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Techno-economic analysis of advanced biofuel production based on bio-oil gasification

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Keywords
techno-economic analysis, bio-oil gasification, fast pyrolysis, Mechanical Engineering

Disciplines
Industrial Engineering | Mechanical Engineering | Systems Engineering

Techno-economic analysis of advanced biofuel production based on bio-oil gasification

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This paper evaluates the economic feasibility of a hybrid production pathway combining fast pyrolysis and bio-oil gasification. The conversion process is simulated with Aspen Plus® for a 2000 t d^{-1} facility. Techno-economic analysis of this fast pyrolysis and bio-oil gasification pathway has been conducted to assess the economic feasibility. A total capital investment of $438 million has been estimated and the minimum fuel selling price (MSP) is $5.6 per gallon of gasoline equivalent. The sensitivity analysis shows that the MSP is most sensitive to internal rate of return, fuel yield, biomass feedstock cost, and fixed capital investment. Monte-Carlo simulation shows that MSP for bio-oil gasification would be more than $6/gal with a probability of 0.24, which indicates this pathway is still at high risk with current economic situation.

Keywords

Techno-economic analysis; Bio-oil gasification; Fast pyrolysis

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1. Introduction

Biofuels are playing an increasingly important role as a cleaner substitute for fossil-based fuels. Second generation biofuels such as corn stover, switchgrass, and woody biomass are made from nonedible plant residues or dedicated energy crop which are less land and water intensive (Carriquiry, et al. 2011). The revised Renewable Fuel Standard (RFS2) has been enacted to accelerate the domestic biofuel production and consumption. The RFS2 mandates that by the year 2022, at least 36 billion gallons per year of renewable fuels will be produced and blended into the transportation fuel, of which at least 16 billion gallons per year should be produced from cellulosic biomass feedstock (Schnepf 2011).

Biomass can be converted to transportation fuels through a variety of production pathways, including biochemical and thermochemical platforms. Biochemical process such as fermentation and anaerobic digestion are limited by high viscosity substrate, high enzymes cost and process complexity (Chen 2012; Zhang, et al. 2014b). Recently, thermochemical conversion of biomass (mainly pyrolysis and gasification) has attracted increasing attention (Brown 2015).

Fast pyrolysis produces bio-oil (70-75 wt%), bio-char (10-15 wt%) and non-condensable gases (10-15 wt%) by thermally decomposing organic feedstock in the absence of oxygen (Butler, et al. 2011; Zhang, et al. 2014a). Fast pyrolysis can be combined with other bio-oil upgrading pathways (e.g., hydrotreating and hydrocracking) to produce transportation fuels and hydrogen (Zhang, et al. 2013c), although bio-oil upgrading is a challenging process for its low fuel quality and conversion efficiency. On the other hand, biomass gasification followed by the Fischer-Tropsch (FT) synthesis is a relatively mature technology to produce liquid fuels (Timilsina and Shrestha 2011) for high efficiency cycles (Knoef and Ahrenfeldt 2005). However, commercialization of biomass gasification is hampered by high logistic cost of feedstock. In recent years, it has been suggested that combing fast pyrolysis and gasification by going through bio-oil gasification would possibly overcome this technical and logistic challenges (Badger and Fransham 2006; Li and Hu 2014; Zhang, et al. 2013a).

There is an increasing literature on techno-economic analysis (TEA) of a variety of advanced biofuel production pathways with a range of feedstock and products. The techno-economic analysis of biomass gasification using corn stover as the feedstock by Swanson et al. claimed that the minimum fuel selling price (MSP) is \$4-5 per gallon of gasoline equivalent (/GGE) and the capital investment requirement is \$500-650 million for a 2000 metric ton per day (t d^{-1}) facility (Swanson, et al. 2010). Zhang et al. conducted a techno-economic analysis of biohydrogen production via bio-oil gasification and concluded that an IRR of 8.4% is realized with the prevailing market price (Zhang, et al. 2013a). Wright et al. reported that the capital cost of gasification plant with a capacity of 550 million GGE per year is about 1.47 billion (Wright, et al. 2008). The capital cost of fast pyrolysis facility with a capacity of 2,000 t d^{-1} is \$200 million and an MSP is 2.11 \$/GGE under purchasing hydrogen scenario (Wright, et al. 2010). Manganaro and Lawal analyze a pathway that combined fast pyrolysis, and FT synthesis. A TPI of \$231 million is calculated for a 2000 MTPD biorefinery with a MSP of 3.74 \$/GGE, assuming an 8% IRR and \$61.20/MT feedstock cost (Manganaro and Lawal 2012).

