



# Techno-economic analysis of monosaccharide production via fast pyrolysis of lignocellulose

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## HIGHLIGHTS

- The cost of producing monosaccharides and fuels via fast pyrolysis was evaluated.
- Capital and feedstock costs, sugar yield and product values strongly influence IRR.
- The process can compete with traditional sugar production under certain scenarios.

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## ABSTRACT

The economic feasibility of a facility producing monosaccharides, hydrogen and transportation fuels via fast pyrolysis and upgrading pathway was evaluated by modeling a 2000 dry metric ton biomass/day facility using Aspen Plus<sup>®</sup>. Equipment sizing and cost were based on Aspen Economic Evaluation<sup>®</sup> software. The results indicate that monosaccharide production capacity could reach 338 metric tons/day. Co-product yields of hydrogen and gasoline were 23.4 and 141 metric tons/day, respectively. The total installed equipment and total capital costs were estimated to be \$210 million and \$326 million, respectively. A facility internal rate of return (IRR) of 11.4% based on market prices of \$3.33/kg hydrogen, \$2.92/gal gasoline and diesel, \$0.64/kg monosaccharide was calculated. Sensitivity analysis demonstrates that fixed capital cost, feedstock cost, product yields, and product credits have the greatest impacts on facility IRR. Further research is needed to optimize yield of sugar via the proposed process to improve economic feasibility.

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## 1. Introduction

Biofuels have historically been produced via a biochemical platform (i.e., pathways employing microorganisms) from sugar feedstocks, such as sucrose (from sugar cane) and glucose (from grain). Controversies regarding the production of biofuels from food crops and the growing of biofuel feedstocks on cropland have encouraged research into the development of sugars derived from lignocellulosic biomass for the production of biofuels, with the

**Abbreviations:** IRR, internal rate of return; TEA, techno-economic analyses; MSP, minimum selling price; GGE, gallon gasoline equivalent; EIA, Energy Information Agency; ESP, electrostatic precipitator; SF1, stage fraction 1; SF2, stage fraction 2; PSA, pressure swing adsorption; DCFROR, discounted cash flow rate of return; NPV, net present value; USDA, United States Department of Agriculture; HFCS, high fructose corn syrup; LHV, low heat value.

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most notable example being cellulosic ethanol. Whereas first generation ethanol plants use sugar crops such as sugar cane or sugar beet or starch crops such as corn or wheat, second-generation plants obtain sugars from lignocellulosic biomass. Unlike grain crops, lignocellulosic feedstocks have the potential to meet a significant fraction of U.S. demand for transportation fuels (Service, 2007). Lignocellulosic biomass, a natural composite of cellulose fibers, hemicellulose, and lignin, is recalcitrant to biological deconstruction, which complicates its use in biofuel production via biochemical pathways. The feedstock pretreatments to release sugars greatly increase production costs.

Techno-economic analyses (TEA) have found a minimum selling price (MSP) of cellulosic ethanol in the near-term ranging from \$3.23 to \$6.18 per gallon gasoline equivalent (GGE) (Aden, 2008; Gnansounou and Dauriat, 2010; Hamelinck et al., 2005; Kazi et al., 2010; Piccolo and Bezzo, 2009). This price is substantially higher than that of sugarcane ethanol with an estimated MSP range of \$1.94–2.04/gge (Cavalett et al., 2011; Tao and Aden, 2009). The

20-year average projected pre-tax price (i.e., prior to the imposition of federal and state excise taxes) for gasoline is \$2.86/gallon (EIA, 2011), suggesting that cellulosic ethanol will only be competitive with gasoline if petroleum prices endure a sustained rise that is substantially higher than projected, or if production costs are substantially reduced.

It is also possible to produce fermentable sugars from biomass via a thermochemical platform. Fast pyrolysis of pure cellulose produces bio-oil containing significant yields of the anhydrosugar, levoglucosan (Patwardhan et al., 2009). Levoglucosan is readily hydrolyzed to glucose and thus pyrolysis of cellulose is a potential pathway for the production of fermentable sugars. The presence of naturally occurring alkali and alkaline earth metals in lignocellulosic biomass has hampered the production of pyrolytic sugars since these metals catalyze the decomposition of cellulose into low molecular weight species at the expense of levoglucosan (Patwardhan et al., 2010). Kuzhiyil et al. (2012) have reported the ability to passivate the catalytic activity of alkali and alkaline earth metals by infusion of small quantities of sulfuric or phosphoric acid into the feedstock as a pretreatment prior to pyrolysis. The process is very efficient, with bio-oil from acid-infused lignocellulose feedstock yielding similar quantities of anhydrosugars as from pure cellulose.

