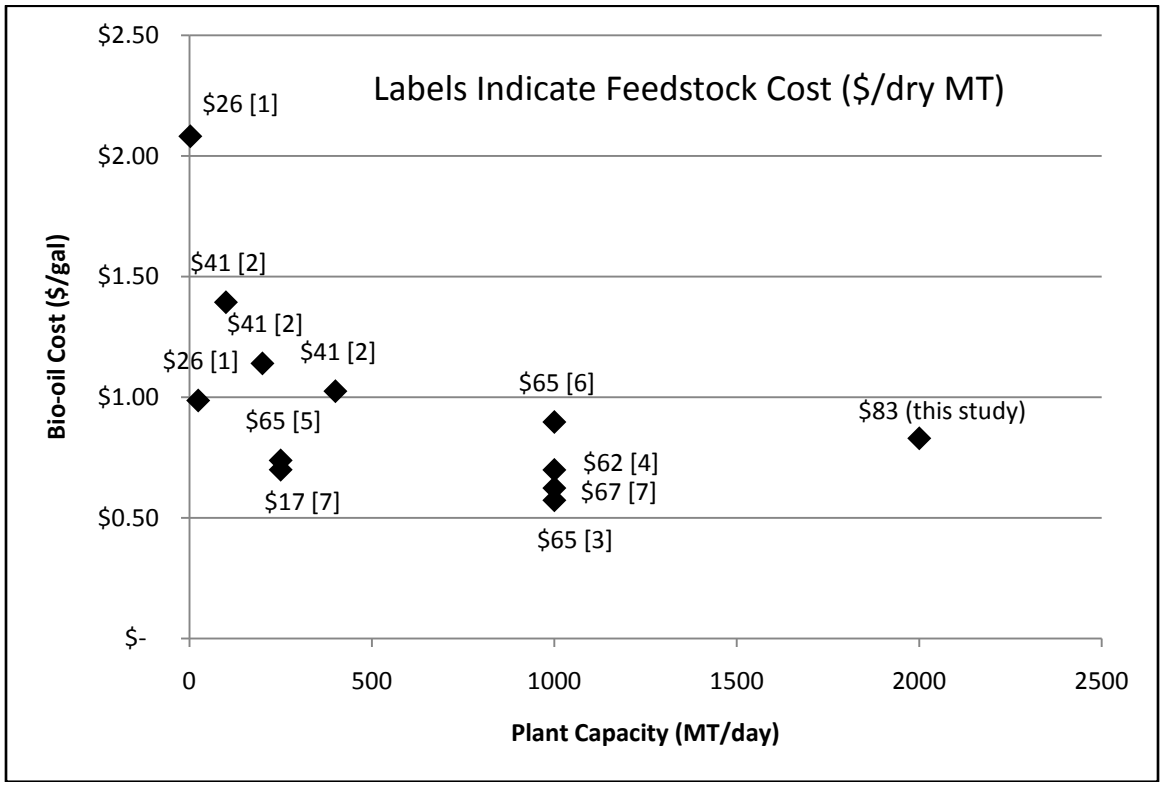


Figure 1. Fast pyrolysis oil costs from previous biomass fast pyrolysis studies [1-7] adapted for inflation from Ringer et al. [8]

Figure 1 shows bio-oil costs for previous biomass fast pyrolysis techno-economic studies. Bio-oil cost estimated in this study is also included [$0.83/gal for 2,000 metric ton per day (MT/day) capacity] in Figure 1. Capital cost estimates for these studies range from $143 million [6] to $37 million [7] for 1,000 MT/day capacities. A recent study by the National Renewable Energy Laboratory (NREL) for a 550 dry MT/day wood fast pyrolysis plant producing 28 million gallons (106 million liters) of bio-oil per year finds the capital cost to be $48.2 million and the operating cost to be $9.6 million/year (based on $0.62/gal product value).

Few studies explore the upgrading of bio-oil to naphtha-range and diesel-range products. Although the technology for bio-oil hydroprocessing is based on commercially available equipment, actual implementations are in development. UOP LLC is one of the main

developers of this technology, and they have published various studies on bio-oil upgrading. A 2005 study finds that gasoline from bio-oil is economically attractive if bio-oil is available at $18/bbl ($0.43/gal) and crude oil sells for $50/bbl [9]. A recent article estimates the cost of naphtha-range and diesel-range fuel from corn stover to be $1.80/gal [10].

Model and Process Description

Biomass fast pyrolysis is a thermochemical process that converts feedstock into gaseous, solid, and liquid products through the heating of biomass in the absence of oxygen. A computational model is developed in this study to simulate this process in order to conduct a techno-economic study of transportation biofuels via fast pyrolysis and bio-oil upgrading. To accomplish this goal, Aspen software is employed to calculate mass and energy balances and economic costs related to the process.

This model is based on the conversion of corn stover to naphtha-range and diesel-range distillation fractions (gasoline blend stock and diesel blend stock, respectively, hereinafter referred to as naphtha and diesel). For modeling, the upgraded pyrolysis oil products are modeled as C8 and C10 hydrocarbons. The biomass plant assumed here processes 2,000 dry MT/day of corn stover using common equipment found in thermochemical conversion facilities. Modifications to existing equipment will be necessary to develop this process; major changes are discussed in this report. The impacts of process uncertainties are considered in the sensitivity and pioneer plant analyses. General processing steps include biomass pretreatment, fast pyrolysis, solids removal, bio-oil collection, char combustion, and bio-oil upgrading.

An overall description of the biomass fast pyrolysis process to produce naphtha and diesel is shown in Figure 2. The hydrogen production scenario employs optional equipment to generate requisite hydrogen. Biomass with 25% moisture content is dried to 7% moisture and ground to 3-mm-diameter size prior to being fed into a fluid bed pyrolyzer operating at 480°C and atmospheric pressure. Standard cyclones remove solids consisting mostly of char particles entrained in the vapors exiting the pyrolyzer. Vapors are condensed in indirect contact heat exchangers, yielding liquid bio-oil that can be safely stored at ambient conditions prior to upgrading to transportation fuels. Non-condensable gases are recycled to the pyrolysis reactor after being combusted to provide process heat. This analysis assumes that pyrolysis solid products are sent to a combustor to provide heat for the drying and pyrolysis process. Excess solids consisting of char are sold as a low-heating-value coal substitute. Bio-oil upgrading, which is discussed in the hydroprocessing section, generates fuel compatible with existing infrastructure.

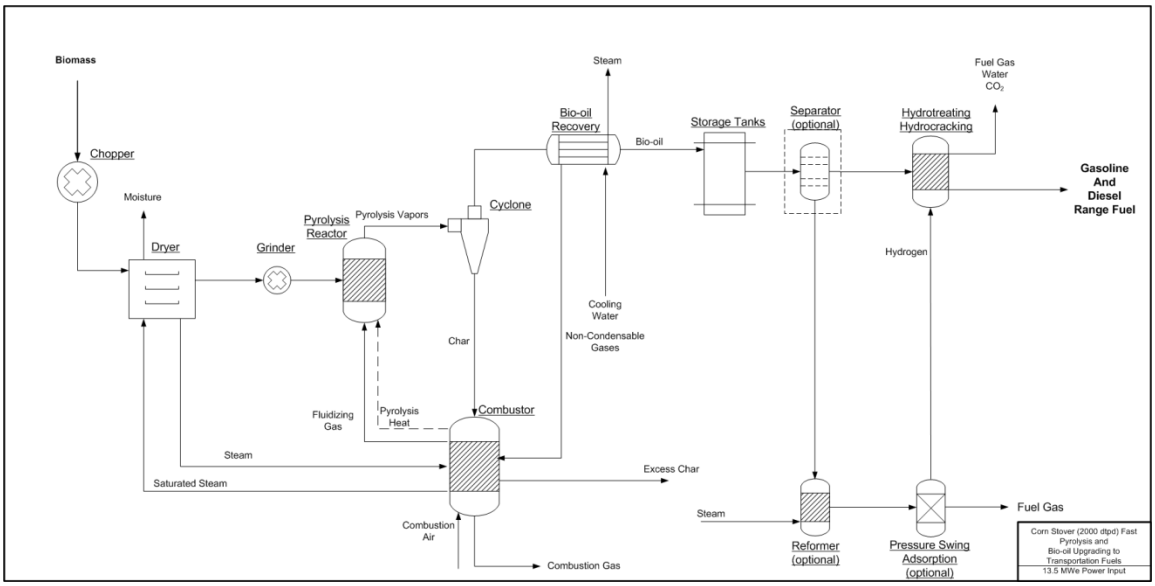


Figure 2. Combined biomass fast pyrolysis and hydrogen production/purchase upgrading process diagram

Biomass condition, as typically delivered, is an important factor not studied in detail in this analysis. Typical feedstock collection methods remove significant soil matter with the biomass from the ground. Soil matter reduces the thermal value of biomass and poses maintenance difficulties for combustion equipment. Ash content can cause fouling and plugging of high-temperature equipment. Minerals catalyze thermal decomposition reactions that are detrimental to the production of quality pyrolysis oil. Biomass washing using water or acid-removal techniques can reduce alkali content in biomass [11]. Mineral effects are not considered in this study because of the limited knowledge of the mechanisms by which alkali content affects pyrolysis yields.

The upgrading process considered in this study is bio-oil hydrotreating and hydrocracking. Hydrotreating and hydrocracking (hydroprocessing) are commonly employed in the petroleum industry to remove undesired compounds such as sulfur from crude oil and to break large hydrocarbon molecules to produce clean naphtha and diesel. Bio-oil typically contains significant quantities of oxygenated compounds that are undesirable for combustion in vehicle engines. Hydrotreating can convert oxygen found in bio-oil to water and carbon dioxide molecules, leaving hydrocarbons that are suitable for internal combustion engines. Complex hydrocarbon compounds are found in bio-oil, and hydrocracking is a potential method to decompose these heavy compounds into naphtha and diesel.

Process modeling is accomplished by employing Aspen Plus software to develop mass and energy calculations. Assumptions and operating conditions are taken from the literature and experimental data when available. Economic analysis uses a combination of Aspen Icarus software equipment cost and sizing and spreadsheet investment analysis calculations. Major assumptions made for this analysis match those of companion analyses for producing transportation fuel from biomass via biochemical and gasification technologies [12, 13]. See Appendix D for more details.

Design Basis

The purpose of the assessed process is to convert biomass into liquid fuels suitable for transportation applications. This is achieved by converting biomass into bio-oil, which is subsequently upgraded to transportation fuels. The design-basis model employs nine distinct sections as described in Table 1.

Table 1. Process Model Sections, Descriptions, and Key Assumptions

Section Name	Section Description	Key Assumptions
Chopping	Particle size reduction to 10 mm	Incoming biomass average size of 10 to 25 mm
Drying	Biomass drying to 7% moisture	Steam drying at 200°C
Grinding	Particle size reduction to 3 mm	Incoming biomass maximum size of <10 mm
Pyrolysis	Biomass conversion to pyrolysis products	480°C and 1 atm 2.75 kg of fluidizing gas per kg of biomass Heat provided by char combustion
Solids Removal	Removal of entrained solid particles from vapor stream	~90% particle removal
Bio-Oil Recovery	Collection of condensing vapors	Rapid condensation to about 50°C 95% collection of aerosols
Storage	Storage of bio-oil and char	4 weeks storage capacity
Combustion	Provides process heat and steam generation	120% excess air combustion 1,100°C gas temperature 200°C steam generation
Hydroprocessing	Upgrading of bio-oil to naphtha-range and diesel-range product fractions	Hydrogen production from oil by aqueous phase reforming P >1,000 psia and T >300°C

Biomass is modeled on a proximate and ultimate analysis basis. There is scarce information in the literature that is specific to corn stover pyrolysis. Nonetheless, this process feedstock is modeled using information from Table 2. This corn stover analysis is adapted from U.S. Department of Agriculture (USDA) experimental data [14]. The ash content value is specified as 6% to meet the requirements of this study, and other values are adjusted accordingly. Char analysis is based on laboratory results [15], shown in Table 3, and these values are not modified in the model.

Table 2. Corn Stover Ultimate and Proximate Analysis [14]

Ultimate Analysis (dry basis)	
Element	**Value (wt %)**
Ash	6
Carbon	47.28
Hydrogen	5.06
Nitrogen	0.8
Chlorine	0
Sulfur	0.22
Oxygen	40.63
Proximate Analysis (wet basis)	
Element	**Value (wt %)**
Moisture	25.0
Fixed Content	17.7
Volatile Matter	52.8
Ash	4.5

Table 3. Char Ultimate and Proximate Analysis [15]

Ultimate Analysis (dry basis)	
Element	**Value (wt %)**
Ash	33.3
Carbon	51.2
Hydrogen	2.12
Nitrogen	0.45
Chlorine	0.471
Sulfur	0.935
Oxygen	11.5
Proximate Analysis (dry, ash-free)	
Element	**Value (wt %)**
Moisture	0
Fixed Content	51.21
Volatile Matter	49.79
Ash	0

Biomass pyrolysis generates a large variety of organic and inorganic compounds that make modeling efforts difficult. Hundreds of compounds have been identified in bio-oil— the primary fast pyrolysis product [16]. A common approach is to employ model compounds to represent chemical groups based on their significance and quantity. This model adapts pyrolysis oil and gas composition from research by NREL as described in the "Pyrolysis" section of this report [8].

Two models are developed to study the performance of biomass pyrolysis for different scenarios: a hydrogen production scenario employing bio-oil reforming to generate requisite hydrogen for bio-oil upgrading, and a hydrogen purchase scenario using merchant hydrogen for bio-oil upgrading. Pyrolysis is a flexible process that can be designed with numerous configurations and scaled to various capacities. Small-scale

pyrolysis is suitable for distributed processing scenarios that could lower costs associated with biomass transportation. Table 4 shows a description of the scenarios explored in this study.

The source of hydrogen for the hydrogen purchase scenario is not specified, although non-fossil energy sources are preferred to achieve the highest reductions in greenhouse gas emissions associated with hydrogen production. The advantage of this approach is that it maximizes the conversion of carbon in biomass to biofuels [17].

These scenarios employ many of the same process sections described in Table 1 and share most of the general assumptions. Assumptions for biomass fast pyrolysis to generate bio-oil are identical for both scenarios; the scenarios only differ by the bio-oil upgrading technology. Hydrogen production fast pyrolysis and oil upgrading employs a portion of the bio-oil produced to generate the required hydrogen for oil hydroprocessing. Additional equipment, including a reformer and gas compressor, are required by the hydrogen production system. The second scenario forgoes the additional investment by purchasing hydrogen from a remote source. Scenarios are based on 2,000 MT/day corn stover input.

Table 4. Corn Stover (2,000 MT/day) Fast Pyrolysis Scenarios

Scenario	Description
Hydrogen Production Fast Pyrolysis with Bio-Oil Upgrading	Large-scale pyrolysis with oil hydroprocessing employing hydrogen derived from bio-oil reforming
Hydrogen Purchase Fast Pyrolysis with Bio-Oil Upgrading	Large-scale pyrolysis with oil hydroprocessing employing off-site generation of hydrogen

Chopping/Grinding

Delivered feedstock typically requires processing prior to being fed into a pyrolysis reactor to avoid penalties that reduce yields and increase heat requirements. Mechanical particle-size reduction and drying are commonly used in thermochemical processes.

Grinding biomass is an expensive and energy-intensive process. Grinding costs can add up to $11/MT of biomass [18]. Specific energy requirements can vary based on equipment and feedstock conditions. A common assumption is that 50 kWh of energy is required per ton of ground biomass [19]. This model employs research by Mani et al. [20] that correlates the grinder screen size to the energy requirement for a hammer mill based on various types of biomass, including corn stover. According to their model, the energy consumption for grinding biomass from a mean chop size of 7.15 mm to between 3.5 and 0.5 mm is approximated by the following equation:

$$\text{Energy [kWh} * \text{ton}^{-1}] = 5.31 * \text{size}^2 - 30.86 * \text{size} + 55.45$$

There are various advantages and disadvantages to using hammer mills for biomass grinding. Hammer mills can employ various screen sizes and work with friable material such as fiber. They incur low capital costs and require minimal maintenance. On the other hand, disadvantages of hammer mills are that they create excess noise and pollution; they

are less efficient than other types of grinders; and they produce a less uniform particle size output than roller mills do [21].