Although there are a large number of techno-economic analyses on biomass gasification or fast pyrolysis, the process design and techno-economic analysis of the hybrid pathway have not been studied extensively. This study aims to model the production process and evaluate the economic feasibility based on nth plant design. In this study, an integrated production pathway combining fast pyrolysis and bio-oil gasification is investigated. Cellulosic biomass such as corn stover is firstly converted to bio-oil through fast pyrolysis and then bio-oil will go through gasification process to produce the syngas followed by catalytic FT synthesis and hydroprocessing to produce transportation fuels. This hybrid pathway offers several advantages. Firstly, bio-oil can be produced in relatively small-sized fast pyrolysis plants at distributed locations and shipped to centralized biorefinery so that high cost of shipping bulky solid biomass over long distance could be avoided. Secondly, liquids are relatively easy to pump to high pressure than solids, so high pressure gasification technology can be implemented to improve conversion efficiency. Thirdly, as most of nitrogen and potassium are left in biochar after the fast pyrolysis, bio-oil has reduced level of ash and other contaminants, which makes the syngas cleanup easier (López-González, et al. 2014; Van Rossum, et al. 2007; Venderbosch, et al. 2002).

The rest of paper is organized as follows: in Section 2, the methodology is presented with a focus on the process design. Then, techno-economic analysis results and analysis are discussed in Section 3. Finally, we conclude the paper in Section 4 with summary of research findings and discussion of research directions.

2. Methods

In this section, the method for this techno-economic study is presented. Materials and technologies are firstly selected according to commonly adopted criteria. Aspen Plus® process engineering software is employed to develop the detailed process model. Capital and operation costs of the plant are evaluated using the output of process models and literature data.

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2.1. Material and Technologies

A variety of feedstock and operational design decisions are to a certain degree subjective and flexible for the bio-oil gasification pathway, e.g., gasification conditions, syngas cleanup techniques, and fuel synthesis methods. These system operating parameters are selected following commonly adopted criteria in literatures: (i) the technology should be commercialized in the next 5-8 years; (ii) adequate feedstock should be provided by the current agricultural system; (iii) the final products are compatible with the current transportation fuels (Anex, et al. 2010; Swanson, et al. 2010).

Iowa possesses the largest quantity of corn stover, an important type of cellulosic biomass, in the United States (Tyndall, et al. 2011). Corn stover is therefore selected as feedstock of this pathway in this study. The ultimate and proximate analyses of corn stover are listed in Table 1 (Wright, et al. 2010). The plant capacity is set to be 2000 t d^{-1} dry biomass for consistency and comparison with literatures (Anex, et al. 2010; Swanson, et al. 2010; Wright, et al. 2010). The fluidized bed gasifier operates in low temperature (870 ºC) for gasification and FT synthesis is adopted for transportation fuel production.

{Insert Table 1 here}

2.2. Process Design

A thorough process model was established in Aspen Plus®. The model developed in this study is based on several previous models developed at Iowa State University (Swanson, et al. 2010; Wright, et al. 2010; Zhang, et al. 2013a). A schematic of the generalized process flow diagram is shown in Figure 1. Detailed information of functions for each area is included in Table2. The major components include biomass preprocessing, bio-oil production (fast pyrolysis), bio-oil gasification, syngas cleanup, and fuel synthesis.

{Insert Table 2 here}

{Insert Figure 1 here}

<u>Biomass preprocessing & fast pyrolysis process</u>

Biomass preprocessing (chopping, drying, and grinding) are conducted before the pyrolysis process. Solids removal and bio-oil recovery are included to condense and collect the bio-oil. In the biomass pretreatment including chopping, drying and grinding, biomass with 25% moisture is dried to 7% moisture and ground from 10-25mm to 3 mm diameter size prior to feeding into the pyrolyzer. The fluidized bed pyrolyzer operates at 500 ℃ and atmospheric pressure. As shown in Table 3, data from previous techno-economic analysis of pyrolysis-based biofuels are employed to build RYield module in Aspen Plus® (Badger and Fransham 2006; Wright, et al. 2010).