Unfortunately, bio-oil is conventionally recovered as a single fraction, which results in the sugars being dissolved in the aqueous phase of the oil along with other soluble compounds such as carboxylic acids and aldehydes and complicates economic recovery of the sugars. Pollard et al. (2011) have devised a system for recovering bio-oil in several stages according to boiling point. Thus, instead of a single fraction of bio-oil, several fractions are produced, including high boiling point “heavy ends” consisting mostly of water-insoluble phenolic oligomers and water-soluble sugars and light ends consisting mostly of water, carboxylic acids, and aldehydes. Rover et al. (2011) have devised a simple washing procedure that recovers a concentrated stream of these sugars, which can be subsequently hydrolyzed to monosaccharide. The final product has high dextrose equivalence, making it closer to dextrose than glucose syrup (a solution of monosaccharides and oligosaccharides) when estimating its market value. The remaining water-insoluble fraction, or raffinate, is mostly phenolic oligomers that can be converted to gasoline and diesel via hydroprocessing (Elliott et al., 2009). Tests of the raffinate indicated that it was less viscous and acidic than whole bio-oil, which might facilitate its conversion to fuels (Rover et al., 2011, pers. comm.). The light ends can be reformed into hydrogen (Kechagiopoulos et al., 2006; Wang et al., 1998; Wright et al., 2010).

The ability to produce dextrose-like monosaccharides by pyrolyzing acid pretreated biomass has the potential to improve fast pyrolysis and upgrading economic feasibility. Past TEAs have found the 20-year IRR of fast pyrolysis facilities to be very sensitive to market values and yields of pyrolytic products (Brown and Hu,

2011; Brown et al., 2011; Brown et al., 2012; Jones et al., 2009; Manganaro and Lawal, 2012; Sarkar and Kumar, 2010; Wright et al., 2010). While past TEAs reported IRRs of 10% for fast pyrolysis and upgrading pathway at market prices in 2008 (Jones et al., 2009; Wright et al., 2010), discussions with industry representatives indicate that a 25% IRR is necessary for advanced biofuels technology to be attractive in capital markets (BIC, 2010). A TEA of the pathway under 20-year market price forecasts by the Energy Information Agency (EIA) calculates an IRR of 8% over the next 20 years (Brown and Hu, 2011), indicating that the pathway is not yet commercially attractive despite its ability to generate a positive IRR. The ability to yield another value-added product via fast pyrolysis might improve the economic feasibility of the pathway.

The objective of the present study was to quantify the economic feasibility of the production of monosaccharides as well as gasoline, diesel, hydrogen, and char via fast pyrolysis and upgrading. A process model of the pyrolysis reactor, sugars extractor, hydroprocessor, and steam reformer was formulated, the facility capital and operating costs were estimated, and the 20-year facility IRR was calculated.

## 2. Methods

### 2.1. Process modeling

The proposed pyrolytic process for producing sugar, gasoline, diesel, hydrogen, and char from bio-oil is described in Fig. 1. Seven areas were employed to simulate biomass depolymerization and product upgrading. These steps were pretreatment, fast pyrolysis, solids removal, bio-oil recovery, hydrolysis and extraction, hydroprocessing and heat generation. Descriptions of each are presented in Table 1.

In Area 100, the biomass is sent to a chopper where it is reduced to a particle diameter of approximately 10 mm. The biomass then enters a continuous spray rotary drum where sulfuric acid (0.4 wt.% of biomass) is infused into the biomass for the purpose of enhancing sugar production during pyrolysis. Acid is added as a mixture of sulfuric acid and water in a 1:1 mass ratio. The biomass then enters a drier where the moisture content is reduced from 25 to 7 wt.%. After drying, the biomass is ground to obtain particles 3 mm in diameter.

Table 2 shows the detailed properties of the red oak feedstock and char (Brewer et al., 2011; Ellens, 2009). Feedstock reception, storage and handling were not explicitly modeled in Aspen Plus; however, the analysis accounted for the equipment associated with these processes, including forklift, bale transport conveyor, magnetic separator and similar ancillary equipment.

Fast pyrolysis is the thermal decomposition of biomass in the absence of oxygen with a very short vapor residence time (~1 s) at a temperature of approximately 500 °C and a pressure of

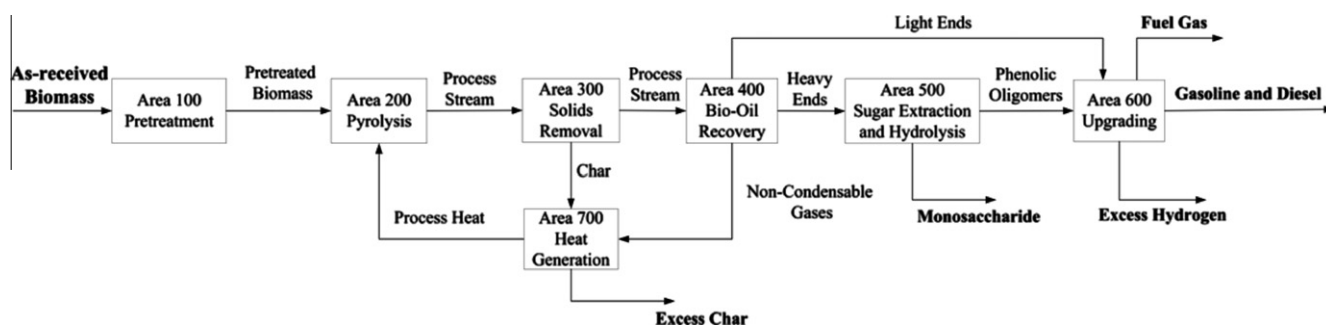


Fig. 1. Process diagram for production of sugars and hydrocarbons from bio-oil.