Drying

Feedstock drying is very important for thermochemical processes. Moisture embedded in the feed consumes process heat and contributes to lower process yields. For reasonable pyrolysis performance, moisture content of less than 7% is recommended [22]. Biomass drying typically requires about 50% more energy than the theoretical minimum of 2,442 kJ per kg of moisture evaporated [23].

Dryers can be generally classified as direct or indirect dryers based on how heat is provided. Direct drying involves contact between the heating medium and the feed. Direct dryers can further be classified as either air or superheated-steam dryers. Most commercial dryers employ heated air or process gas to dry the feed. An example of an air dryer is the rotary dryer, which has the advantage of being less sensitive to particle size and can accept hot flue gases. An important disadvantage of air dryers is the potential fire hazard due to the nature of their operation. On the other hand, steam dryers pose less of a fire hazard and emit no air emissions. The disadvantages of steam dryers are higher capital costs and small particle size requirement [24]. This study assumes that biomass is steam dried to 7% moisture.

Steam dryers can employ excess steam generated by a process plant. Thermochemical plants typically require steam as a means to provide, or remove, heat from different equipment. Harmful volatile organic compounds (VOCs) can, however, become entrained in the evaporating moisture. The implication is that dryer steam will require water treatment to reduce the accumulation of large quantities of unwanted chemicals and to prevent the release of these compounds once the steam is discarded. Unfortunately, there is not enough information to properly model the release of VOCs from biomass drying at this time.

Pyrolysis

Fast pyrolysis of biomass is a thermal process that requires temperatures near 500°C, rapid heat transfer, and low residence times. Various reactor designs have been proposed for this process [25]. Because of concerns over the scalability of existing reactor designs, this study assumes that multiple 500 MT/day reactors are employed in parallel. This size is selected based on assumptions from a report by NREL [8]. Commercial units as large as 200 MT/day are currently in operation. Pyrolysis product distribution is adapted from USDA data [14] using the bio-oil and non-condensable gas (NCG) composition shown in Table 6. Bio-oil and NCG composition is modified from a previous NREL analysis [8]. Bio-oil compounds are selected based on available Aspen Plus software compounds and may not share the same properties as compounds selected by NREL. Table 5 shows various pyrolysis yields for corn stover [14, 26]. Table 6 includes the initial pyrolysis product yields employed in this study. The final yield is adjusted to ensure mole and mass balance.

Table 5. Corn Stover Fast Pyrolysis Yields

Material Yields (wt % dry basis)	NREL [26]	NREL [26]	USDA [14][a]
Non-Condensable Gas	14.3	11.7	21.9
Oil	57.6	55.0	61.6[b]
Water	4.9	7.9	-
Char/Ash	19.4	19.5	17
Total	**96.2**	**94.1**	**100**

[a] Current results adjusted from these yield values. [b] Includes water content.

The USDA corn stover fast pyrolysis yield data are listed in Table 5. This study employs a more detailed pyrolysis product composition, shown in Table 6, as a starting point for pyrolysis yield calculations. The pyrolysis product composition reported in this study varies from the original references to achieve a mole balance. For example, the USDA oil yield is listed as 61.6 wt %, but calculations estimate oil yields of 63 wt % (dry basis). The USDA corn stover had a moisture content of 2.5% at the pyrolysis reactor, while our analysis assumes 7% moisture content, which increases the combined water and oil yield to 72 wt % (about 4% of reaction water is generated during the pyrolysis process).

Table 6. Pyrolysis Product Composition

Gas Compounds	Composition (kg/100 kg of dry biomass)
Carbon Dioxide	5.42
Carbon Monoxide	6.56
Methane	0.035
Ethane	0.142
Hydrogen	0.588[a]
Propane	0.152
Ammonia	0.0121
Bio-Oil Compounds	
Acetic Acid	5.93
Propionic Acid	7.31
Methoxyphenol	0.61
Ethylphenol	3.80
Formic Acid	3.41
Propyl-Benzoate	16.36
Phenol	0.46
Toluene	2.27
Furfural	18.98
Benzene	0.77
Other Compounds	
Water	10.80
Char/Ash	16.39

[a] Corrected to 0.02 kg/kg of biomass based on engineering judgment.

Yield adjustments from the original analysis are accomplished by using yield factors. Different factors are applied to the yield of individual compounds until a 100% mole

balance within 1% is achieved. This effort is made to maintain carbon, hydrogen, oxygen, and nitrogen mole balance plus ash mass balance throughout the model. Closer attention is given to carbon, hydrogen, and oxygen because these are the most relevant elements in both the feedstock and final products.

Cleanup

Gases exiting from the pyrolysis reactor contain entrained particles of various sizes. Compared to particles generated from gasification, pyrolysis particles can be much smaller, with sizes less than 25 microns. The particle size is important because it affects the design and performance of cleaning equipment such as cyclones and filters. This model assumes that a set of parallel cyclones is employed to remove 90% of entrained char particles. Baghouse filters can be considered as a secondary collection unit, but conventional baghouse filters require modifications to be employed in pyrolysis applications. Even then, they may still cause yield reductions and require costly maintenance due to coking and vapor condensation on the filter surface.

Char collected from the cleanup section is sent to the combustion section, where it is employed to provide process heat. A portion of the char is burned, while the rest is collected and sold as a by-product.

Oil Collection

There are various possible approaches to the collection of pyrolysis oil. To collect high-quality oil and maintain high yields, vapors should be condensed within fractions of a second after exiting the pyrolysis reactor. Longer residence time allows secondary reactions to take place in the gas phase and reduces the quantity of oil collected.

To accomplish rapid condensation of pyrolysis vapors, this model employs an indirect heat exchanger to transfer heat from the vapors to a water stream. This design allows for the generation of excess steam. Although not considered in this report due to lack of reliable data, staged condensation of bio-oil allows for the collection of oil fractions with attractive properties. A simple example is to condense a majority of water in a specific condenser and a higher concentration of oil in a different condenser. This takes advantage of the fact that different compounds will condense at different temperatures in similar fashion to the crude distillation process. After most of the oil is condensed, an electrostatic precipitator (ESP) unit collects remaining droplets known as aerosols by using high voltage charges. The formation of aerosols is a complex process, with scarce information available on the mechanism involved. This model simply assumes that any remaining char entrained in the vapors is collected in the ESP unit.

NCGs include significant amounts of methane and other combustible gases. NCGs are sent to the combustor to provide heat for biomass drying and the pyrolysis reaction. Combustion gases are then recycled to the reactor on a 1.6 kg gas/kg dry biomass ratio to provide process heat and aid in fluidizing the reactor.

Storage
Bio-oil and char are collected in the storage section, which can store up to 4 weeks of product. Bio-oil storage equipment must be made of stainless steel material to prevent corrosion from bio-oil acids. Char contains volatile material and when handled improperly can pose a fire hazard. Furthermore, the small size of char particles poses an inhalation hazard for people handling the material. Storage for the naphtha and diesel products is similarly sized for 4 weeks of capacity.

Combustion
Process heat is required to operate the pyrolysis reactor. Heat can be provided to small-scale reactors by employing guard heaters with insulation to prevent heat loss. Large-scale reactors require a direct form of heating, which could consist of using hot gases as the fluidizing agent. A possible scheme is to combust pyrolysis gases in the combustion section prior to recycling back to the pyrolysis reactor.

The current model assumes that recycled NCGs and a fraction of pyrolysis char (27.5 MJ/kg [15]) are combusted to provide the necessary process heat for the pyrolysis process and steam generation. Combustion air, at 90% of the stoichiometric requirement, is sent to the combustion reactor. Combustion gases are cooled to provide additional heat to generate steam. Finally, cyclones collect ash from the combustion gases at a solids disposal cost of $18/short ton [27].

Hydroprocessing
Hydrotreating is an exothermic process commonly used in the oil industry to selectively remove impurities that could affect downstream equipment. Hydrotreating takes place in a hydrogen-rich environment (about 95 mol % or 5% by weight). Typical process conditions for hydrotreating are 7–10 MPa (1,000–1,500 psi) pressure and 300°−400°C using a cobalt-molybdenum catalyst.

Hydrocracking breaks down heavy molecules into shorter chains. For example, heavy hydrocarbons with 30 or more carbon atoms can be split into chains within the diesel (C_{12}) or gasoline (C_8) range. Process conditions are a bit more severe than for hydrotreating, with pressures of 10–14 MPa (1,500–2,000 psi) and temperatures of 400°−450°C using a nickel-molybdenum catalyst.

Bio-oil contains a large variety of heavy and light compounds. Bio-oil includes a water-soluble aqueous phase that can be reformed to produce the required hydrogen and contains heavy molecules that can be hydrocracked to lighter molecules. Bio-oil may be suitable for both hydrotreating at an oil refinery where hydrogen can be provided separately and hydrogen production processing that employs the oil aqueous phase to generate hydrogen. In 2005, UOP LLC published one of the few publically available reports on bio-oil hydrotreating [9]. Their initial report employed bio-oil's aqueous phase to generate hydrogen to process the pyrolysis lignin (heavy) phase. Table 7 shows yields from UOP's 2005 report.

Table 7. Pyrolysis Lignin Hydrotreating/Hydrocracking Yield [9]

Feed	**Wt %**
Pyrolytic Lignin	100
Hydrogen	4–5
Product	
Light Ends	15
Gasoline	30
Diesel	8
Water, Carbon Dioxide	51–52

A schematic of the process proposed by UOP is shown in Figure 3. This process employs gravity separation to separate pyrolysis lignin from the water-soluble bio-oil compounds (aqueous phase oil). Aqueous phase oil is mixed with steam and sent to a high-temperature pre-reformer and converted into syngas. This syngas is then fed into a reformer with methane to produce hydrogen. Various reactions including water-gas-shift take place in the reformer to produce hydrogen. Analysis from Marquevich et al. [28] is employed in this analysis to model the reforming process. The general chemical formula for reactions taking place in the reformer is:

$$C_nH_mO_k + (2n - k) * H_2O \rightarrow nCO_2 + (2n + m/2 - k) * H_2$$

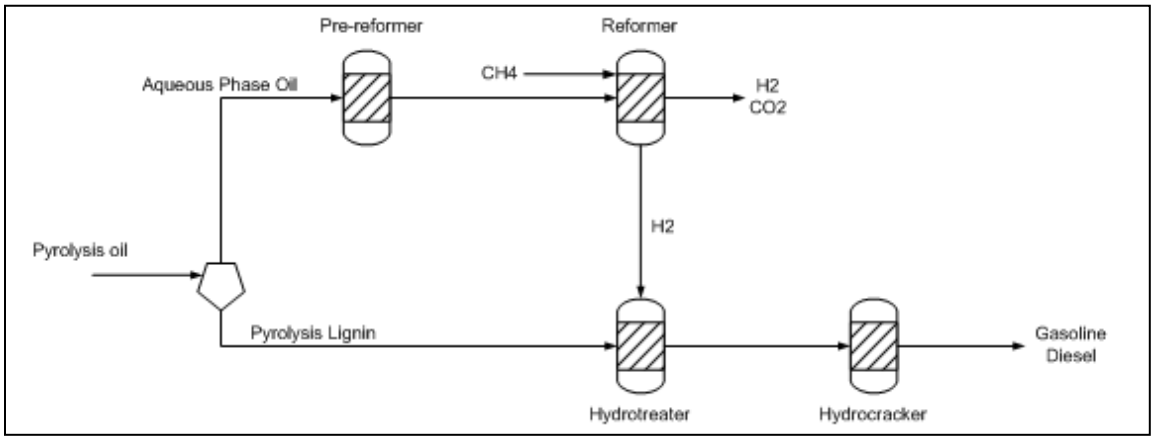


Figure 3. Upgrading of pyrolysis oil to naphtha-range and diesel-range products

This study modifies the original UOP analysis by assuming that a smaller fraction of oil is separated for reforming purposes. Bio-oil's aqueous phase can account for up to 70% of the bio-oil weight, which leaves only 30% for upgrading. This study assumes that the process can be optimized to only separate as much bio-oil as required to produce the needed hydrogen. Estimates show that 38% of the bio-oil needs to be reformed into hydrogen to upgrade the remaining bio-oil.

UOP updated their bio-oil hydrotreating analysis in 2008 [10]. This more recent study explores converting all bio-oil components to transportation fuels, therefore maximizing yields from pyrolysis oils. This approach requires a remote source of hydrogen such as an oil refinery. Yields for this scenario are shown in Table 8.

Table 8. Pyrolysis Oil Hydrotreating/Hydrocracking Yields [10]

Feed	Wt %
Pyrolysis Oil	100
Hydrogen	3–4.5
Products	
Naphtha Range	21
Diesel Range	21
Water, Carbon Dioxide, Lights	60

Hydrocracking typically processes hydrocarbons of long (greater than 30) carbon chain lengths. The longest carbon chain for a compound used in this model has 10 carbon atoms, and therefore hydrocracking is not properly modeled. Nevertheless, a hydrocracking unit is included because this is an important component in the conversion of bio-oil to transportation fuels. Unit size and cost are based on the mass and energy requirements calculated by Aspen Plus software. It is known that metals (P, K, Na, Ca, and other) originating from corn stover will be contained in bio-oil. Traditionally, the concentration of metals is lowered to 10 ppm or less by utilizing a guard bed with sacrificial catalyst to prevent significant poisoning of the hydrotreating catalyst. Typical levels of total metal content in bio-oil can be as high as 500 ppm. It is recognized that technology to remove metals specifically found in bio-oil is not well understood and is under current development. Therefore, in this model a placeholder component for the guard bed is included. Capital costs for the guard bed are estimated as 15% of the hydroprocessing capital expenditure.

Bio-oil hydroprocessing generates significant amounts of fuel gas consisting mostly of methane and lesser amounts of carbon monoxide. A product value of $5 per million Btu is employed in this study, which is comparable to the price of low-cost industrial natural gas [29].

Economics

This study employs Aspen Icarus software to estimate equipment costs and a method from Peters and Timmerhaus to calculate project investment expenditures [30]. Estimates based on this methodology are typically accurate within 30%. The profitability of a given process can be determined from the operating costs and profitability analysis. NREL developed a discounted cash flow rate of return analysis spreadsheet to calculate the product value (PV), which is used in this study.

The plant is designed based on the current state of technology and is assumed to be the n^{th} plant of its kind. Economic analysis for a pioneer plant is developed using risk analysis formulas that take into account the maturity of the knowledge and accuracy of the simulation. The online time is 328 days per year (equivalent capacity factor of 90%). Construction time of less than 24 months is considered. The startup period is 25% of the construction time (6 months). During this period, an average of 50% production is achieved with expenditures of about 75% of variable expenses and 100% of fixed expenses. Contingency is calculated as a 20% factor of total installed equipment cost and indirect costs. A pioneer contingency factor of 30% is employed. Equipment costing data

and installation factors are collected from direct quotation, published data, and Aspen Icarus software evaluation with preference given in the order shown here.

Feedstock cost is assumed to be $75/dry short ton ($83/dry metric ton) and includes delivery cost. Electricity cost is assumed to be $0.054/kWh. Catalyst replacement costs are estimated at $1.77 million/year based on costs for crude oil processing [31].

Working capital is assumed to be 15% of the total capital investment. It is assumed that the product, transportation fuel, will be made and shipped and payment received in 30 days. Annual maintenance materials are 2% of the total installed equipment cost.

General overhead is a factor of 60% applied to the total salaries and covers items such as safety, general engineering, general plant maintenance, payroll overhead (including benefits), plant security, janitorial and similar services, phone, light, heat, and plant communications.

The total plant investment cost is determined by applying overhead and contingency factors to installed equipment costs. Insurance and taxes are considered as 1.5% of the total installed equipment cost.

To determine the product value per gallon of naphtha and diesel, a discounted cash flow analysis is used after determining the major three costs areas: (i) total project investment, (ii) variable operating costs, and (iii) fixed operating cost. A 10% discounted cash flow rate of return is used over a 20-year plant life. The plant is considered 100% equity financed. The IRS modified accelerated cost recovery system is employed to calculate the federal tax return, with depreciation based on a declining balance (DB) method. This allows for the shortest recovery period and largest deductions. The general plant depreciation period is assumed to be 7 years. Property listed with a recovery period of less than 10 years uses a 200% DB depreciation method. A 20-year-recovery-period property uses 150% DB depreciation. State tax is not considered for these calculations because the location of the plant is not specified. Return on investment is calculated on a per-gallon basis and income tax is averaged over the plant life.