{Insert Table 3 here}

Standard cyclones remove solids consisting mostly of char particles entrained in the vapors exiting the pyrolyzer (90% particle removal rate (Wright 2010)). It is assumed that the solid products and non-condensable gases are sent to a combustor to provide heat for the drying and pyrolysis process. The char composition analysis is shown in Table 1 (Wright, et al. 2010). Ash and char are removed from the raw bio-oil through the cyclones with 90% particle removal rate. The electrostatic precipitators (ESP) and condensers are used to collect liquid phase in bio-oil recovery process.

Bio-oil gasification process

In the bio-oil gasification system (as shown in Figure 2(a)), 95% purity oxygen and steam are employed as the gasifying agent. The bio-oil is a mixture of all fractions from the fast pyrolysis, so-called "whole bio-oil". The gasifier operates at a pressure of 28 bar and a temperature of 870 ℃ (Swanson, et al. 2010). The mass ratios of oxygen to bio-oil are set to be 0.3 and the mass ratios of steam to bio-oil are set to be 0.2. After gasification, a separator is used to remove the slag. The syngas contains some particulate as well as all the ammonia, hydrogen sulfide, and other contaminants which need cleanup. A direct water quench is employed to reduce the syngas temperature to about 40 ℃ to condense tar and most of ammonia and ammonium chloride (Zhang, et al. 2013a). Carbon dioxide and nitrogen hydrogen sulfide are removed in acid gas removal system with monoethanolamine.

{Insert Figure 2 here}

FT synthesis process

In the catalytic FT synthesis, one mole of CO reacts with two moles of H_2 to form mainly aliphatic straight-chain hydrocarbons (Equation (1)). Typical FT catalysts are based on iron or cobalt. The optimal ratio of H_2/CO is around 2.1 according to the previous study (Swanson, et al. 2010). When the feed gas H_2/CO ratio is lower than 2.1, water-gas shift (WGS) reaction (Equation (2)) is used to increase the ratio to 2.1. Typical operation conditions for FT synthesis, when aiming for long-chain products, are under temperatures of 200-250 ºC and pressures of 25-60 bars (Anderson, et al. 1984).

$$\mathrm{CO} + 2.1\mathrm{H_2} \Rightarrow -(\mathrm{CH_2})- + \mathrm{H_2O} \quad (1)$$

$$\mathrm{CO} + \mathrm{H_2O} \Leftrightarrow \mathrm{CO_2} + \mathrm{H_2} \quad (2)$$

As shown in Figure 2(b), major operations in this area include zinc oxide/activated carbon gas polishing, syngas booster compression, steam methane reforming (SMR), WGS, pressure swing adsorption (PSA), FT synthesis, FT product separation, and unconverted syngas recycle (Swanson, et al. 2010).

Appropriate pretreatment must be taken so that the syngas entering FT synthesis contaminants below 200 ppb sulfur and 10 ppm ammonia at a pressure of 25 bar (Spath and Dayton 2003). First, a zinc oxide and activated carbon gas polishing is used to polish sulfur and trace contaminants. Next, the syngas stream is compressed to 25 bar in syngas booster compression unit. Experimental data indicate that there is a significant amount of methane and ethane in the syngas stream in the low temperature bio-oil gasification scenario. Thus, a SMR is utilized to reduce those components. As mentioned, a WGS unit is included to adjust syngas H_2/CO ratio to just above the optimal value for FT synthesis. After that, PSA is used to provide hydrogen for the hydroprocessing section. Next, the syngas reacts over a cobalt-based catalyst in a fixed-bed FT reactor at 200 ºC. The Anderson-Schulz-Flory alpha chain growth model described by Song et al. is used to predict the FT product distribution (Song, et al. 2004). After the gas is cooled, the liquid hydrocarbons and water are separated before the hydroprocessing section. The unconverted syngas is partially recycled back into the FT reactor while the other portions go back to the acid gas removal system in syngas cleanup section.