**Table 1**  
Description of each step employed in the fast pyrolysis system.

| Area | Description   |
|------|---|
| 100  | Biomass is chopped and ground to 3 mm diameter; biomass is pretreated with sulfuric acid; biomass is dried to 7% moisture   |
| 200  | Pretreated biomass is pyrolyzed in a fluidized bed reactor operating at 500 °C and 1 atm pressure   |
| 300  | Char and other particulate matter is removed from pyrolysis vapors via high efficiency cyclones   |
| 400  | Bio-oil is collected as stage fractions of heavy ends and light ends  |
| 500  | Sugars are washed from the bio-oil heavy ends; sugars are hydrolyzed to monosaccharides   |
| 600  | Transportation fuels are produced by hydroprocessing the phenolic oligomers obtained from the heavy ends of the bio-oil; hydrogen is produced by catalytic reforming of the light ends of the bio-oil |
| 700  | Excess char and non-condensable gasses from pyrolysis vapors are combusted with air to supply heat and fluidizing agent for pyrolysis   |

**Table 2**  
Ultimate and proximate analyses for red oak feedstock and char (wt.%) (Brewer et al. (2011); Ellens (2009)).

| Ultimate analysis (dry basis) |         |      | Proximate analysis (wet basis) |         |      |
|-------------------------------|---------|------|--------------------------------|---------|------|
| Element                       | Red oak | Char | Element                        | Red oak | Char |
| Carbon                        | 48.7    | 75.7 | Moisture                       | 3.9     | 3.7  |
| Hydrogen                      | 6.8     | 4.2  | Fixed content                  | 12.6    | 64.9 |
| Nitrogen                      | 0.07    | 0.3  | Volatile matter                | 81.9    | 27.1 |
| Chlorine                      | 0       | 0    | Ash                            | 1.7     | 4.3  |
| Sulfur                        | 0       | 0.01 |                                |         |      |
| Oxygen                        | 44.0    | 15.5 |                                |         |      |

1 atm (Bridgewater, 2007). In Area 200, a fluidized pyrolyzer was simulated as a jacketed reactor using the RYield module in Aspen Plus. A daily throughput of 2000 dry metric tons of biomass was assumed to be processed in four, 500 dry metric ton per day reactors instead of a single, large reactor. Composition and yield data for bio-oil and non-condensable gasses come from pilot-scale pyrolysis trials at the National Renewable Energy Laboratory (NREL) (Ringer et al., 2006). Data on thermal depolymerization of biomass to sugars came from Iowa State University, which found that plant polysaccharides in red oak were converted to anhydrosugar monosaccharides in yields of around 16 wt.% with higher yields expected with additional development of the process (Dalluge and Rover, 2011 pers. comm.). In the absence of a complete characterization of the other components of this sugar-rich bio-oil, these results were integrated with more complete bio-oil characterization data from NREL to estimate the mass fractions for the other constituents of the bio-oil (see Table 3).

The hot process stream exiting the fluidized bed pyrolyzer is a mixture of non-condensable gasses, condensable vapors, aerosols, and entrained particles of char and ash. In Area 300, high flow gas cyclones operating at 90% efficiency remove solid particles

**Table 3**  
Pyrolysis products (wt.% of biomass feedstock) (Ringer et al., 2006).

| Bio-oil                      |      | Gases           |       |
|------------------------------|------|-----------------|-------|
| Water                        | 21.2 | Nitrogen        | 0     |
| Methyl acetate               | 0.1  | Carbon dioxide  | 6.9   |
| Hydroxyacetaldehyde          | 5.3  | Carbon monoxide | 4.9   |
| Acetic acid                  | 5.0  | Methane         | 0.9   |
| Hydroxyacetone               | 0.8  | Ethane          | 0.4   |
| 2-Furanone                   | 0.2  | Hydrogen        | 0.004 |
| 3-Methyl-1,2-cyclopentadione | 0.2  | Propane         | 0.2   |
| Phenol                       | 0.03 | Ammonia         | 0     |
| 1-Hydroxy-2-butanone         | 0.2  | Steam           | 3.0   |
| Furfural                     | 0.2  | Acetylene       | 0.2   |
| 2-Cyclopenten-1-one          | 0.1  | Butene          | 0.2   |
| $\gamma$ -Butyrolactone      | 0.1  | Total           | 15.7  |
| 5-Hydroxymethylfurfural      | 0.2  |                 |       |
| Sugars                       | 16.0 | Solids          |       |
| Methanol                     | 0.2  | Char            | 19.1  |
| Lignin                       | 15.4 | Ash             | 0.04  |
| Total                        | 65.2 | Total           | 19.1  |

from the rest of the process stream. Area 400 collects bio-oil as two-stage fractions: heavy ends and light ends. The heavy ends, containing most of the sugars, are recovered by rapidly cooling the process stream from around 500 °C to just above the dew point of water (around 80 °C). This causes vapors to either condense on the walls of a condensing chamber or to form aerosols entrained in the process stream, which are collected with an electrostatic precipitator. The light ends of the bio-oil, consisting mostly of water and acetic acid, are collected by a condenser operated below room temperature followed by an electrostatic precipitator to recover any fume. Several stages of bio-oil fractionation can be employed in this manner (Pollard et al., 2011), although for this study, two stages were sufficient. The non-condensable gasses consist of hydrogen, carbon monoxide, and light hydrocarbons. These exit the bio-oil recovery system and are routed to Area 700 where they are burned, along with part of the biochar, to provide process heat to the pyrolyzer in Area 200. Excess biochar was assumed to be sold as boiler fuel.