In the hydrogen purchase scenario explored for this study, bio-oil is upgraded employing hydrogen from an external source. The purchase price of hydrogen considered here is $1.50/gallon of gasoline equivalent (GGE) or nearly $1.50/kg.

Sensitivity Analysis

Table 9 shows the sensitivity analysis parameters selected for this study. These parameters can have a strong impact on the performance and economics of the process. Capital cost is selected as a sensitivity variable because of the uncertainty associated with the estimate in this study. Upgrading of bio-oil to naphtha and diesel is a developing technology with little public information about the system performance. Upgrading yields can be strongly affected by bio-oil quality and catalyst performance and require further research.

Table 9. Sensitivity Analysis Parameters for Fast Pyrolysis and Upgrading

Sensitivity Analysis	Favorable	Base Case	Unfavorable
Biomass Cost ($/ton)	$50	$75	$100
Bio-Oil Yield (wt/wt feed)	0.7	0.63	0.55
Fuel Gas Credit Value ($/MMBTU)	$10	$5	$2.5
Char Value ($/ton)	$30	$20	$10
Capital Cost (millions $)	$173	$247	$321
Catalyst Cost (millions $)	$0.88	$1.77	$3.53
Fuel Yield (wt/wt feed)	0.3	0.25	0.2
Hydrogen Purchase Scenario			
Capital Cost (millions $)	$120	$172	$223
Fuel Yield (wt/wt feed)	0.47	0.42	0.37
Hydrogen Price ($/GGE)	$1	$1.5	$2

Pioneer Plant Analysis

RAND Corporation analysis was employed to estimate the costs associated with construction and operation of a pioneer plant [32]. This analysis appropriately takes into account the risk associated with building a first-of-a-kind plant or processing unit. The methodology used is based on statistical regressions for plant performance and cost growth. Plant performance is based on the assumption that a pioneer plant can require a few years to operate at peak capacity, and therefore a revenue penalty is assigned due to reduced output. Cost growth estimates the total project investment cost of a pioneer plant, which is typically higher than an equivalent n^{th} plant. Plant performance and cost growth are estimated using the equations shown below and the parameters from Table 10.

Plant Performance = 85.77 – 9.69*NEWSTEPS + 0.33*BALEQS – 4.12*WASTE – 17.91*SOLIDS

Cost Growth = 1.1219 – 0.00297*PCTNEW – 0.02125*IMPURITIES – 0.01137*COMPLEXITY + 0.00111*INCLUSIVENESS – C_1*PROJECT DEFINITION

Table 10. Description of Pioneer Plant Analysis Parameters

Parameters	Range	Definition
NEWSTEPS	≥ 0	Number of new process areas
BALEQS	0–100	Percentage of mass and energy balance equations based on commercial plant data
SOLIDS	0 or 1	A factor based on the presence of solids
WASTE	0–5	A factor of waste disposal
Plant Performance	**0–86**	
PCTNEW	0–100	Percentage of equipment cost for new (under-developed) equipment
IMPURITIES	0–5	A factor of impurities present in the process
COMPLEXITY	0–5	Number of consecutively linked plant areas
INCLUSIVENESS	0–100	Percentage of land purchase/lease, initial plant inventory/parts/catalysts, and pre-operating personnel costs included in the analysis
PROJECT DEFINITION	2–8	A factor of level of detail in the analysis
Cost Growth		

Once the risk analysis parameters are selected and plant performance and cost growth are calculated, the total project investment (TPI) and first year operating costs are calculated as follows:

$$\text{TPI (Pioneer Plant)} = \text{TPI } (n^{th})/\text{Cost Growth}$$

$$\text{Operating Cost } (1^{st} \text{ year}) = \text{Operating Cost} * \text{Plant Performance}$$

The plant performance factor increases by 20% every year until it reaches 100%, at which point the plant is operating at full capacity. It is important to note that if a plant fails to reach a 40% plant performance factor within the first year of operation, the plant is unlikely to achieve full nameplate capacity without significant capital investment. Subsequent plants will likely achieve improved performance. Table 11 includes the selected parameter values and pioneer plant analysis results. The plant performance results suggest that a biomass fast pyrolysis plant based on our assumptions will fail to achieve full nameplate capacity.

Table 11. Selected Pioneer Plant Analysis Parameters for Biomass Fast Pyrolysis and Upgrading

RAND Analysis	Optimistic	Base Case	Pessimistic	Range
NEWSTEPS	2	3	4	-
BALEQS	0%	0%	0%	0–100%
SOLIDS	1	1	1	0,1
WASTE	3	4	5	0–5
Plant Performance	**36.12**	**22.31**	**8.50**	**0–100**
PCTNEW	25.00%	60.00%	75.00%	0%–100%
IMPURITIES	3	4	5	0–5
COMPLEXITY	4	5	6	2–7
INCLUSIVENESS	66%	0%	0%	0%–100%
PROJECT DEFINITION	6	8	8	2–8
Cost Growth	**0.56**	**0.29**	**0.22**	**0-1**

Results

Table 12 includes a comparison of key results from this study. The hydrogen production scenario has higher capital costs than the hydrogen purchase scenario due to additional equipment required for bio-oil upgrading. Annual operating costs are higher for the hydrogen purchase scenario at a hydrogen price of $1.50/GGE (nearly $1.50/kg). Higher fuel yields from upgrading all available bio-oil offset the increase in costs, resulting in a lower product value for the hydrogen purchase scenario. Detailed analyses of these results are provided in the following sections.

Table 12. Summary of n^{th} Plant Cost Results

	Hydrogen Production	Hydrogen Purchase
Capital Cost (millions $)	$287	$200
Annual Operating Cost (millions $)	$109	$123
Fuel Yield (million gallons/year)	35.4	58.2
PV ($/gal gasoline equivalent)	$3.09	$2.11
Pioneer Plant Capital Cost (million $)	$911	$585
Pioneer PV ($/GGE)	$6.55	$3.41

An interesting comparison for these scenarios involves detailing the primary energy flows, which are shown in Figure 4 [30, 31]. The hydrogen production scenario has a biomass-to-fuel efficiency of almost 40% and an overall efficiency of 77% when energy in the excess char and fuel gas are considered. The hydrogen purchase scenario achieves a biomass-to-fuel efficiency of 70%, primarily due to bypassing a bio-oil reforming step. Purchase of external hydrogen is shown here as a negative energy input, which brings the overall efficiency to the same as the hydrogen production scenario (77%).

Efforts are taken to ensure that these scenarios avoid consumption of fossil fuel resources. Purchased electricity can come from nuclear, hydroelectric, or wind power,

which are all prevalent in the Corn Belt. Most available hydrogen is derived from fossil fuel processing, although it is conceivable that hydrogen could be produced by hydrolysis of water with electricity from non-fossil sources. The hydrogen-production scenario produces more fuel gas due to the bio-oil reforming process, which causes a decrease in liquid fuel yield.

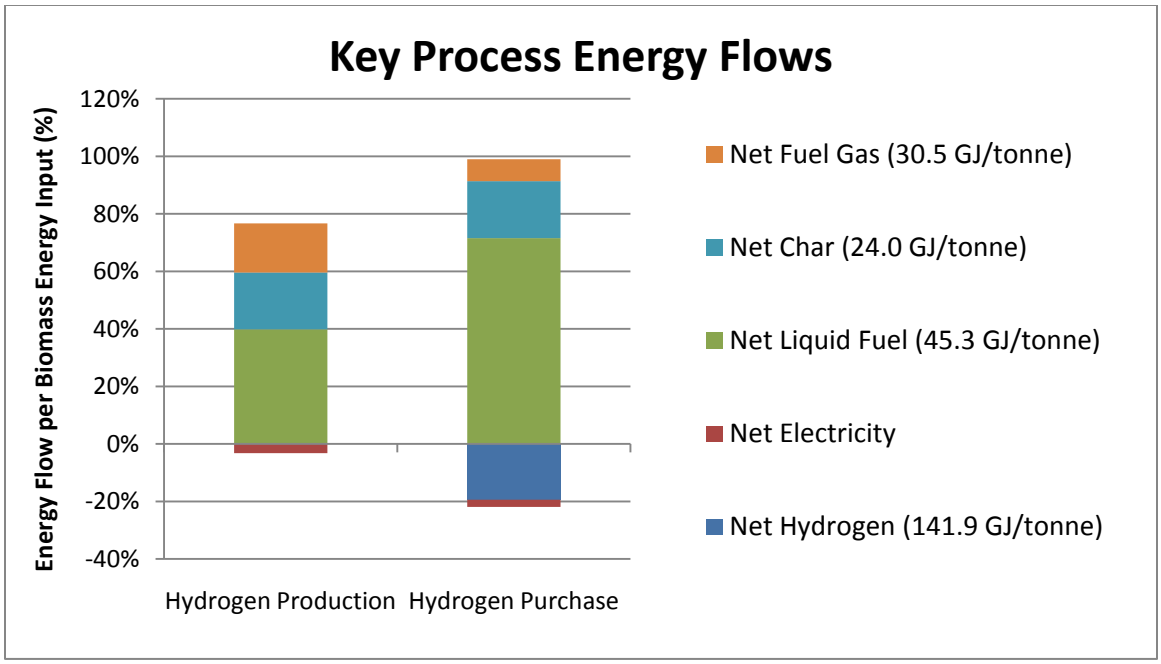


Figure 4. Comparison of process energy flows for 2,000 MT/day biomass fast pyrolysis and upgrading [33, 34]

Hydrogen-Production Scenario

The product value for a 2,000 MT/day corn stover hydrogen-production fast pyrolysis and upgrading plant is $3.09/gal ($0.82/liter) of fuel. This corresponds to a bio-oil production cost of $0.83/gal ($0.22/liter). Capital expenditures for this plant are estimated at $287 million.

Estimated bio-oil yield for a 2,000 MT/day corn stover pyrolysis plant is 104 million gallons (394 million liters) per year, representing 72% yield by weight of the dry biomass input with 15% water content. For the hydrogen-production scenario, 38% of the bio-oil is reformed to produce 1,500 kg/hour of hydrogen. Biomass-to-liquid-fuel efficiency for the hydrogen-production scenario is estimated at 36%. Fuel yield for the hydrogen-production scenario is 35.4 million gallons (134 million liters) of fuel per year.

Capital costs are shown in Table 13. Hydroprocessing incurs the largest expenditure. Large-scale hydroprocessing is typically employed in industry to take advantage of economies of scale. At the plant capacity assumed in this study (2,000 MT/day), hydroprocessing costs are relatively expensive on a per-gallon-of-output basis. Storage

costs include 1 month of fuel storage. The project contingency for both scenarios is assumed as 20% of total direct and indirect costs.

Table 13. Capital Costs for 2,000 MT/day Hydrogen-Production Fast Pyrolysis and Upgrading Scenario

Capital Costs (millions $)	
Hydroprocessing	$48.7
Combustion	$47.3
Pyrolysis and Oil Recovery	$28.0
Pretreatment	$20.2
Utilities	$9.1
Storage	$5.8
Total Equipment Installed Cost	**$159.1**
Indirect Costs	$46.9
(% of TEIC + IC)	20%
Project Contingency	$41.2
Total Project Investment (TPI)	$287.4
Installed Cost per Annual Gallon	$4.50
Total Project Investment per Annual Gallon	$8.12
Lang Factor	5.46

Operating costs for the hydrogen-production scenario are shown in Table 14. Feedstock costs contribute half the cost of producing fuel from biomass in this scenario. Co-product credits, primarily from the sale of fuel gas, generate significant income.

Electricity costs are estimated to contribute $0.164/gal to the cost of fuel. Investment in power generation could yield additional income from the sale of excess electricity, but this scenario is not explored in this study.

Table 14. Operating Costs for 2,000 MT/day Hydrogen-Production Fast Pyrolysis and Upgrading Scenario

Operating Costs (cents/gal product)	
Feedstock	153.8
Electricity	16.4
Solids Disposal	5.1
Catalyst	5.1
Fixed Costs	32.5
Co-Product Credits	-31.9
Capital Depreciation	33.6
Average Income Tax	26.4
Average Return on Investment (10% IRR)	63.5
Total	309.4
Operating Costs (millions $/yr)	
Feedstock	$54.4
Electricity	$5.8
Solids Disposal	$1.8
Catalyst	$1.8
Fixed Costs	$11.5
Co-Product Credits	-$11.3
Capital Depreciation	$11.9
Average Income Tax	$9.3
Average Return on Investment	$22.5
Total	$109.5

Sensitivity analysis results shown in Figure 5 indicate a strong impact from fuel yield. This implies that slight improvements in the bio-oil upgrading process could reduce the cost of fuel significantly, whereas lower yields could cause a rapid increase in fuel cost. Biomass cost sensitivity is also important, not only because of its impact, but also because the cost to acquire feedstock can vary widely between locations and throughout the year. Overall, sensitivity results suggest a greater negative sensitivity.

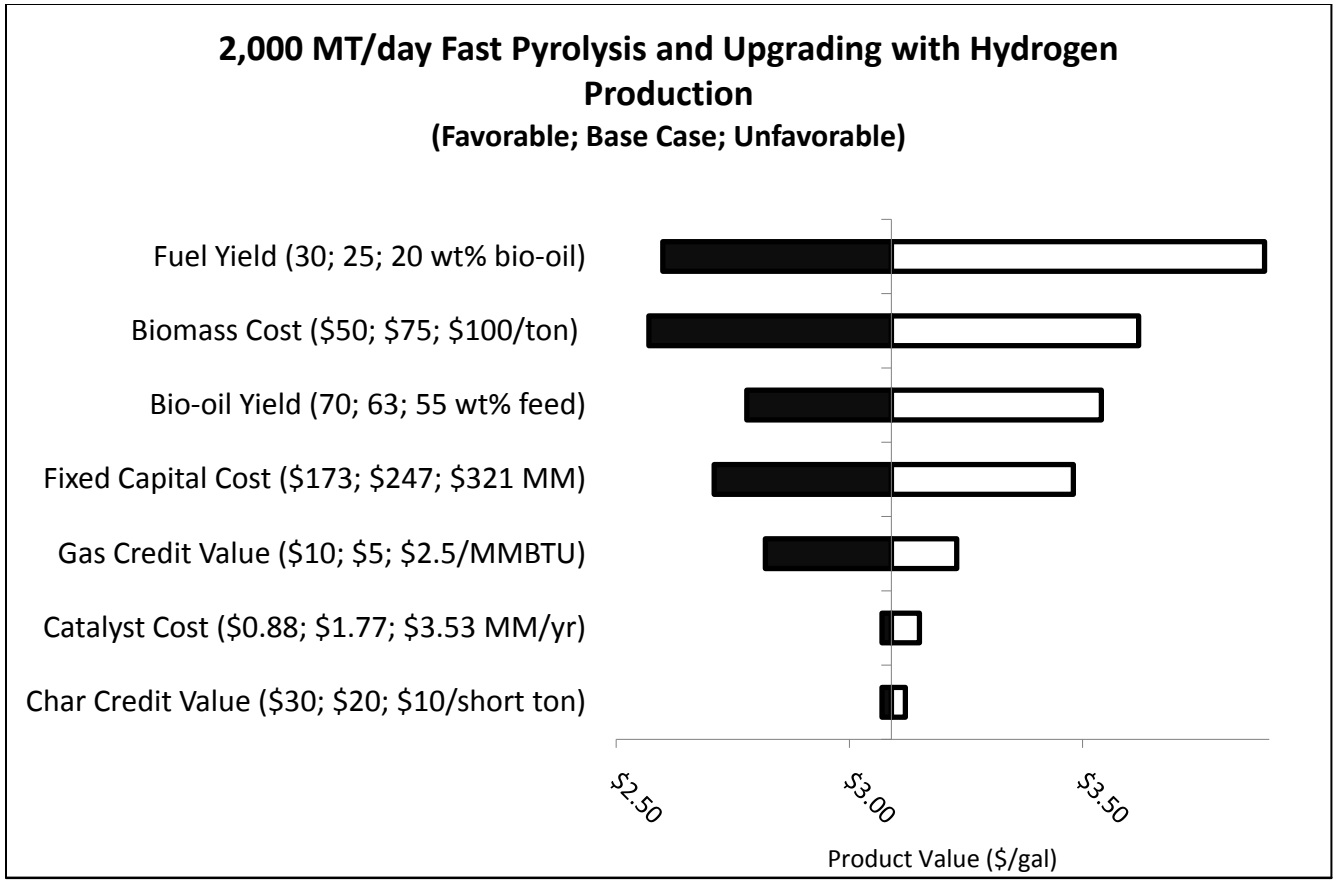


Figure 5. Sensitivity analysis for 2,000 MT/day hydrogen-production fast pyrolysis and upgrading scenario

Table 15 shows the risk analysis results for the hydrogen-production scenario. These are the estimated costs for a pioneer plant based on the current process analysis. Absent of learning effects from available commercial implementations, building a first-of-a-kind biomass fast pyrolysis and upgrading plant is expected to cost about $864 million. Fuel costs from this plant are $6.55/gal.

Table 15. Pioneer Analysis for 2,000 MT/day Hydrogen-Production Fast Pyrolysis and Upgrading Scenario

	n^{th} Plant	Optimistic	Base Case	Pessimistic
Capital Cost (millions $)	$287	$479.8	$911.6	$1,236.3
Product Value ($/GGE)	$3.09	$4.32	$6.55	$8.23

Hydrogen-Purchase Scenario

The product value for a 2,000 MT/day fast pyrolysis and upgrading plant with external hydrogen production is $2.11/gal of fuel. Capital expenditures for this plant are estimated at $200 million.

The hydrogen-purchase scenario employs 2,040 kg/h of hydrogen to upgrade 60,000 kg/h of bio-oil. Feedstock-to-liquid-fuel efficiency for the hydrogen-purchase scenario is estimated at 50% and includes the hydrogen energy input (120 MJ/kg or 0.98 GGE/kg of hydrogen). Fuel production for the hydrogen-purchase scenario is 58.2 million gallons per year.

Table 16 includes capital costs for the hydrogen-purchase scenario. Compared to the hydrogen-production scenario, the hydroprocessing section has a much lower cost because it does not include equipment to reform and compress hydrogen.

Table 16. Capital Costs for 2,000 MT/day Hydrogen-Purchase Fast Pyrolysis and Upgrading Scenario

Capital Costs (millions $)	
Combustion	$45.9
Pyrolysis and Oil Recovery	$28.0
Pretreatment	$20.2
Hydroprocessing	$14.8
Storage	$1.7
Total Equipment Installed Cost	**$110.6**
Indirect Costs	$70.9
(% of TEIC + IC)	20%
Project Contingency	$32.6
Total Project Investment (TPI)	$200
Installed Equipment Cost per Annual Gallon	$1.90
Total Project Investment per Annual Gallon	$3.43
Lang Factor	5.46

Table 17 shows the operating costs for the hydrogen-purchase scenario. Feedstock costs contribute almost half the cost of fuel. Although most operating expenditures are comparable to the hydrogen-production scenario, the increase in fuel yield reduces the per-gallon cost of fuel.

Table 17. Operating Costs for 2,000 MT/day Hydrogen-Purchase Fast Pyrolysis and Upgrading Scenario

Operating Costs (cents/gal product)	
Feedstock	93.5
Hydrogen	40.7
Electricity	8.4
Solids Disposal	3.1
Catalyst	3.1
Fixed Costs	15.5
Co-Product Credits	-10.1
Capital Depreciation	14.8
Average Income Tax	11.7
Average Return on Investment (10% IRR)	31.1
Total	**211.4**
Operating Costs (millions $/yr)	
Feedstock	$54.4
Hydrogen	$23.7
Electricity	$4.9
Solids Disposal	$1.8
Catalyst	$1.8
Fixed Costs	$8.8
Co-Product Credits	-$5.9
Capital Depreciation	$8.6
Average Income Tax	$6.8
Average Return on Investment	$18.1
Total	**$123.0**

Sensitivity analysis results for the hydrogen-purchase scenario are shown in Figure 6. Biomass cost has a significant impact on the cost of fuel, and at $100/ton, the product value estimate is $2.43/gal.

Risk analysis for this scenario shows a similar trend to the hydrogen-production pioneer plant. As shown in Table 18, capital cost estimates for base case assumptions are $585 million with pioneer plant fuel costs of $3.41/gal.

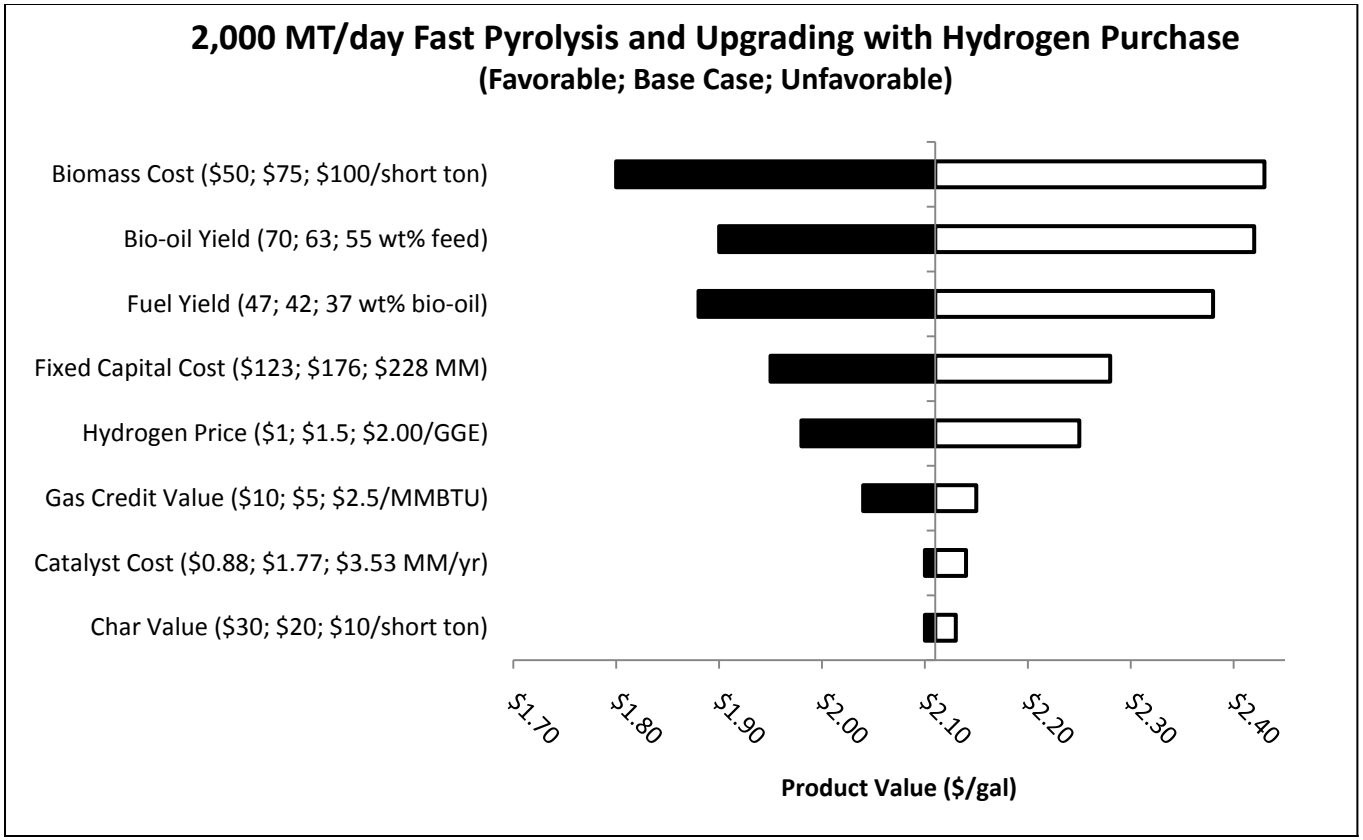


Figure 6. Sensitivity analysis for 2,000 MT/day hydrogen-purchase fast pyrolysis and upgrading scenario

Table 18. Pioneer Analysis for 2,000 MT/day Hydrogen-Purchase Fast Pyrolysis and Upgrading Scenario

	n^{th} Plant	Optimistic	Base Case	Pessimistic
Capital Cost (millions $)	$200	$307.9	$584.9	$793.2
Product Value ($/GGE)	$2.11	$2.54	$3.41	$4.07

Conclusions

This techno-economic study explores the cost of converting corn stover into naphtha-range and diesel-range stock fuel via fast pyrolysis and bio-oil upgrading. Based on the current analysis, naphtha and diesel can potentially be produced from biomass by an n^{th} plant at a competitive product value (PV) of $3.09/gal or $2.11/gal when hydrogen is procured from a remote source. For a pioneer plant, projected fuel PV increases to $6.55/gal for hydrogen-production technology and $3.41/gal for hydrogen-purchase technology.

Two scenarios are modeled in this study. One is a hydrogen-production scenario employing bio-oil reforming to generate requisite hydrogen. The other is a hydrogen-purchase scenario that relies on merchant hydrogen. Both scenarios process 2,000 metric tons per day of corn stover to generate 35 million and 58 million gallons of naphtha-range and diesel-range stock fuel, respectively. The hydrogen-production scenario sacrifices a portion of bio-oil to produce hydrogen, which results in lower yields compared to the purchase scenario. n^{th} plant capital costs are estimated at $287 million for the hydrogen-production scenario and $200 million for the hydrogen-purchase scenario. The difference is primarily due to the high cost of reforming equipment.

The source of hydrogen for the hydrogen-purchase scenario is not specified in this report. Non-fossil energy sources are preferable to achieve the highest reductions in greenhouse gas emissions associated with transportation fuel production. The advantage of this approach is that it maximizes the conversion of carbon in biomass to biofuels.

Sensitivity analysis of key process variables finds fuel conversion yields to have the most impact on the final cost of transportation fuel. Variations of 5% in the bio-oil upgrading yield result in product values of $2.60−$3.89/gal for the hydrogen-production scenario. Biomass cost and bio-oil yield are found to have significant impact on the cost of fuel for both scenarios. The capital cost sensitivity range is ±30% and has a relatively small impact, with product values ranging between $2.71 and $3.48.

Pioneer plant analysis is employed in this study to estimate the capital and operating costs of a first-of-a-kind biomass fast pyrolysis and upgrading plant. Results indicate that a pioneer hydrogen-production fast pyrolysis and upgrading plant could require an investment of $912 million and have a product value of $6.55. These high costs indicate that some aspects of this technology, notably the bio-oil upgrading process, require further development to reduce uncertainties in costs.

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Appendix A. Equipment Cost and Description

Table A-1. Equipment Costs for Fast Pyrolysis and Bio-Oil Upgrading with Hydrogen Production

Number Required	Number Spares	No. Req. Variable	Area	Equipment Name	Equipment Type	Equipment Cost	Total Equipment Cost (with spares)	Installed Cost
1	0	1	Upgrading	Liquid/Gas Fuel Separator	DVT CYLINDER	$283,200	$283,200	$855,264
1	0	1	Upgrading	Bio-oil Separation Vessel	DVT CYLINDER	$145,100	$145,100	$438,202
1	1	2	Upgrading	Reforming Bio-oil Pump	DCP CENTRIF	$47,600	$95,200	$287,504
1	1	2	Upgrading	Hydroprocessing Pump	DCP CENTRIF	$114,200	$228,400	$689,768
1	1	2	Upgrading	Reforming Water Pump	DCP CENTRIF	$48,700	$97,400	$294,148
1	0	1	Upgrading	Hydrotreater/hydrocracker	DAT REACTOR	$329,700	$329,700	$995,694
1	1	2	Upgrading	Reforming Water Heater	DHE FLOAT HEAD	$737,700	$1,475,400	$4,455,708
1	1	2	Upgrading	Reforming Preheater	DHE FLOAT HEAD	$241,100	$482,200	$1,456,244
1	1	2	Upgrading	hydrogen Cooler	DHE FLOAT HEAD	$599,300	$1,198,600	$3,619,772
1	1	2	Upgrading	Reformer Recycle Heater	DHE FLOAT HEAD	$739,500	$1,479,000	$4,466,580
1	0	1	Upgrading	Catalyst Guard Bed	C	$827,815	$827,815	$2,500,000
1	0	1	Upgrading	Reforming Steam/Water Sep	DVT CYLINDER	$180,700	$180,700	$545,714
1	0	1	Upgrading	Flue Gas Water Condenser	DVT CYLINDER	$137,700	$137,700	$415,854
1	0	1	Upgrading	hydrogen Compressor	DGC CENTRIF	$2,015,500	$2,015,500	$6,086,810
1	0	1	Upgrading	hydrogen Compressor 2	DGC CENTRIF	$3,655,600	$3,655,600	$11,039,912
1	0	1	Upgrading	Pressure Swing Adsorption	DVT CYLINDER	$76,300	$76,300	$230,426
1	0	1	Upgrading	Bio-oil Reformer	DVT JACKETED	$184,800	$184,800	$558,096
1	0	1	Upgrading	Bio-oil Prereformer	DVT JACKETED	$209,300	$209,300	$632,086
1	0	1	Upgrading	Flue Gas Blower	EFN CENTRIF	$214,700	$214,700	$648,394
1	0	1	Upgrading	Flue Gas Combustor	EFU VERTICAL	$2,800,100	$2,800,100	$8,456,302

4	0	4	Pyrolysis	Screw Feeder	ECO SCREW	$162,500	$650,000	$1,963,000
1	1	2	Pyrolysis	Condenser Water Pump	DCP CENTRIF	$669,700	$1,339,400	$4,044,988
1	1	2	Pyrolysis	Bio-oil Condenser	DHE FLOAT HEAD	$998,700	$1,997,400	$6,032,148
4	0	4	Pyrolysis	Biomass Feeding Bin	DVT CONE BTM	$41,400	$165,600	$500,112
1	0	1	Pyrolysis	Electro-Static Precipitator	EDC ELC L VOLT	$292,400	$292,400	$883,048
1	0	1	Pyrolysis	NCG/Oil Separation	DVT CYLINDER	$215,400	$215,400	$650,508
1	0	1	Pyrolysis	Pyrolysis Vapor Cyclones	EDC CYCLONE	$1,262,000	$1,262,000	$3,811,240
4	0	4	Pyrolysis	Pyrolysis Fluid Bed	DVT JACKETED	$836,000	$3,344,000	$10,098,880
1	0	1	Combustion	Char/Ash Feeding Bin	DVT CONE BTM	$165,700	$165,700	$500,414
1	0	1	Combustion	Screw Feeder	ECO SCREW	$379,600	$379,600	$1,146,392
1	0	1	Combustion	Water Softener System	C	$1,858,000	$1,858,000	$5,611,160
1	0	1	Combustion	Deaerator	C	$223,000	$223,000	$673,460
1	0	1	Combustion	Solids Combustor	DVT JACKETED	$244,300	$244,300	$737,786
1	1	2	Combustion	BFW Pump	DCP CENTRIF	$129,700	$259,400	$783,388
1	1	2	Combustion	BFW Heater	DHE FLOAT HEAD	$2,543,600	$5,087,200	$15,363,344
1	1	2	Combustion	BFW Preheater	DHE FLOAT HEAD	$3,141,200	$6,282,400	$18,972,848
1	0	1	Combustion	Combustor Cyclones	EDC CYCLONE	$1,104,700	$1,104,700	$3,336,194
1	0	1	Combustion	Combustion Gas Blower	EFN CENTRIF	$62,200	$62,200	$187,844
2	0	2	Pretreatment	Bale Transport Conveyor	C	$533,000	$1,066,000	$3,219,320
2	0	2	Pretreatment	Bale Unwrapping Conveyor	C	$200,000	$400,000	$1,208,000
1	0	1	Pretreatment	Discharge Conveyor	C	$67,000	$67,000	$202,340
1	0	1	Pretreatment	Truck Scales	C	$45,000	$45,000	$135,900
4	0	4	Pretreatment	Truck Unloading Forklift	C	$24,000	$96,000	$289,920
4	0	4	Pretreatment	Bale Moving Forklift	C	$24,000	$96,000	$289,920