The bio-oil fractions are processed into glucose, hydrocarbons, and hydrogen in Areas 500 and 600, as detailed in Fig. 2. The heavy ends enter Area 500 where a wash percolator recovers water-soluble sugars from water-insoluble phenolic oligomers. The sugars, which include mostly monosaccharides but also levoglucosan (an anhydrosugar) and a few disaccharides, are acid-hydrolyzed at 180 °C to yield glucose. It is important that the sugars are substantially hydrolyzed, as incomplete hydrolysis results in glucose syrup, a mixture of monosaccharides and oligosaccharides, which is less valuable than dextrose, a glucose-rich product (USDA, 2011). One means of differentiating between these two products is to determine their dextrose equivalent value (DE), which in turn is an indication of the degree of depolymerization of the product. Whereas glucose syrup has a DE of 38–44 (Jackson, 1995), dextrose has a DE greater than 60 (Wyman, 1996). Although the DE of pyrolytic sugars has yet to be determined, sugar analysis indicates that it is comprised of more than 90% monomers, high enough for the product to be considered dextrose (Rover et al., 2011 pers. comm.).

Area 600, designated as upgrading, consists of both a steam reformer to generate hydrogen from the light ends of the bio-oil and a hydrocracker for converting the phenolic oligomers to hydrocarbons. Steam reforming, which was assumed to follow the reaction mechanism described by Marquovich et al. (1999), converts the light ends to a syngas, containing hydrogen, carbon monoxide, carbon dioxide, and light hydrocarbons. A pressure swing adsorption (PSA) unit is used to separate pure hydrogen from the remainder of the syngas (fuel gas).

The fuel gas represents energy in excess of in-plant requirements and is sold for process heat applications. Part of the hydrogen is consumed in hydroprocessing with the balance sold at market prices. As detailed in Fig. 2, phenolic oligomers from the wash percolator and hydrogen from the steam reformer are reacted over a nickel-molybdenum catalyst at 400–450 °C and a pressure of 1500–2000 psi to produce hydrocarbons suitable for production of gasoline and diesel.