1	0	1	Pretreatment	Concrete Storage Slab	C	$600,000	$600,000	$1,812,000
1	0	1	Pretreatment	Belt Press	C	$133,000	$133,000	$401,660
1	0	1	Pretreatment	Magnetic Separator	C	$19,000	$19,000	$57,380
1	0	1	Pretreatment	Biomass Chopping Screen	EVS ONE DECK	$22,500	$22,500	$67,950
4	0	4	Pretreatment	Rotary Dryer	ERD DIRECT	$681,400	$2,725,600	$8,231,312
1	0	1	Pretreatment	Biomass Grinding Screen	EVS ONE DECK	$23,000	$23,000	$69,460
1	0	1	Pretreatment	Steam Blower	EFN CENTRIF	$803,300	$803,300	$2,425,966
1	0	1	Pretreatment	Grinding Hammer Mill	ECR HAMMER MED	$302,200	$302,200	$912,644
1	0	1	Pretreatment	Chopper	ECR HAMMER MED	$302,200	$302,200	$912,644
							$0	$0
4	0	4	Storage	Fuel Storage	DVT STORAGE	$469,800	$1,879,200	$5,675,184
1	1	2	Storage	Liquid Fuel Pump	DCP CENTRIF	$24,300	$48,600	$146,772
1	0	1	Utilities	Cooling Tower	ECTWCOOLING WP	$3,005,100	$3,005,100	$9,075,402
					Totals	**$35,197,515**	**$52,683,115**	**$159,103,006**

Table A-2. Equipment Costs for Fast Pyrolysis and Bio-Oil Upgrading with Hydrogen Purchase

Number Required	Number Spares	No. Req. Variable	Area	Equipment Name	Equipment Type	Equipment Cost	Total Equipment Cost (with spares)	Installed Cost
1	1	2	Upgrading	Hydroprocessing Pump	DCP CENTRIF	$114,200	$228,400	$689,768
1	0	1	Upgrading	Hydrotreater/hydrocracker	DAT REACTOR	$329,700	$329,700	$995,694
1	0	1	Upgrading	Catalyst Guard Bed	C	$827,815	$827,815	$2,500,000
1	1	2	Upgrading	Flue Gas Water Condenser	DVT CYLINDER	$137,700	$275,400	$831,708
1	1	2	Upgrading	Flue Gas Blower	EFN CENTRIF	$214,700	$429,400	$1,296,788
1	0	1	Upgrading	Flue Gas Combustor	EFU VERTICAL	$2,800,100	$2,800,100	$8,456,302
4	0	4	Pyrolysis	Screw Feeder	ECO SCREW	$162,500	$650,000	$1,963,000
1	1	2	Pyrolysis	Condenser Water Pump	DCP CENTRIF	$669,700	$1,339,400	$4,044,988
1	1	2	Pyrolysis	Bio-oil Condenser	DHE FLOAT HEAD	$998,700	$1,997,400	$6,032,148
4	0	4	Pyrolysis	Biomass Feeding Bin	DVT CONE BTM	$41,400	$165,600	$500,112
1	0	1	Pyrolysis	Electro-Static Precipitator	EDC ELC L VOLT	$292,400	$292,400	$883,048
1	0	1	Pyrolysis	NCG/Oil Separation	DVT CYLINDER	$215,400	$215,400	$650,508
1	0	1	Pyrolysis	Pyrolysis Vapor Cyclones	EDC CYCLONE	$1,262,000	$1,262,000	$3,811,240
4	0	4	Pyrolysis	Pyrolysis Fluid Bed	DVT JACKETED	$836,000	$3,344,000	$10,098,880
1	0	1	Combustion	Char/Ash Feeding Bin	DVT CONE BTM	$165,700	$165,700	$500,414
1	0	1	Combustion	Screw Feeder	ECO SCREW	$379,600	$379,600	$1,146,392
1	0	1	Combustion	Water Softener System	C	$1,858,000	$1,858,000	$5,611,160
1	0	1	Combustion	Deaerator	C	$223,000	$223,000	$673,460
1	0	1	Combustion	Solids Combustor	DVT JACKETED	$244,300	$244,300	$737,786
1	1	2	Combustion	BFW Pump	DCP CENTRIF	$129,700	$259,400	$783,388

1	1	2	Combustion	BFW Heater	DHE FLOAT HEAD	$2,543,600	$5,087,200	$15,363,344
1	1	2	Combustion	BFW Preheater	DHE FLOAT HEAD	$3,141,200	$6,282,400	$18,972,848
1	0	1	Combustion	Combustor Cyclones	EDC CYCLONE	$653,300	$653,300	$1,972,966
1	0	1	Combustion	Combustion Gas Blower	EFN CENTRIF	$62,200	$62,200	$187,844
2	0	2	Pretreatment	Bale Transport Conveyor	C	$533,000	$1,066,000	$3,219,320
2	0	2	Pretreatment	Bale Unwrapping Conveyor	C	$200,000	$400,000	$1,208,000
1	0	1	Pretreatment	Discharge Conveyor	C	$67,000	$67,000	$202,340
1	0	1	Pretreatment	Truck Scales	C	$45,000	$45,000	$135,900
4	0	4	Pretreatment	Truck Unloading Forklift	C	$24,000	$96,000	$289,920
4	0	4	Pretreatment	Bale Moving Forklift	C	$24,000	$96,000	$289,920
1	0	1	Pretreatment	Concrete Storage Slab	C	$600,000	$600,000	$1,812,000
1	0	1	Pretreatment	Belt Press	C	$133,000	$133,000	$401,660
1	0	1	Pretreatment	Magnetic Separator	C	$19,000	$19,000	$57,380
1	0	1	Pretreatment	Biomass Chopping Screen	EVS ONE DECK	$22,500	$22,500	$67,950
4	0	4	Pretreatment	Rotary Dryer	ERD DIRECT	$681,400	$2,725,600	$8,231,312
1	0	1	Pretreatment	Biomass Grinding Screen	EVS ONE DECK	$23,000	$23,000	$69,460
1	0	1	Pretreatment	Steam Blower	EFN CENTRIF	$803,300	$803,300	$2,425,966
1	0	1	Pretreatment	Grinding Hammer Mill	ECR HAMMER MED	$302,200	$302,200	$912,644
1	0	1	Pretreatment	Chopper	ECR HAMMER MED	$302,200	$302,200	$912,644
1	0	1	Storage	Fuel Storage	DVT STORAGE	$497,800	$497,800	$1,503,356
1	1	2	Storage	Liquid Fuel Pump	DCP CENTRIF	$24,300	$48,600	$146,772
					Totals	**$25,609,715**	**$39,624,415**	**$119,665,732**

Appendix B. Aspen Plus Software Mass Flow and Balance Results and Process Flow Diagrams

Table B-1. Aspen Plus Software Section Mass and Mole Balances

	In	Out	% Change			In	Out	% Change
Plant					**Recovery**			
Carbon	0.91	0.91	100.03%		Carbon	1.12	1.12	100.00%
Hydrogen	2.54	2.54	100.05%		Hydrogen	1.51	1.51	100.00%
Oxygen	1.92	1.92	100.24%		Oxygen	1.52	1.52	100.00%
Nitrogen	2.44	2.44	99.92%		Nitrogen	1.50	1.50	100.00%
Ash	1.39	1.37	98.33%		Ash	0.21	0.21	100.00%
Mass	287,148	287,252	100.04%		Mass	218,126	218,126	100.00%
Drying					**Storage**			
Carbon	0.91	0.91	100.00%		Carbon	0.67	0.67	100.00%
Hydrogen	2.03	2.03	100.00%		Hydrogen	1.20	1.20	100.00%
Oxygen	1.02	1.02	100.00%		Oxygen	0.45	0.45	100.00%
Nitrogen	0.01	0.01	100.00%		Nitrogen	0.00	0.00	100.00%
Ash	1.39	1.39	100.00%		Ash	1.00	1.00	100.00%
Mass	110,905	110,905	100.00%		Mass	62,957	62,957	100.00%
Pyrolysis					**Combustion**			
Carbon	1.29	1.29	100.38%		Carbon	0.73	0.73	100.00%
Hydrogen	1.61	1.61	100.05%		Hydrogen	0.47	0.47	100.00%
Oxygen	1.53	1.53	100.07%		Oxygen	1.30	1.30	100.00%
Nitrogen	1.51	1.51	99.88%		Nitrogen	2.26	2.26	100.00%
Ash	1.42	1.39	98.36%		Ash	1.19	1.19	100.00%
Mass	230,902	231,006	100.04%		Mass	226,395	226,395	100.00%
Cleanup					**Hydrotreating**			
Carbon	1.29	1.29	100.00%		Carbon	0.56	0.56	99.18%
Hydrogen	1.61	1.61	100.00%		Hydrogen	1.65	1.65	100.03%
Oxygen	1.53	1.53	100.00%		Oxygen	1.14	1.14	100.30%
Nitrogen	1.51	1.51	100.00%		Nitrogen	1.67	1.67	100.00%
Ash	1.39	1.39	100.00%		Ash	0.20	0.20	100.00%
Mass	231,005	231,005	100.00%		Mass	180,854	180,854	100.00%

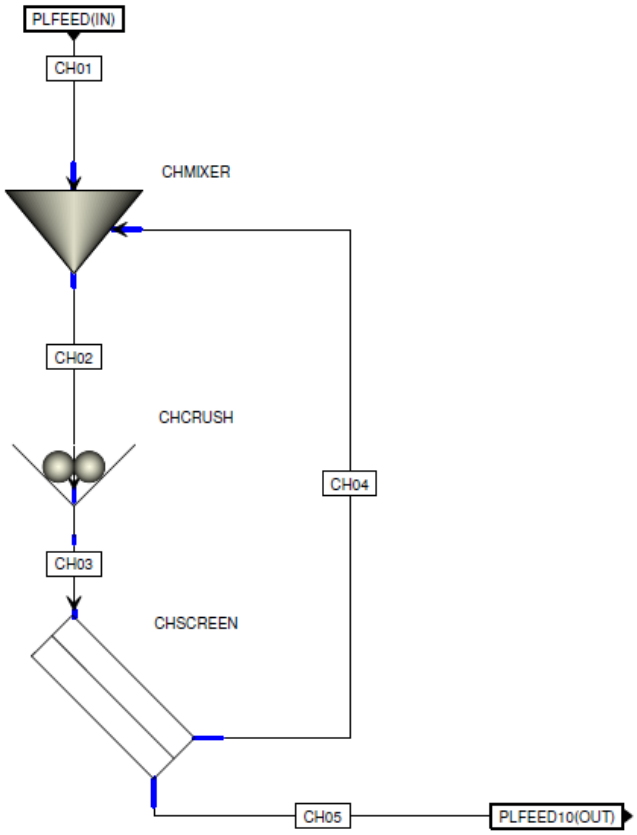


Figure B-1. Biomass chopping process flow diagram (PFD)

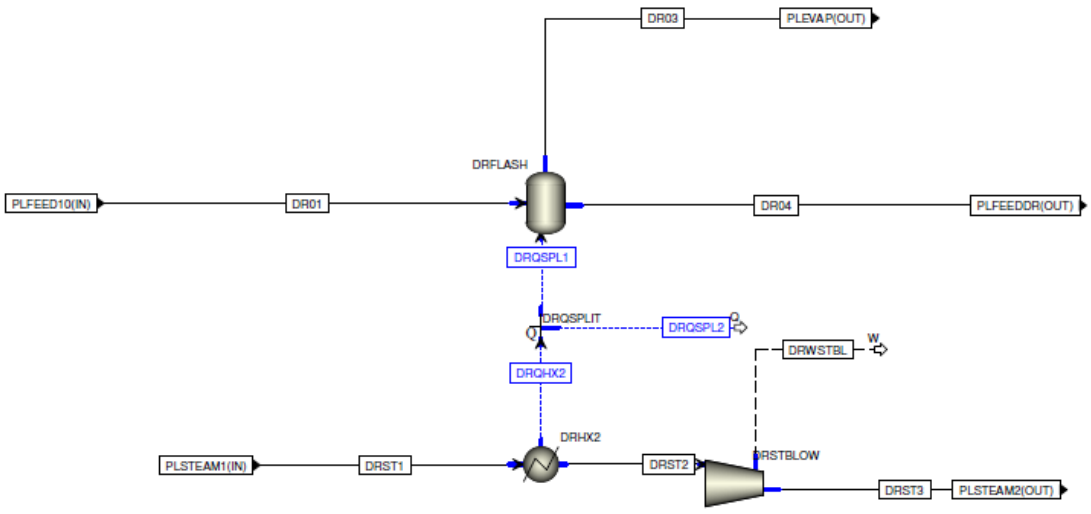


Figure B-2. Biomass drying PFD

Table B-2. Stream Table for Biomass Drying

	DR01	DR03	DR04	DRST1	DRST2	DRST3
Temperature °C	25	100	100	196.7	120	132.8
Pressure bar	1.013	1.013	1.013	14.479	1.983	2.181
Mass Flow kg/hr						
N2	0	0	0	0	0	0
O2	0	0	0	0	0	0
H2	0	0	0	0	0	0
CO	0	0	0	0	0	0
CO2	0	0	0	0	0	0
WATER	27778	21633	6145	0	0	0
NH3	0	0	0	0	0	0
CH4	0	0	0	0	0	0
C2H4	0	0	0	0	0	0
C3H6	0	0	0	0	0	0
AR	0	0	0	0	0	0
C2H4O2	0	0	0	0	0	0
C3H6O2	0	0	0	0	0	0
C7H8O2	0	0	0	0	0	0
C8H10O	0	0	0	0	0	0
CH2O2	0	0	0	0	0	0
C10H12O2	0	0	0	0	0	0
C6H6O	0	0	0	0	0	0
C7H8	0	0	0	0	0	0
C5H4O2	0	0	0	0	0	0
C6H6	0	0	0	0	0	0
NO	0	0	0	0	0	0
NO2	0	0	0	0	0	0
SULF	0	0	0	0	0	0
METHANOL	0	0	0	0	0	0
C8H18	0	0	0	0	0	0
C10H22	0	0	0	0	0	0
CL2	0	0	0	0	0	0
STEAM	0	0	0	767236	767236	767236
SO2	0	0	0	0	0	0
CHAR	0	0	0	0	0	0
STOVER	83334	0	83334	0	0	0
ASH	0	0	0	0	0	0

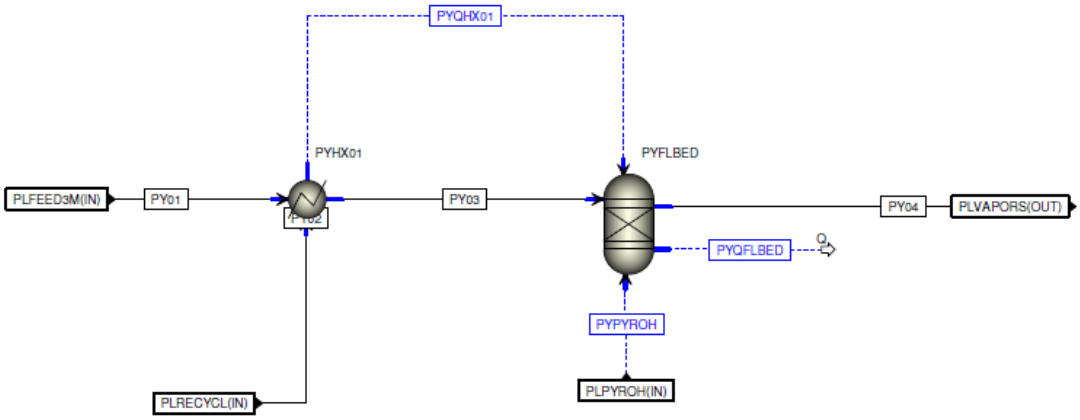


Figure B-3. Biomass pyrolysis PFD

Table B-3. Stream Table for Biomass Pyrolysis

	PY01	PY02	PY03	PY04
Temperature °C	100	56.7	250	480
Pressure bar	1.013	1.098	1.013	1.013
Mass Flow kg/hr				
N2	0	75485	75485	75840
O2	0	0	0	0
H2	0	0	0	3
CO	0	328	328	4953
CO2	0	57392	57392	67651
WATER	6145	7564	13709	16849
NH3	0	0	0	0
CH4	0	53	53	806
C2H4	0	1	1	19
C3H6	0	1	1	22
AR	0	0	0	0
C2H4O2	0	46	46	2818
C3H6O2	0	79	79	13723
C7H8O2	0	0	0	433
C8H10O	0	1	1	2836
CH2O2	0	362	362	12882
C10H12O2	0	3	3	13698
C6H6O	0	0	0	36
C7H8	0	22	22	363
C5H4O2	0	41	41	3032
C6H6	0	0	0	6
NO	0	0	0	0
NO2	0	0	0	0
SULF	0	0	0	0
METHANOL	0	0	0	0
C8H18	0	0	0	0
C10H22	0	0	0	0
CL2	0	0	0	0
STEAM	0	0	0	0
SO2	0	0	0	0
CHAR	0	288	288	15121
STOVER	83334	0	83334	0
ASH	0	0	0	55

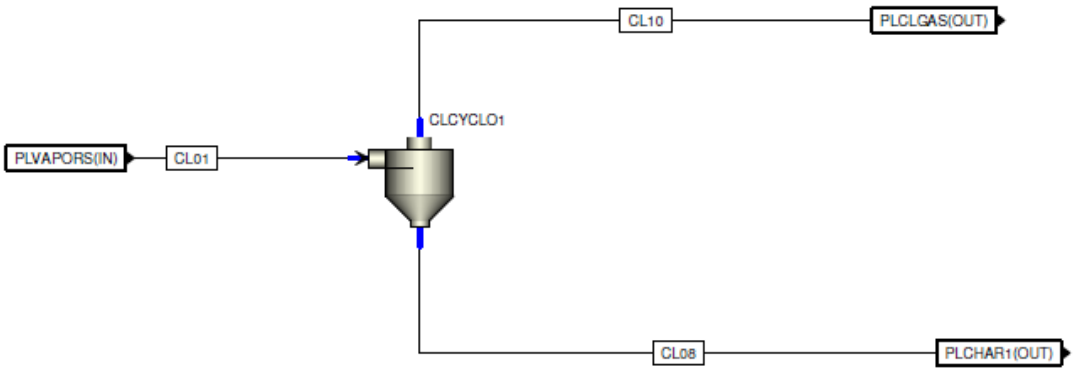


Figure B-4. Pyrolysis gas cleaning PFD

Table B-4. Stream Table for Pyrolysis Gas Cleaning

	CL01	CL08	CL10
Temperature °C	480	480	480
Pressure bar	1.013	0.997	0.997
Mass Flow kg/hr			
N2	75840	0	75840
O2	0	0	0
H2	3	0	3
CO	4953	0	4953
CO2	67651	0	67651
WATER	16849	0	16849
NH3	0	0	0
CH4	806	0	806
C2H4	19	0	19
C3H6	22	0	22
AR	0	0	0
C2H4O2	2818	0	2818
C3H6O2	13723	0	13723
C7H8O2	433	0	433
C8H10O	2836	0	2836
CH2O2	12882	0	12882
C10H12O2	13698	0	13698
C6H6O	36	0	36
C7H8	363	0	363
C5H4O2	3032	0	3032
C6H6	6	0	6
NO	0	0	0
NO2	0	0	0
SULF	0	0	0
METHANOL	0	0	0
C8H18	0	0	0
C10H22	0	0	0
CL2	0	0	0
STEAM	0	0	0
SO2	0	0	0
CHAR	15121	12853	2268
STOVER	0	0	0
ASH	55	47	8

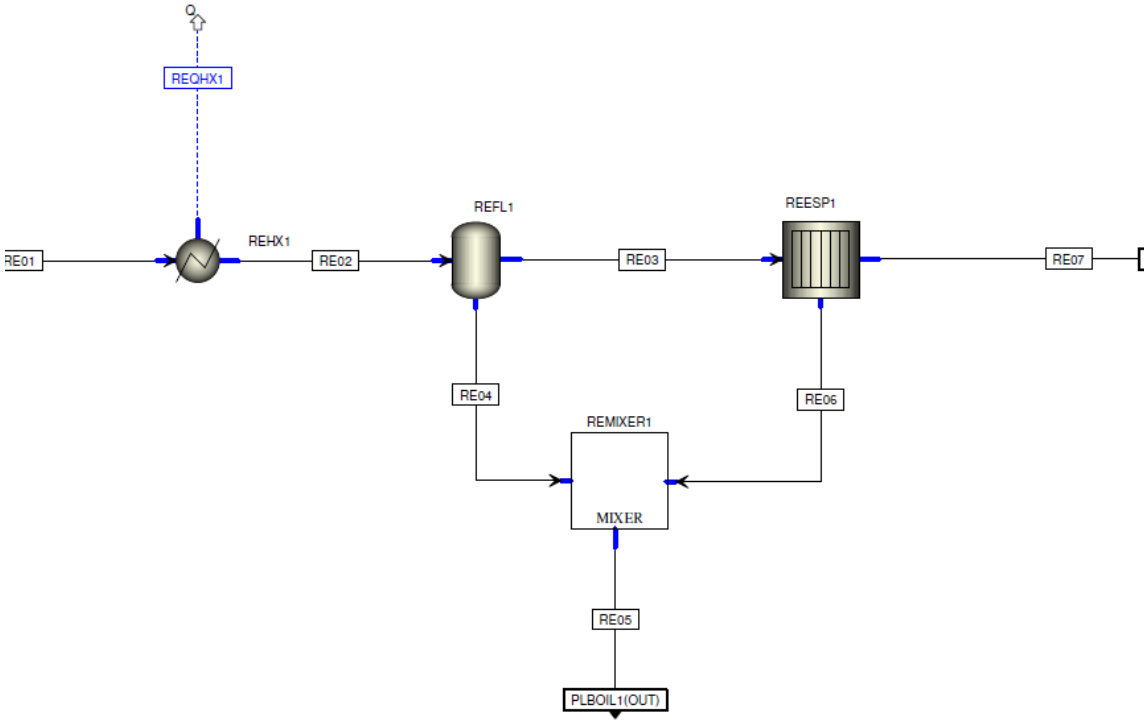


Figure B-5. Bio-oil recovery PFD

Table B-5. Stream Table for Bio-Oil Recovery

	RE01	RE02	RE03	RE04	RE05	RE06	RE07
Temperature °C	480	150	50	50	50	50	50
Pressure bar	0.997	0.997	1.014	1.014	1.013	1.013	1.013
Mass Flow kg/hr							
N2	75840	75840	75840	0	0	0	75840
O2	0	0	0	0	0	0	0
H2	3	3	3	0	0	0	3
CO	4953	4953	4953	0	0	0	4953
CO2	67651	67651	67606	46	46	0	67606
WATER	16849	16849	6159	10690	10690	0	6159
NH3	0	0	0	0	0	0	0
CH4	806	806	806	0	0	0	806
C2H4	19	19	19	0	0	0	19
C3H6	22	22	21	0	0	0	21
AR	0	0	0	0	0	0	0
C2H4O2	2818	2818	689	2129	2129	0	689
C3H6O2	13723	13723	1197	12526	12526	0	1197
C7H8O2	433	433	0	432	432	0	0
C8H10O	2836	2836	8	2828	2828	0	8
CH2O2	12882	12882	5472	7410	7410	0	5472
C10H12O2	13698	13698	44	13653	13653	0	44
C6H6O	36	36	1	35	35	0	1
C7H8	363	363	338	25	25	0	338
C5H4O2	3032	3032	615	2417	2417	0	615
C6H6	6	6	6	0	0	0	6
NO	0	0	0	0	0	0	0
NO2	0	0	0	0	0	0	0
SULF	0	0	0	0	0	0	0
METHANOL	0	0	0	0	0	0	0
C8H18	0	0	0	0	0	0	0
C10H22	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
CHAR	2268	2268	454	1815	2210	396	58
STOVER	0	0	0	0	0	0	0
ASH	8	8	2	7	8	1	0

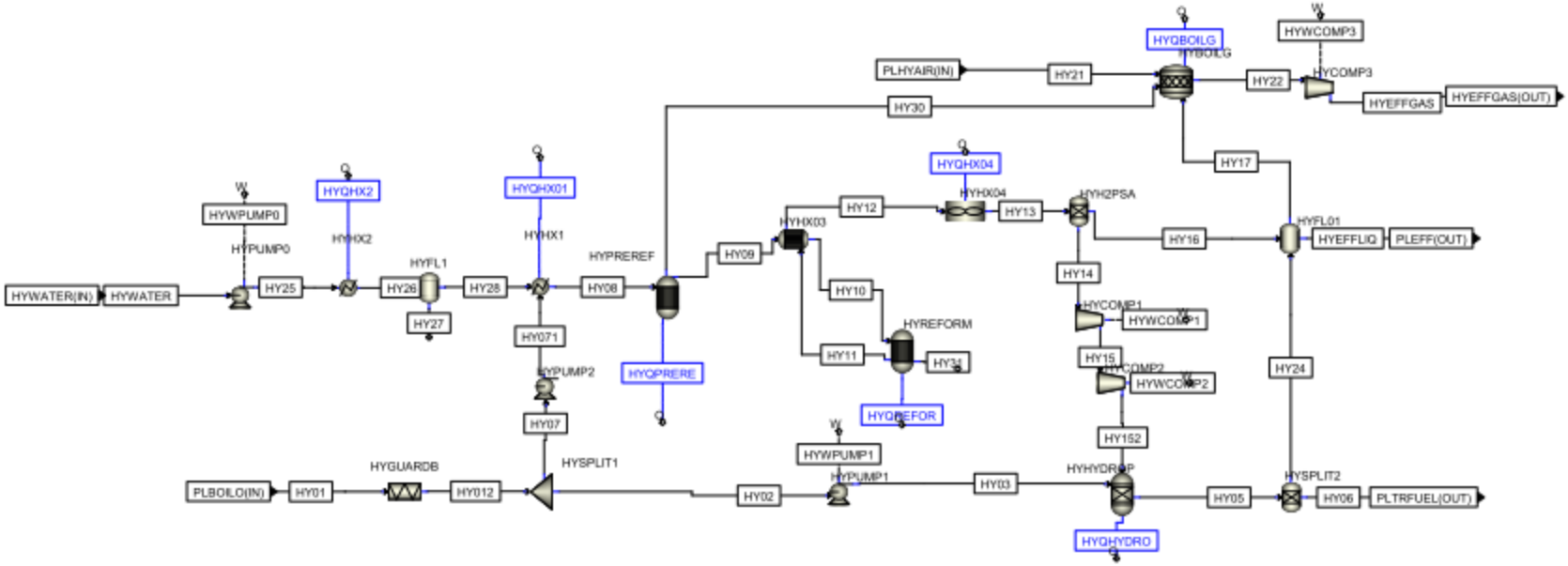


Figure B-6. Bio-oil reforming and hydroprocessing PFD

Table B-6. Stream Table for Bio-Oil Reforming and Hydroprocessing

	HY01	HY02	HY03	HY05	HY06	HY07	HY08
Temperature °C	50	50	56.2	450	140.7	50	260
Pressure bar	1.014	1.014	68.948	68.948	1.014	1.014	13.79
Mass Flow kg/hr							
N2	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0
H2	0	0	0	1	0	0	0
CO	0	0	0	0	0	0	0
CO2	46	28	28	8500	0	18	18
WATER	10690	6576	6576	10501	0	4115	24204
NH3	0	0	0	0	0	0	0
CH4	0	0	0	1070	0	0	0
C2H4	0	0	0	0	0	0	0
C3H6	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0
C2H4O2	2129	1310	1310	0	0	820	820
C3H6O2	12526	7705	7705	0	0	4821	4821
C7H8O2	432	266	266	0	0	166	166
C8H10O	2828	1739	1739	0	0	1088	1088
CH2O2	7410	4558	4558	0	0	2852	2852
C10H12O2	13653	8398	8398	0	0	5255	5255
C6H6O	35	22	22	0	0	14	14
C7H8	25	15	15	0	0	10	10
C5H4O2	2417	1487	1487	0	0	930	930
C6H6	0	0	0	0	0	0	0
METHANOL	0	0	0	0	0	0	0
C8H18	0	0	0	6691	6691	0	0
C10H22	0	0	0	6691	6691	0	0
CL2	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
CHAR	2210	1360	1360	1360	0	851	851
STOVER	0	0	0	0	0	0	0
ASH	8	5	5	5	0	3	3

(continued)	HY09	HY10	HY11	HY012	HY12	HY13	HY14
Temperature °C	400	696.1	700	50	410	40	38.8
Pressure bar	13.79	13.79	13.79	1.014	13.79	13.79	9.632
Mass Flow kg/hr							
N2	0	0	0	0	0	0	0
O2	0	0	0	0	0	0	0
H2	0	0	1350	0	1350	1350	1350
CO	0	0	5029	0	5029	5029	0
CO2	18	18	15763	46	15763	15763	0
WATER	24204	24204	14862	10690	14862	14862	0
NH3	0	0	0	0	0	0	0
CH4	0	0	3175	0	3175	3175	0
C2H4	0	0	0	0	0	0	0
C3H6	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0
C2H4O2	820	820	0	2129	0	0	0
C3H6O2	4821	4821	0	12526	0	0	0
C7H8O2	166	166	0	432	0	0	0
C8H10O	1088	1088	0	2828	0	0	0
CH2O2	2852	2852	0	7410	0	0	0
C10H12O2	5255	5255	0	13653	0	0	0
C6H6O	14	14	0	35	0	0	0
C7H8	10	10	0	25	0	0	0
C5H4O2	930	930	0	2417	0	0	0
C6H6	0	0	0	0	0	0	0
METHANOL	0	0	0	0	0	0	0
C8H18	0	0	0	0	0	0	0
C10H22	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
CHAR	0	0	0	2210	0	0	0
STOVER	0	0	0	0	0	0	0
ASH	0	0	0	8	0	0	0

(continued)	HY15	HY16	HY17	HY21	HY22	HY24	HY25
Temperature °C	174.5	38.8	26.4	25	50.7	25	25.6
Pressure bar	25	9.632	1.013	1.014	1.014	1.014	13.79
Mass Flow kg/hr							
N2	0	0	0	78064	78065	0	0
O2	0	0	0	23703	3951	0	0
H2	1350	0	1	0	0	1	0
CO	0	5029	5029	0	0	0	0
CO2	0	15763	24244	0	43750	8500	0
WATER	0	14862	631	0	10137	10501	20089
NH3	0	0	0	0	0	0	0
CH4	0	3175	4230	0	0	1070	0
C2H4	0	0	0	0	0	0	0
C3H6	0	0	0	0	0	0	0
AR	0	0	0	0	0	0	0
C2H4O2	0	0	0	0	0	0	0
C3H6O2	0	0	0	0	0	0	0
C7H8O2	0	0	0	0	0	0	0
C8H10O	0	0	0	0	0	0	0
CH2O2	0	0	0	0	0	0	0
C10H12O2	0	0	0	0	0	0	0
C6H6O	0	0	0	0	0	0	0
C7H8	0	0	0	0	0	0	0
C5H4O2	0	0	0	0	0	0	0
C6H6	0	0	0	0	0	0	0
METHANOL	0	0	0	0	0	0	0
C8H18	0	0	0	0	0	0	0
C10H22	0	0	0	0	0	0	0
CL2	0	0	0	0	0	0	0
STEAM	0	0	0	0	0	0	0
SO2	0	0	0	0	0	0	0
CHAR	0	0	0	0	851	1360	0
STOVER	0	0	0	0	0	0	0
ASH	0	0	0	0	3	5	0

(continued)	HY26	HY27	HY28	HY30	HY31	HY071	HY152
Temperature °C	371.1		370	400		51.3	382
Pressure bar	13.79		13.79	13.79		13.79	69
Mass Flow kg/hr							
N2	0		0	0		0	0
O2	0		0	0		0	0
H2	0		0	0		0	1350
CO	0		0	0		0	0
CO2	0		0	0		18	0
WATER	20089		20089	0		4115	0
NH3	0		0	0		0	0
CH4	0		0	0		0	0
C2H4	0		0	0		0	0
C3H6	0		0	0		0	0
AR	0		0	0		0	0
C2H4O2	0		0	0		820	0
C3H6O2	0		0	0		4821	0
C7H8O2	0		0	0		166	0
C8H10O	0		0	0		1088	0
CH2O2	0		0	0		2852	0
C10H12O2	0		0	0		5255	0
C6H6O	0		0	0		14	0
C7H8	0		0	0		10	0
C5H4O2	0		0	0		930	0
C6H6	0		0	0		0	0
METHANOL	0		0	0		0	0
C8H18	0		0	0		0	0
C10H22	0		0	0		0	0
CL2	0		0	0		0	0
STEAM	0		0	0		0	0
SO2	0		0	0		0	0
CHAR	0		0	851		851	0
STOVER	0		0	0		0	0
ASH	0		0	3		3	0

(continued)	HYEFFGAS	HYEFFLIQ	HYWATER
Temperature °C	62	26.4	25
Pressure bar	1.115	1.013	1.013
Mass Flow kg/hr			
N2	78065	0	0
O2	3951	0	0
H2	0	0	0
CO	0	0	0
CO2	43750	19	0
WATER	10137	24732	20089
NH3	0	0	0
CH4	0	15	0
C2H4	0	0	0
C3H6	0	0	0
AR	0	0	0
C2H4O2	0	0	0
C3H6O2	0	0	0
C7H8O2	0	0	0
C8H10O	0	0	0
CH2O2	0	0	0
C10H12O2	0	0	0
C6H6O	0	0	0
C7H8	0	0	0
C5H4O2	0	0	0
C6H6	0	0	0
METHANOL	0	0	0
C8H18	0	0	0
C10H22	0	0	0
CL2	0	0	0
STEAM	0	0	0
SO2	0	0	0
CHAR	851	1360	0
STOVER	0	0	0
ASH	3	5	0

Appendix C. Selection Process of Pyrolysis Reactor Technologies for Further Analysis under Iowa State University/ConocoPhillips Company/National Renewable Energy Laboratory Collaborative Study

Purpose

The purpose of this section is to describe the selection process employed to down select pyrolysis technologies and bio-oil upgrading to transportation fuels for further research and development.