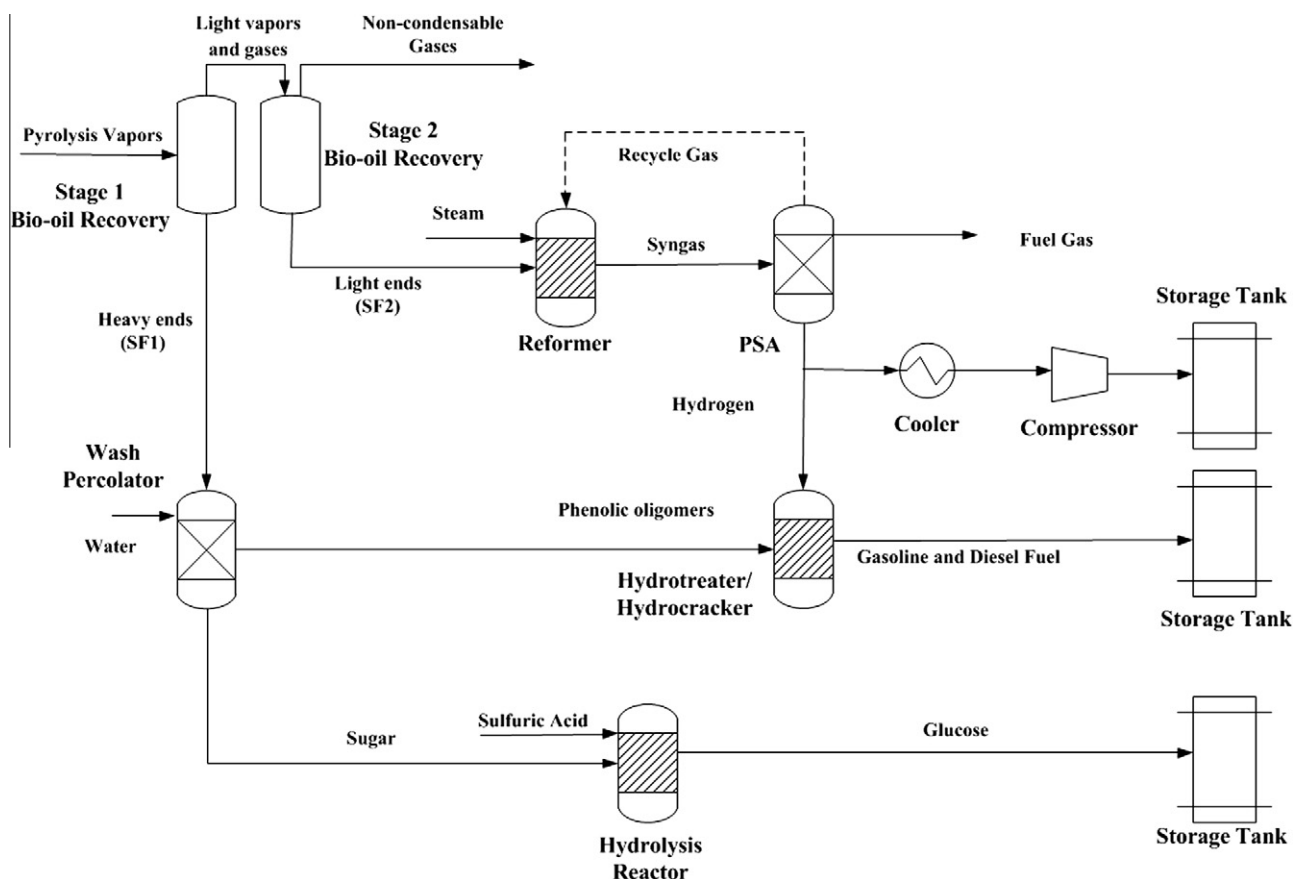


Fig. 2. Process diagram for processing bio-oil (recovery of sugars in Area 500 and steam reforming light ends and hydroprocessing phenolic oligomers in Area 600).

## 2.2. Cost estimation

Equipment sizing and cost estimation were based on the Aspen Economic Evaluation software (Table 4). The methodology developed by Peters et al., 2003 for installation costs and rate of return analysis was employed for the capital analysis. A Lang factor of 5.46 was employed to calculate the total investment cost based on the total purchased equipment cost, which has been employed in previous analyses of pyrolysis-based biofuels production

**Table 4**  
Methodology for capital cost estimation for nth plant.

| Parameter                             | Assumption         |
|---------------------------------------|--------------------|
| Total purchased equipment cost (TPEC) | 100%               |
| Purchased equipment installation      | 39%                |
| Instrumentation and controls          | 26%                |
| Piping                                | 10%                |
| Electrical systems                    | 31%                |
| Buildings (including services)        | 29%                |
| Yard improvements                     | 12%                |
| Service facilities                    | 55%                |
| Total installed cost (TIC)            | 3.02*TPEC          |
| Indirect cost (IC)                    | 0.89*TPEC          |
| Engineering                           | 32%                |
| Construction                          | 34%                |
| Legal and contractors fees            | 23%                |
| Total direct and indirect costs(TDIC) | TIC + IC           |
| Contingency                           | 20% of TDIC        |
| Fixed capital investment (FCI)        | TDIC + contingency |
| Working capital (WC)                  | 15% of FCI         |
| Land use                              | 6% of TPEC         |
| Total capital investment (with land)  | FCI + WC + land    |

(Wright et al., 2010). A modified discounted cash flow rate of return (DCFROR) analysis spreadsheet was employed to calculate the facility IRR in this study. The process design was based on the current state of technology and was assumed to be an nth plant operating for 20 years. All cost estimation assumed 2010 dollars in calculating net present value (NPV).

Table 5 details the variable operating parameters employed in the analysis. The U.S. spot price for sulfuric acid has ranged widely from \$50/metric ton to \$425/metric ton, so this analysis assumed a price of \$238/metric ton (Boyd, 2010). The red oak feedstock was assumed to be purchased for \$83/metric ton, which is in line with existing facility-gate cost estimates for forest thinnings and logging residues (Council & Production, 2011; DOE, 2011). The cost of electricity and market prices for synthetic gasoline and diesel, hydrogen and char were based on the averages for EIA 20-year

**Table 5**  
Material and operating parameters employed in the evaluation.

| Material                  | Price                              |
|---------------------------|------------------------------------|
| Feedstock                 | \$83/metric ton                    |
| Electricity               | \$0.061/kwh                        |
| Process water             | \$0.032/metric ton                 |
| Sulfuric acid             | \$0.23/kg                          |
| Hydroprocessing catalyst  | \$1.77 million/year (Meyers, 1997) |
| Solids disposal cost      | \$19.84/metric ton                 |
| Fuel gas                  | \$5/MMBTU                          |
| Dextrose                  | \$0.64/kg                          |
| Synthetic gasoline/diesel | \$2.92/gallon                      |
| Hydrogen                  | \$3.33/kg                          |
| Char                      | \$18.21/metric ton                 |



price forecasts (EIA, 2011). This analysis used pre-tax prices for all products and \$0.48/gal of gasoline and \$0.53/gal of diesel were subtracted from the EIA prices, which included national and state taxes, to account for this (API, 2011). The price of hydrogen was calculated as a function of natural gas prices under the assumption that it was produced via steam reforming (NREL, 2006). Char, with a heating value that can be as low as one-half that of coal, was assumed to be sold as boiler fuel at one-half the 20-year average price of U.S. coal.

In addition to variable operating costs, fixed operating costs were also considered, which included labor, overhead, maintenance, insurance and taxes. Total labor cost was estimated to be \$1.89 million while total fixed operating costs of were \$10.55 million. While the DE of the sugars produced via fast pyrolysis were not determined, monomeric sugars have high DE levels (Wyman, 1996). Detailed analysis of the sugar shows very high levels of monosaccharides, which is characteristic of dextrose. Accordingly, for marketing purposes, this analysis assumed that the sugar product was dextrose. While historical prices for dextrose are available from the United States Department of Agriculture (USDA, 2011), projected prices are not available and were therefore calculated based on the historical 11-year correlation ( $R_2 = 0.93$ ) between the prices of dextrose and high fructose corn syrup (HFCS) and 10-year FAPRI price forecasts for HFCS (FAPRI, 2010). The average forecast dextrose price for 2011–2020 calculated via this methodology was \$0.64/kg.

A Monte-Carlo simulation was performed to obtain an IRR distribution for the proposed facility. The simulation was performed with Crystal Ball® software and analyzed through JMP® software.

### 3. Results and discussion

#### 3.1. Process modeling results

The input/output mass and energy balances are shown in Table 6 for a 2000 dry metric tons/day fast pyrolysis and upgrading facility that uses red oak as cellulosic feedstock. Monosaccharides are the main output product as measured by net energy output (198 GJ/h). Of the char product, 57% is combusted for process heat. Nevertheless, char in excess of what is required for process heat is the second largest net energy output at 182 GJ/h. The fuel gas from the pressure swing absorption unit contributes a relatively modest net energy output of 15 GJ/hr. Of the hydrogen generated through the reforming of light ends, 40% goes to the hydroprocessing operations. The net energy output of the excess hydrogen is 117 GJ/hr. Transportation fuels (synthetic gasoline and diesel fuel) represent 29% of the total energy output of 913 GJ/h. With a total energy input of 1607 GJ/h, the overall energy efficiency for this facility on a LHV basis is approximate 57%.

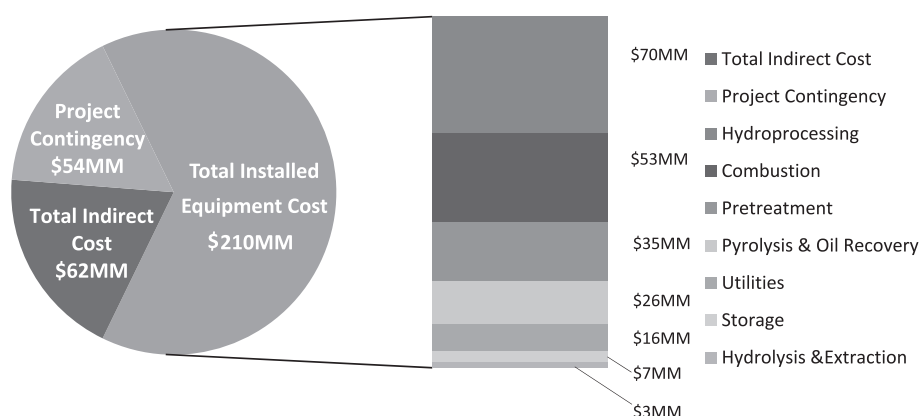
#### 3.2. Economics estimation results

The total investment cost is illustrated in Fig. 3. It is the sum of the project contingency, indirect costs and installed equipment costs of \$54, \$62 and \$210 million, respectively. The installed equipment costs for different areas are analyzed in the stacked bar chart in Fig. 3. The installed equipment costs for the hydroprocessing area is the largest portion of the total installed equipment cost, accounting for approximately 33% of the total, or \$70 million. This cost is driven by the employment of a hydrotreater and hydrocracker, a reformer, and a boiler in the hydroprocessing area to convert bio-oil fractions to synthetic gasoline, diesel, and hydrogen. The installed equipment costs for the combustion area are also quite large with a value of \$53 million due to the high costs of the boiler and boiler feedwater preheaters. Pretreatment equipment costs are primarily driven by the inclusion of a continuous spray rotary drum and drier. The steam for biomass drying is also generated in the pretreatment area, requiring the purchase of steam blowers and heat exchangers.

Total annual facility operating costs are \$111 million (see Fig. 4). Feedstock costs are responsible for the majority of the operating costs at \$54.4 million annually, followed by capital depreciation (\$16.3 million), fixed costs (\$14.4 million), and corporate income tax payments (\$12.4 million). Catalyst costs (\$1.8 million) include hydroprocessing catalysts and sulfuric acid for pretreatment and hydrolysis. These results are similar to other techno-

**Table 6**  
Mass and energy balance for production of monosaccharide and hydrocarbons from 2000 metric tons/day red oak fast pyrolysis plant.