Summary

Six different pyrolysis technologies were initially chosen to compare based on selected criteria. These technologies, based on the reactor design, were bubbling bed, circulating (transported) bed, auger, free fall (entrained flow), ablative, and catalytic pyrolysis (fluid bed). Hydrothermal processing was added during the down-selection process based on its similar primary liquid product. Two bio-oil upgrading technologies were considered: hydrogenation (hydrocracking) and gasification via the Fischer-Tropsch process.

Bubbling-bed pyrolysis was chosen as the base case because of the availability of reliable and established data. Positive and negative ratings were given to other technologies based on how they compared to the base case on various criteria. These ratings allowed for the selection of technologies for focus in the current study. **Bubbling-bed, auger, and free-fall** pyrolysis reactors were chosen for further research. Further inquiries into hydrothermal processing and hydrogenation will be done as well.

Background

A process design matrix was developed to aid in the collection of data for various technologies. These technologies were grouped into biochemical, gasification, pyrolysis, upgrading, and supporting technologies. The selected criteria included capital expenditure, operating costs, plant efficiency, carbon efficiency, capacity factor, plant size (typical), complexity of process, level of technology development, and energy content, among others. Based on these criteria, circulating (transported) bed, auger, free fall (entrained flow), ablative, and catalytic pyrolysis (fluid bed) were rated and compared to bubbling bed pyrolysis.

This down selection process was an informal process, based on the knowledge and experience of the team. No specific references were consulted in this process.

Selection Process

Bubbling-bed pyrolysis was chosen as the base-case pyrolysis technology because of the availability of reliable and established information. Following is a short description of how each of the considered alternative technologies compares to the base case in terms of the process design criteria.

Circulating (Transported) Bed

This technology requires a higher capital cost due to the need for additional reactor vessels and related equipment. Operating costs are also expected to be higher because of increased fluidizing gas requirements and the recirculation of inert bed material. Based on available knowledge, circulating bed has a lower plant and carbon efficiency. The capacity factor rating, defined as the availability of the system, is lower than the base case due to additional maintenance requirements associated with the use of additional equipment. Circulating-bed reactors, similar to bubbling-bed reactors, are better suited for larger sizes and are therefore not attractive economically for a distributed processing scenario. The complexity of this process is higher than for the base case. In terms of level of technology development, this configuration had an identical rating to the bubbling-bed technology.

Auger

The auger reactor was given an identical capital expenditure rating to the base case based on the assumption that increases in capital requirements for the reactor could be offset by savings in auxiliary equipment. Augers are expected to reduce operating costs. Plant efficiency was given an identical rating, while the carbon efficiency is expected to suffer under this configuration due to lower yields. The use of mechanical moving parts reduced the capacity factor rating. An auger reactor's typical size is well suited for a distributed processing scenario where smaller is better. Process steps might be eliminated, reducing the complexity of the process. The level of technology development is considered low for this technology.

Free Fall (Entrained Flow)

A free-fall reactor is the only option with a lower capital expense requirement. This reactor is also expected to have lower operating costs. Its plant efficiency is lower than the bubbling bed's efficiency, and the same applies to the carbon efficiency due to lower yields. This is the only option with a better capacity factor than the base case. The typical plant size is comparable to the bubbling bed. The complexity is expected to be lower, although the level of development is low.

Ablative

Capital expenditure for ablative reactors is expected to be comparable to the bubbling-bed technology. Operating costs are rated lower than the base case. Plant efficiency is similar to the bubbling bed. The carbon efficiency is lower for this technology, as is the capacity factor due to the mechanical requirements. Ablative reactors are better suited to small-scale operations. The complexity and level of technology development both received a negative rating.

Catalytic Pyrolysis

The catalytic pyrolysis analysis was based on the assumption that it employed a fluid-bed reactor design. There has not been much research done in this area, but like catalytic gasification, catalytic pyrolysis suffers from sulfur and chlorine poisoning of the catalysts. This option received negative ratings for all criteria with the exception of plant size, for which it is expected to have the same typical size as the base case.

Other Technologies

Hydrothermal processing is a different process than pyrolysis. This process involves pressurizing biomass in an aqueous solution using a batch process that requires heat and significant amounts of water. The hydrocarbon portion of the product contains less oxygen than typical bio-oil and is closer in composition to fossil-based oil. Because it is currently at a very early development stage, not enough techno-economic information is available, although up-to-date commercialization efforts have not been promising. This process yields a superior product, in terms of energy content, compared to pyrolysis, and this is the main reason for its inclusion in this study.

Hydrogenation/hydrocracking was selected for further study as well. This upgrading technology would be used at a large-scale facility. The level of technology development for this process is considered to be low at this time.

Action

The next step in this collaboration is to prepare Aspen software models for the selected technologies: bubbling-bed, auger, and free-fall pyrolysis. Hydrogenation/hydrocracking will also be evaluated. For this purpose, a work plan is being developed. In January of 2008, the plan will be presented regarding the Aspen software models' level of detail and assumptions.

Appendix D. Assumptions for Pyrolysis Technologies Techno-Economic Studies

Plant Size, Location, and Construction

- Various plant sizes will be regarded as economically feasible plant size, specifically plant capacities of 5, 50, and 550 tons/day (dry feedstock)
- Scenarios will aggregate plants to a total capacity of 2,000 tons/day
- Biomass collection area for 2,000 tons/day has a 50 mile radius, and transportation costs are 23% of feedstock costs with linear scaling for smaller-sized areas
- The plant produces pyrolysis gas, bio-oil, and charcoal
- The plant is considered to be located in the middle of corn farmland
    - 25% of the land will be tied up in infrastructure (roads and buildings), and
    - 75% of the farm land plants corn
- The plant will be designed based on the state of the technology and would be the n^{th} plant of its kind
- The online time would be 350 days/year (equivalent capacity factor of 96%)
- Construction time of less than 24 months is considered based on judgment
    - Startup period would be 25% of the construction time (6 months)
        - During this period, an average of 50% production will be achieved with expenditure of about 75% of variable expenses and 100% of fixed expenses

Feedstock

- Corn stover (comprised of stalks, leaves, cobs and husks) is considered as feedstock
    - The feedstock will be delivered to the feed handling area of the plant
    - Moisture content in the feedstock is 15% (wet basis)
    - Variation of feed compositions will be incorporated in the model
    - The feedstock transportation and management protocol are not considered
    - Feed cost is assumed to be $50/dry MT with credit for reduced transportation cost

Material and Energy Balance

Material Balance

- Biomass will be modeled using ultimate and proximate analyses data and the Aspen software coal model
- Biomass handling will be specified to less than 5-mm grinding size and 7% moisture content, also depending on specific process requirements

- No biomass material is lost during washing
- Various bio-oil compositions will be considered; bio-oil components of interest are formic acid, hydroxyacetaldehyde, acetic acid, diacetyl, glyoxal, acetol, levoglucosan, cellobiosan, water, and pyrolytic lignin
- Where necessary, nitrogen will be employed as a fluidizing agent and pyrolysis gas recirculation will also be considered
- Carbon efficiency can be calculated based on carbohydrate carbon content, as follows:

$$Carbon\ efficiency\ (\%) = \frac{Carbon\ in\ Bio-oil}{Carbon\ in\ Biomass\ Carbohydrate} \times 100$$

Energy Balance

- An Aspen software yield reactor model will be employed to calculate reaction energy balance
- Combustor energy losses will be factored into the model
- The energy value of the products will be reported to measure an overall energy balance.

Equipment Design, Material of Construction and Costing

Equipment Design

- The reactors will be modeled using experimentally determined conversions of specific reactions (kinetic expressions will be not used because of the level of their development)
- If the size of any equipment is known to change linearly with the inlet flow, that information can be used for equipment scaling (a characteristic of the size might be the heat duty for a heat exchanger if the log-mean temperature difference is known not to change)
- For some equipment, nothing can be easily related to the size, in which case the unit will be resized with each process change (for example, heat exchangers with varying temperature profiles; in this case, the heat exchanger area will be calculated each time the model will be run and the cost will be scaled using the ratio of the new and original areas)

Material of Construction

- Most construction material will consist of stainless steel

Costing

- Equipment costing data and installation factors will be collected from direct quotations, published data, and Aspen Icarus software evaluation, with preference given in the order shown here
- If process changes are made and the equipment size changes, the equipment will be re-costed following the exponential scaling expression:

$$New\ Cost = Original\ \text{Cost}\left(\frac{\text{New size}*}{\text{Original size}*}\right)^{\exp}$$

*or characteristic linearly related to the size

- The purchased equipment cost obtained in a particular year will be indexed to the year of interest (2012) using the Chemical Engineering Index
    - The existing value of the index will be regressed to extrapolate to the future year (2012)

Chemical Costing

- Costs for acids and other chemicals, if considered, will be obtained from quotation
    - The cost of the chemicals will also be indexed following the Industrial Inorganic Chemical Index (from SRI) to estimate the cost of the chemicals in the future year of interest (2012)

Operating Cost

- Working capital is assumed to be 5% of the total capital investment
    - It is assumed that the product will be made and shipped and that payment will be received in 30 days
- Annual maintenance materials cost will be 2% of the total installed equipment cost
- Employee salaries will be indexed to the future year of interest (2012) following the data of the Bureau of Labor Statistics
- Salaries of the yard employees will not include benefits and will be covered in the general overhead category
    - General overhead will be a factor of 60% applied to the total salaries and covers items such as safety, general engineering, general plant maintenance, payroll overhead (including benefits), plant security, janitorial and similar services, phone, light, heat, and plant communications

Wastewater Treatment Plant

- The process will be designed for zero discharge to a municipal treatment plant in a steady state mode
- Any process upset (sudden increase of solids in the wastewater) will not considered in the model
- Rain and snow run-off, equipment washing, and other non-process waters are assumed to flow to the municipal wastewater treatment system; other intermittent loads (process spills) will not be considered in the design

Greenhouse Emissions and Control

- All of the sulfur going into the combustor is converted to SO_2

Cost Analysis

- The total plant investment cost will be determined by applying overhead and contingency factors (NREL experience and literature) to installed equipment costs
- Insurance and taxes will be considered as 1.5% of the total installed equipment cost (Delta-T/NREL/published data)
    - The estimates are location-sensitive
- To determine the product value per gallon of liquid fuel, a discounted cash flow analysis will be used after knowing the major three costs areas: (i) total project investment, (ii) variable operating costs, and (iii) fixed operating cost
    - A 10% discounted cash flow rate of return will be used over a 20-year plant life
    - The plant is considered 100% equity financed
- For federal tax returns, depreciation will be determined as follows:
    - The IRS modified accelerated cost recovery system, including the general depreciation system will be followed; this allows both the 200% and 150% declining balance (DB) methods of depreciation
        - This allows the shortest recovery period and the largest deductions
    - Other property not specifically described in the publication should be depreciated using a 7-year recovery period
        - Property listed with a recovery period less than 10 years will use the 200% DB depreciation method and 20-year-recovery-period property will use the 150% DB depreciation
    - State tax will not be considered for the calculation (because the location of the plant is not specified)
- Return on investment will be calculated on a per gallon basis; income tax will be averaged over the plant life and that average will be calculated on a per gallon basis

Table D-1. Biomass Pyrolysis/Torrefaction Design Matrix

2 matrices - stover and softwood	Capital Expenditure (Total Project Investment)	Operating Costs ($ per year)	Fuel Cost ($/gal)	IRR	Economic Risk/Uncertainty (L, M, H)	Plant Efficiency (Energy in product/(Energy in + Energy bi	Carbon Efficiency (C in Products/C in Feedsto	Process Fuel Yield (liters/tonne)	Capacity Factor/On-Stream Time	Plant Size (tons feed or gallons fuel) *	Complexity of Process (No. of steps)	Level of Technology Development (L, M, H)	Water Use/Discharge (gal water/gal fuel)	Greenhouse Gas Emissions (Nox, Sox, CO, CO	Particulates/Solid Wastes (tons/year)	Air Emissions (tons/year)	Types of Waste Treatment Required	Toxic Materials Generated or Used?	Energy Content (LHV, Btu/gal)	Engine Compatibility	Use Emissions	Infrastructure	Fuel Tox	A
Pyrolysis Technologies (Bio-oil product) (ISU lead)																								
Bubbling Bed	0	0				0	0		0	0	0	0							0					
Circulating / Transport Bed	-	-				-	-		-	-	-	0							0					
Auger	0	+				0	-		-	+	+	-							0					
Free Fall (Entrained Flow)	+	+				-	-		+	0	+	-							0					
Ablative	0	+				0	-		-	+	-	-							0					
Catalytic Pyrolysis (Fluid Bed)	-	-				-	-		-	0	-	-							0					
Hydrothermal Processing (HTP)	-	-				0	-		-	0	-	0							+					
Bio-oil Upgrade to Transportation Fuels																								
Hydrogenation/Hydrocracking	+	+				+	+		+		0	+	-											
Gasification (FT process)																								
Supporting Technologies																								
Lignin depolymerization/catalysis																								
Lignin pyrolysis/hydrogenation																								
Lignin gasification																								
Torrefaction																								
	Economics					**Process**							**Environmental**						**Fuel Properties**					

Table D-2. Hydrogen-Production Scenario Discounted Cash Flow Rate of Return Worksheet

DCFROR Worksheet												
Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$17,222,804	$105,463,626	$56,247,267								
Working Capital				$26,365,906								
Loan Payment					$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales					$93,095,189	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919
By-Product Credit					$4,439,632	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509
Plant Performance					1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Annual Sales					$97,534,821	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428
Annual Manufacturing Cost												
Feedstock					$47,623,352	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688
Electricity					$4,288,930	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634
Catalyst					$1,546,125	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000
Hydrogen Cost					$20,725,229	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976
Solids Disposal Cost					$1,555,347	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539
Fixed Operating Costs					$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600
Total Product Cost					$84,722,583	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437
Annual Depreciation												
General Plant												
DDB					$50,220,774	$35,871,982	$25,622,844	$18,302,031	$13,072,880	$9,337,771	$6,669,837	
SL					$25,110,387	$20,925,323	$17,935,991	$16,014,278	$15,251,693	$15,251,693	$15,251,693	
Remaining Value					$125,551,936	$89,679,954	$64,057,110	$45,755,079	$32,682,199	$23,344,428	$16,674,591	
Actual					$50,220,774	$35,871,982	$25,622,844	$18,302,031	$15,251,693	$15,251,693	$15,251,693	
Steam Plant												
DDB					$0	$0	$0	$0	$0	$0	$0	$0
SL					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Actual					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($37,408,536)	($1,367,991)	$8,881,147	$16,201,960	$19,252,298	$19,252,298	$19,252,298	$34,503,991
Losses Forward						($37,408,536)	($38,776,526)	($29,895,379)	($13,693,419)	$0	$0	$0
Taxable Income					($37,408,536)	($38,776,526)	($29,895,379)	($13,693,419)	$5,558,879	$19,252,298	$19,252,298	$34,503,991
Income Tax					$0	$0	$0	$0	$2,167,963	$7,508,396	$7,508,396	$13,456,557
Annual Cash Income					$12,812,239	$34,503,991	$34,503,991	$34,503,991	$32,336,028	$26,995,595	$26,995,595	$21,047,435
Discount Factor		1.21	1.1	1	0.909090909	0.826446281	0.751314801	0.683013455	0.620921323	0.56447393	0.513158118	0.46650738
Annual Present Value	$215,543,631				$11,647,490	$28,515,695	$25,923,359	$23,566,690	$20,078,130	$15,238,309	$13,853,009	$9,818,784
Total Capital Investment + Interest		$20,839,592.44	$116,009,988.56	$82,613,173.67								
Net Present Worth				$0								
					If income taxable income < 0, tax = $0							
					Loan Interest subtracted from taxable income.							
					Loan payment subtracted from annual cash income							
					Interest on construction loan added to investment							
NPV of Income Tax	$58,488,521				$0	$0	$0	$0	$1,346,134	$4,238,294	$3,852,995	$6,277,583
NPV of Fuel Income	$1,028,551,771				$84,631,990	$102,584,231	$93,258,391	$84,780,356	$77,073,051	$70,066,410	$63,696,736	$57,906,124