| Materials      | Mass (metric tons/day) | LHV (MJ/kg) | Energy (GJ/hr) | Values (\$million/year) |
|----------------|------------------------|-------------|----------------|-------------------------|
| <i>In</i>      |                        |             |                |                         |
| Biomass        | 2000                   | 18.7        | 1560           | 54.43                   |
| Electricity    | n/a                    | n/a         | 49             | 6.52                    |
| <i>Out</i>     |                        |             |                |                         |
| Hydrogen       | 23.4                   | 120         | 117            | 25.65                   |
| Gasoline       | 70.5                   | 44.4        | 130            | 22.97                   |
| Diesel         | 70.5                   | 44.2        | 130            | 22.97                   |
| Monosaccharide | 338                    | 14.1        | 198            | 78.96                   |
| Fuel Gas       | 341                    | 11          | 156            | 5.78                    |
| Char           | 159                    | 27.5        | 182            | 1.05                    |



**Fig. 3.** Installed equipment cost for red oak to monosaccharide and hydrocarbons.

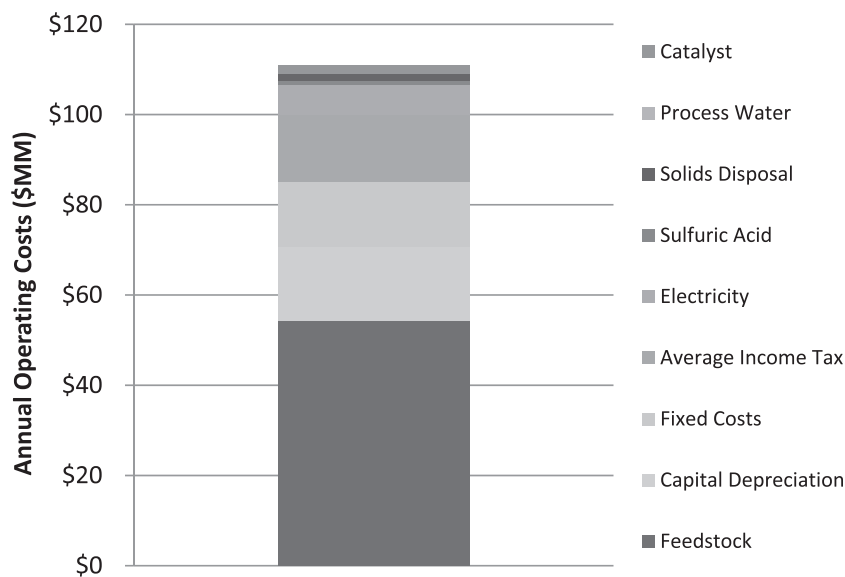


Fig. 4. Annual operating costs for red oak to monosaccharide and hydrocarbons.

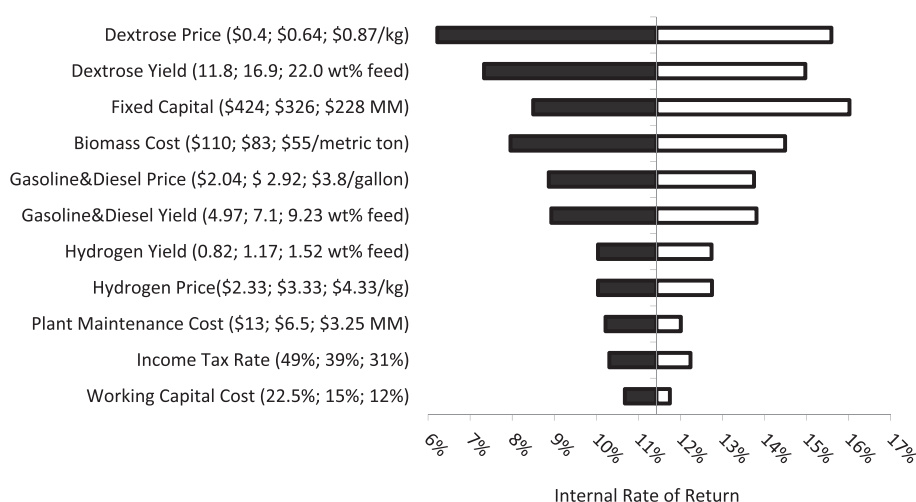


Fig. 5. Sensitivity analysis for facility IRR of a 2000 metric tons/day red oak fast pyrolysis to monosaccharide and hydrocarbons plant.

economic analyses, which find that feedstock costs, fixed costs, and capital depreciation to comprise the vast majority of annual operating costs for fast pyrolysis and hydroprocessing facilities (Brown et al., 2011; Brown et al., 2012; Wright et al., 2010). Based on the calculated capital costs, operating costs, and the assumptions, a facility IRR of 11.4% was obtained through DCFROR analysis, setting a goal of zero NPV at the end of this 20-year project.

### 3.3. Uncertainty analysis

The results of this study are based on an economic analysis that is deterministic and assumes that all of the parameters are precisely known; however, many of the costs and parameters used to evaluate the profitability of the chemical process are subject to significant volatility throughout the project's 20-year life. Uncertainty analysis enables an evaluation of the sensitivity that IRR has to changes in these parameters.

A sensitivity analysis on IRR was conducted by changing several parameters to which IRR is sensitive. These parameters included dextrose price, fixed capital cost, monosaccharide yield, biomass

cost, co-product prices and yields, corporate income tax rate, plant maintenance costs and working capital. As shown in Fig. 5, dextrose price, fixed capital cost, monosaccharide yield, biomass cost, co-products yields, and co-products values have the greatest impact on IRR. An increase in dextrose price from \$0.40/kg to \$0.87/kg increases facility IRR from 6.2% to 15.6%. A  $\pm 30\%$  range in fixed capital cost results in an IRR range of 8.5–16%. A monosaccharide yield increase from 11.8 to 22 wt.% results in an increase of facility IRR from 7.3% to 15%. A  $\pm 33\%$  range in biomass cost results in an IRR range of 8% to 14.5%. Biomass cost is also important in determining facility IRR. When the biomass price was reduced from \$110 to \$55 metric ton, the facility IRR increased from 8% to 14.5%. Co-products in the sensitivity analysis included hydrogen, gasoline, and diesel. Gasoline and diesel yields and prices have considerable effects on facility IRR, which varied from 8.9% to 13.8% when gasoline and diesel prices were increased from \$2.04/gal to \$3.80/gal. When gasoline and diesel yields increased from 4.97% to 9.23%, IRR varied from 8.9% to 13.8%. A  $\pm 30\%$  range in hydrogen yield and hydrogen price both resulted in a facility IRR range of 10–12.7%, which indicates that hydrogen has a

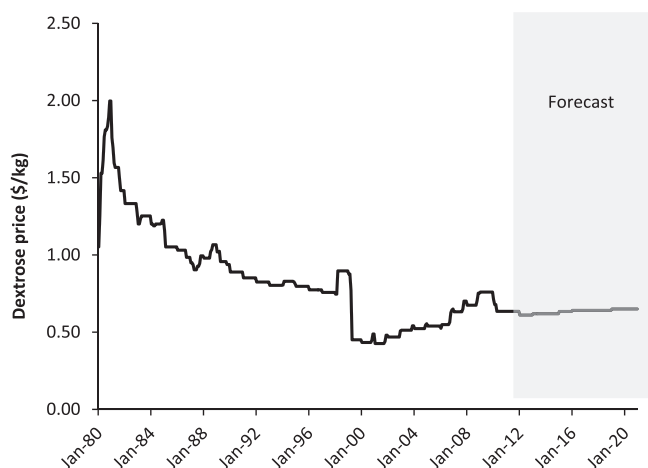


Fig. 6. Historical and projected dextrose prices, 2010 dollars. Adapted from USDA (USDA, 2011), FAPRI (FAPRI, 2010).

comparatively smaller impact on facility IRR than gasoline and diesel among the co-products. Corporate income tax rate, plant maintenance costs, and working capital cost have comparatively small impacts on facility IRR, resulting in a range of IRR between 10.22 and 12%. The cost-effective production of monosaccharides via the biomass fast pyrolysis pathway is mainly affected by capital cost, product yields, and market factors such as product and co-product prices.

The changing variables in the Monte-Carlo were dextrose price, monosaccharide yield, fixed capital cost, biomass cost, gasoline and diesel yield, and gasoline and diesel price as the sensitivity analysis found these parameters to have the greatest impact on facility IRR. All of these variables were assumed to follow triangular distributions with variation ranges discussed in the sensitivity analysis. Two thousand random facility IRRs were generated during the Monte-Carlo simulation. Through the analysis, the expected value of facility IRR (that is, mean IRR) was found to be  $11.2 \pm 3.1\%$ . The minimum IRR is 0.7% and the maximum IRR is 20.5%. The median, 25% quartile, and 75% quartile facility IRRs are 9.1%, 11.3% and 13.4%, respectively. For the cumulative probability distribution of the facility IRR analysis, more than 65% of facilities will have IRRs exceeding 10%. This shows that the economic environment presents a low risk to this project investment.

The fast pyrolysis and hydroprocessing facility in this analysis was assumed to have an operating life of 20 years. Obtaining general price trends for facility products and co-products by examining historical data and forecasts is therefore helpful in accurately determining economic feasibility during that time period. Fig. 6 shows the monthly historical dextrose prices and price forecasts through 2020. This figure shows that dextrose has a history of volatility. While the forecasted average annual prices for dextrose over the next decade suggest that the sugars extraction process will contribute to the overall economic feasibility of the fast pyrolysis facility, this result is highly sensitive to the market price of dextrose (Fig. 5). The IRR for a fast pyrolysis and upgrading facility without monosaccharide production under identical economic assumptions has been found to be 8.15% (Brown and Hu, 2011); a facility employing the proposed sugars extraction process has a greater IRR when dextrose prices are above \$0.49/kg, which is higher than prices experienced as recently as 1997–2006.

#### 4. Conclusions

This paper evaluates the economic feasibility of a biomass fast pyrolysis and upgrading facility producing monosaccharides,

synthetic gasoline and diesel fuel, hydrogen, fuel gas, and char. The analysis shows that monosaccharide production via the biomass fast pyrolysis and upgrading pathway modestly improves the economic feasibility of the pathway assuming sugar yields of 16wt%. As shown by sensitivity analysis, higher sugar yields, which are expected from process improvements, or higher market values would significantly increase IRR, which may allow it to compete with sugars from enzymatic hydrolysis of cellulose or even from starch.

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