9	10	11	12	13	14	15	16	17	18	19	20
											($26,365,906)
$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
$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919	$124,126,919
$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509	$5,919,509
1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428	$130,046,428
$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688
$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634	$4,901,634
$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000
$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976	$23,685,976
$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539	$1,777,539
$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600	$8,983,600
$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437	$95,542,437
$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
$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991	$34,503,991
$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557	$13,456,557
$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435	$21,047,435
0.424097618	0.385543289	0.350493899	0.318630818	0.28966438	0.263331254	0.239392049	0.217629136	0.197844669	0.17985879	0.163507991	0.148643628
$8,926,167	$8,114,697	$7,376,997	$6,706,361	$6,096,692	$5,542,447	$5,038,588	$4,580,535	$4,164,123	$3,785,566	$3,441,424	$3,128,567
											($3,919,124.00)
$5,706,894	$5,188,085	$4,716,441	$4,287,674	$3,897,885	$3,543,532	$3,221,393	$2,928,539	$2,662,308	$2,420,280	$2,200,255	$2,000,231
$52,641,931	$47,856,301	$43,505,728	$39,550,662	$35,955,147	$32,686,497	$29,714,998	$27,013,634	$24,557,849	$22,325,317	$20,295,743	$18,450,676

Table D-3. Hydrogen-Purchase Scenario Discounted Cash Flow Rate of Return Worksheet

DCFROR Worksheet													
Year		-2	-1	0	1	2	3	4	5	6	7	8
Fixed Capital Investment		$22,936,121	$148,313,504	$79,100,536								
Working Capital				$37,078,376								
Loan Payment					$0	$0	$0	$0	$0	$0	$0	$0
Loan Interest Payment		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Loan Principal		$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
Fuel Sales					$82,088,890	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853
By-Product Credit					$8,483,234	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978
Plant Performance					1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Annual Sales					$90,572,124	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831
Annual Manufacturing Cost												
Feedstock					$47,623,352	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688
Electricity					$5,050,411	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898
Catalyst					$1,546,125	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000
Solids Disposal Cost					$53,356	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978
Fixed Operating Costs					$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177
Total Product Cost					$65,756,420	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740
Annual Depreciation												
General Plant												
DDB					$70,625,478	$50,446,770	$36,033,407	$25,738,148	$18,384,391	$13,131,708	$9,379,792	
SL					$35,312,739	$29,427,283	$25,223,385	$22,520,880	$21,448,457	$21,448,457	$21,448,457	
Remaining Value					$176,563,695	$126,116,925	$90,083,518	$64,345,370	$45,960,979	$32,829,270	$23,449,479	
Actual					$70,625,478	$50,446,770	$36,033,407	$25,738,148	$21,448,457	$21,448,457	$21,448,457	
Steam Plant												
DDB					$0	$0	$0	$0	$0	$0	$0	$0
SL					$0	$0	$0	$0	$0	$0	$0	$0
Remaining Value					$0	$0	$0	$0	$0	$0	$0	$0
Actual					$0	$0	$0	$0	$0	$0	$0	$0
Net Revenue					($45,809,775)	($3,193,679)	$11,219,684	$21,514,943	$25,804,634	$25,804,634	$25,804,634	$47,253,091
Losses Forward						($45,809,775)	($49,003,454)	($37,783,770)	($16,268,827)	$0	$0	$0
Taxable Income					($45,809,775)	($49,003,454)	($37,783,770)	($16,268,827)	$9,535,807	$25,804,634	$25,804,634	$47,253,091
Income Tax					$0	$0	$0	$0	$3,718,965	$10,063,807	$10,063,807	$18,428,705
Annual Cash Income					$24,815,704	$47,253,091	$47,253,091	$47,253,091	$43,534,126	$37,189,284	$37,189,284	$28,824,385
Discount Factor		1.21	1.1	1	0.909090909	0.826446281	0.751314801	0.683013455	0.620921323	0.56447393	0.513158118	0.46650738
Annual Present Value	$301,565,008				$22,559,730	$39,052,141	$35,501,947	$32,274,497	$27,031,267	$20,992,381	$19,083,983	$13,446,789
Total Capital Investment + Interest		$27,752,706.12	$163,144,854.55	$116,178,911.57								
Net Present Worth				$0								
					If income taxable income < 0, tax = $0							
					Loan interest subtracted from taxable income.							
					Loan payment subtracted from annual cash income							
					Interest on construction loan added to investment							
NPV of Income Tax	$80,329,568				$0	$0	$0	$0	$2,309,184	$5,680,757	$5,164,324	$8,597,127
NPV of Fuel Income	$906,949,903				$74,626,263	$90,456,077	$82,232,797	$74,757,088	$67,960,989	$61,782,718	$56,166,107	$51,060,097

9	10	11	12	13	14	15	16	17	18	19	20
											($37,078,376)
$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
$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853	$109,451,853
$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978	$11,310,978
1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831	$120,762,831
$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688	$54,426,688
$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898	$5,771,898
$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000	$1,767,000
$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978	$60,978
$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177	$11,483,177
$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740	$73,509,740
$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
$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091
$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091	$47,253,091
$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705	$18,428,705
$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385	$28,824,385
0.424097618	0.385543289	0.350493899	0.318630818	0.28966438	0.263331254	0.239392049	0.217629136	0.197844669	0.17985879	0.163507991	0.148643628
$12,224,353	$11,113,048	$10,102,771	$9,184,338	$8,349,398	$7,590,362	$6,900,329	$6,273,026	$5,702,751	$5,184,319	$4,713,017	$4,284,561
											($5,511,464.33)
$7,815,570	$7,105,064	$6,459,149	$5,871,953	$5,338,140	$4,852,854	$4,411,686	$4,010,623	$3,646,021	$3,314,565	$3,013,241	$2,739,310
$46,418,270	$42,198,427	$38,362,207	$34,874,733	$31,704,303	$28,822,094	$26,201,903	$23,819,912	$21,654,466	$19,685,878	$17,896,253	$16,269,321

Table D-4. Hydrogen-Production Scenario Detailed Operating Cost Analysis

Variable Operating Costs											
Costing Code	Raw Material	Stream No.	kg/hr (or kW)	lb/hr (or HP)	Quoted Price ($/ton, or cents/kWh)	Year of Price Quote	2000 Cost (cents / ton)	2000 Cost ($/lb)	$/hour	MM$/yr (2007)	Cents/gal (2007)
	Feedstock		83,333	183,749	75	2007				54.43	153.86
	Electricity		13,530	29,834	0.054					5.77	16.32
Mixed with Oil Aqueous Phase	Process Water		11,793	26,004	0.029					0.06	0.17
15% of Operating Cost	Catalyst									1.77	5.00
	Subtotal									**62.03**	**175.35**
	Waste Streams										
Ref 67 in NREL Thermochemical ethanol	Solids Disposal Cost		11,340		18					1.78	5.03
	Subtotal									**1.78**	**5.03**
	By-Product Credits										
$5/MMBTU	Fuel Gas (MMBTU)		1,940,929		5					9.70	27.44
	Char ($/ton)		9,223		20					1.61	4.54
	Subtotal									**11.31**	**31.98**
	Total Variable Operating Costs									**52.49**	**148.40**
	Fixed Operating Costs										
	Plant Manager	80000	1	80,000		38					
	Plant Engineer	65000	1	59,560		31					
	Maintenance Supr	60000	1	60,000		29					
	Lab Manager	50000	1	50,000		24					
	Shift Supervisor	37000	5	169,518		18					
	Lab Technician	25000	2	45,816		12					
	Maintenance Tech	28000	8	205,254		13					
	Shift Operators	25000	20	458,156		12					
	Yard Employees	20000	32	586,440		10					
	General Manager	100000	0	0		48					
	Clerks & Secretaries	20000	3	54,979		10					
	Total Salaries		74	1,769,722						1.77	5.00
	Overhead/Maint	60%	of Labor & Supervison	1,061,833		18.40	PEP 2003 (uses 37.66)			1.06	3.00
	Maintenance	2%	of TPI	4,756,232			**avg salary (w/ benefits)**			4.76	13.45
	Insurance & Taxes	1.5%	of TPI	3,567,174			38264.26			3.57	10.08
	Costs			11,154,961						**11.15**	**31.54**
	Total Cash Cost									**63.65**	**179.93**
	Annual Capital Charge									**41.85**	**118.32**
	Annual Operating Cost									105.50	298.26

Table D-5. Hydrogen-Purchase Scenario Detailed Operating Cost Analysis

Variable Operating Costs											
Costing Code	Raw Material	Stream No.	kg/hr (or kW)	lb/hr (or HP)	Quoted Price ($/ton, or cents/kWh)	Year of Price Quote	2000 Cost (cents / ton)	2000 Cost ($/lb)	$/hour	MM$/yr (2007)	Cents/gal (2007)
	Feedstock		83,333	183,749	75	2007				54.43	93.51
	Electricity		11,490	25,335	0.054					4.90	8.42
Mixed with Oil Aqueous Phase	Fresh Water		0	0	0.029					0.00	
15% of Operating Cost	Catalyst									1.767	
	Hydrogen		2,041		1333					23.69	40.70
	Subtotal									**84.78**	**142.63**
	Waste Streams										
Ref 67 in NREL Thermochemical ethanol	Solids Disposal Cost		11,340		18					1.78	3.05
	Subtotal									**1.78**	**3.05**
	By-Product Credits										
$5/MMBTU	Fuel Gas (MMBTU)		862,635		5					4.31	7.41
	Char ($/ton)		9,223		20					1.61	2.76
	Subtotal									**5.92**	**10.17**
	Total Variable Operating Costs									**80.64**	**135.51**
	Fixed Operating Costs										
	Plant Manager	80000	1	80,000		38					
	Plant Engineer	65000	1	59,560		31					
	Maintenance Supr	60000	1	60,000		29					
	Lab Manager	50000	1	50,000		24					
	Shift Supervisor	37000	5	169,518		18					
	Lab Technician	25000	2	45,816		12					
	Maintenance Tech	28000	8	205,254		13					
	Shift Operators	25000	20	458,156		12					
	Yard Employees	20000	32	586,440		10					
	General Manager	100000	0	0		48					
	Clerks & Secretaries	20000	3	54,979		10					
	Total Salaries		74	1,769,722						1.77	3.04
	Overhead/Maint	60%	of Labor & Supervison	1,061,833		18.40	PEP 2003 (uses 37.66)			1.06	1.82
	Maintenance	2%	of TPI	3,305,992			**avg salary (w/ benefits)**			3.31	5.68
	Insurance & Taxes	1.5%	of TPI	2,479,494			38264.26			2.48	4.26
	Costs			8,617,041						**8.62**	**14.81**
	Total Cash Cost									**89.26**	**150.32**
	Annual Capital Charge									**29.09**	**49.98**
	Annual Operating Cost									118.35	200.30

Table D-6. Hydrogen-Production Scenario Detailed Capital Investment Analysis

Total Capital Investment					
	2007 Dollars		Peters & Timmerhaus 5th Edition		This method of cos
Total Purchased Equipment Cost (TPEC)	**$52,683,115**	100%			
Purchased Equipment Installation	$20,546,415	39%	Percent of TPEC	13%	% of TIC
Instrumentation and Controls	$13,697,610	26%	Percent of TPEC	9%	% of TIC
Piping	$16,331,766	31%	Percent of TPEC	10%	% of TIC
Electrical Systems	$5,268,311	10%	Percent of TPEC	3%	% of TIC
Buildings (including services)	$15,278,103	29%	Percent of TPEC	10%	% of TIC
Yard Improvements	$6,321,974	12%	Percent of TPEC	4%	% of TIC
Service Facilities	$28,975,713	55%		18%	
Total Installed Cost (TIC)	**$159,103,006**	3.02			
Indirect Costs					
Engineering	$16,858,597	32%	Percent of TPEC	11%	% of TIC
Construction	$17,912,259	34%	Percent of TPEC	11%	% of TIC
Legal and Contractors Fees	$12,117,116	23%	Percent of TPEC	8%	% of TIC
Total Indirect	**$46,887,972**	4.69			
Project Contingency	$41,198,196	78.2%	Percent of TPEC	20%	% of TIC + IC
(Working Capital shown in DCFROR)	**$37,078,376**				
Total Fixed Capital Investment	**$247,189,174**	4.69			
Non-depreciated Direct Costs					
Land	$3,160,987	6.00%	Percent of TPEC		
Total Investment (with Land)	$287,428,536				
Lang Factor	5.46				

Table D-7. Hydrogen-Purchase Scenario Detailed Capital Investment Analysis

Total Capital Investment					
	2007 Dollars		Peters & Timmerhaus 5th Edition		This method of c
Total Purchased Equipment Cost (TPEC)	**$36,619,315**	100%			
Purchased Equipment Installation	$14,281,533	39%	Percent of TPEC	13%	% of TIC
Instrumentation and Controls	$9,521,022	26%	Percent of TPEC	9%	% of TIC
Piping	$11,351,988	31%	Percent of TPEC	10%	% of TIC
Electrical Systems	$3,661,931	10%	Percent of TPEC	3%	% of TIC
Buildings (including services)	$10,619,601	29%	Percent of TPEC	10%	% of TIC
Yard Improvements	$4,394,318	12%	Percent of TPEC	4%	% of TIC
Service Facilities	$20,140,623	55%		18%	
Total Installed Cost (TIC)	**$110,590,330**	3.02			
Indirect Costs					
Engineering	$11,718,181	32%	Percent of TPEC	11%	% of TIC
Construction	$12,450,567	34%	Percent of TPEC	11%	% of TIC
Legal and Contractors Fees	$8,422,442	23%	Percent of TPEC	8%	% of TIC
Total Indirect	**$32,591,190**	3.91		0.2	
Project Contingency	$28,636,304	78%	Percent of TPEC	20%	% of TIC + IC
(Working Capital shown in DCFROR)	**$25,772,674**				
Total Fixed Capital Investment	**$171,817,824**	4.69			
Non-depreciated Direct Costs					
Land	$2,197,159	6.00%	Percent of TPEC		
Total Investment (with Land)	$199,787,656				
Lang Factor	5.46				

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14. ABSTRACT *(Maximum 200 Words)*
This study develops techno-economic models for assessment of the conversion of biomass to valuable fuel products via fast pyrolysis and bio-oil upgrading. The upgrading process produces a mixture of naphtha-range (gasoline blend stock) and diesel-range (diesel blend stock) products. This study analyzes the economics of two scenarios: onsite hydrogen production by reforming bio-oil, and hydrogen purchase from an outside source. The study results for an nth plant indicate that petroleum fractions in the naphtha distillation range and in the diesel distillation range are produced from corn stover at a product value of $3.09/gal ($0.82/liter) with onsite hydrogen production or $2.11/gal ($0.56/liter) with hydrogen purchase. These values correspond to a $0.83/gal ($0.21/liter) cost to produce the bio-oil. Based on these nth plant numbers, product value for a pioneer hydrogen-producing plant is about $6.55/gal ($1.73/liter) and for a pioneer hydrogen-purchasing plant is about $3.41/ gal ($0.92/liter). Sensitivity analysis identifies fuel yield as a key variable for the hydrogen-production scenario. Biomass cost is important for both scenarios. Changing feedstock cost from $50–$100 per short ton changes the price of fuel in the hydrogen production scenario from $2.57–$3.62/gal ($0.68–$0.96/liter).

15. SUBJECT TERMS
biomass; biofuels; techno-economic analysis; pyrolysis; bio-oil; thermochemical; hydroprocessing

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