3. Economic Analysis

Literature data and Aspen Economic Evaluation® software are employed to estimate the facility cost for this pathway. Unit costs for equipment are scaled from base equipment costs by using Equation (3). $Cost_{new}$ is the scaled new equipment cost and $Cost_0$ is the base equipment cost; $size_{new}$ is the size of new equipment and $size_0$ is the size of base equipment; I is the inflation index of calculated year and I_0 is the inflation index of the base year. n is the specific scaling factor for a particular type of equipment ranging from 0.6 to 0.8. The scaling factor and some base equipment cost come from literature (Swanson, et al. 2010; Wright, et al. 2010; Zhang, et al. 2013a). The estimated costs have been adjusted to the 2013 US dollars.

$$Cost_{new} = \left(\frac{I}{I_0}\right) * Cost_0 * \left[\frac{size_{new}}{size_0}\right]^n \quad (3)$$

Aspen Economic Evaluation software is employed to estimate equipment size and calculate project capital expenditures. The methodology developed by Peters et al. is used for calculating installation costs (Peters, et al. 1968). A total installation factor of 3.02 is used to estimate the installed equipment costs (Zhang, et al. 2013a). A Lang Factor of 5.46 is chosen to estimate the total capital investment (TCI) (Wright, et al. 2010; Zhang, et al. 2013a; Zhang, et al. 2013b). Table 4(a) provides a summary of methodology for capital cost estimation.

{Insert Table 4 here}

Table 4(b) provides the assumptions and references for the material and operating cost estimation. The electricity price are based on the average 20-year forecast from Energy Information Administration (EIA) Annual Energy Outlook 2014 (EIA 2014). The facility-gate corn stover feedstock price is assumed to be 83 \$ t^{-1} (Downing, et al. 2011). The solid and waste water disposal costs are based on biomass gasification design (Swanson, et al. 2010). The costs for process water, fuel gas, and steam are based on bio-oil gasification to produce hydrogen (Zhang, et al. 2013a).

A modified National Renewable Energy Laboratory (NREL) discounted cash flow rate of return (DCFROR) analysis spreadsheet is employed to evaluate the economic feasibility with the IRR under the prevailing market conditions. Assumptions in DCFROR analysis are listed in Table 4(c) (Zhang, et al. 2013a). The process design is assumed to be the *n*th plant with a life cycle of 20 years based on the current state of technology.

4. Results and Analysis

4.1. Process Modeling

The raw corn stover is assumed to be with 25% moisture, and the moisture level is reduced to 7% with pretreatment. The fast pyrolysis process has a capacity of 2000 t d^{-1} dry corn stover and the yield of wet bio-oil (with a moisture content of 15%) is 63%, which means it will yield 1260 t d^{-1} of wet bio-oil. The transportation fuel yield for gasoline and diesel are 170 t d^{-1} and 69 t d^{-1}, representing 13.5% and 5.5% of the wet bio-oil, respectively. The comparisons of fuel yield for different pathways are included in Table 5.

{Insert Table 5 here}

Gasification experiments have been conducted with whole red oak bio-oil at Iowa State University. The gasification reactor runs at 850 °C. Pure oxygen was maintained at an equivalence ratio of 25% for full combustion. The bio-oil gasification yields are estimated based on the experiment with similar feedstock and literature data (Swanson, et al. 2010; Zhang, et al. 2013a). Table 6 shows the comparison of gasification conditions and syngas composition.

{Insert Table 6 here}

4.2. Economics Results

Estimated total installed equipment cost (TIEC) for 2000 t d^{-1} facility is $273 million. As mentioned in Table 4, total capital investment (TCI) is the summation of total installed equipment cost, total indirect cost ($92 million), project contingency ($73 million), working capital cost ($66 million), and land use ($6 million), which is $510 million in this study.

The free on board (FOB) equipment cost and installed equipment cost are breakdown to process area. Figure 3(a) shows the percentage of equipment cost and installed cost for each model area. Fast pyrolysis, combustion and fuel synthesis contribute 48% of equipment cost and 42% of installed cost.

{Insert Figure 3 here}

Stream mass flows in the Aspen Plus model and current market prices of the products are used to calculate the total annual operating costs. The fixed operating costs include salaries, maintenance cost, and insurance. The costs of cooling water, steam, waste disposal etc. are included in other variable operating costs category. As show in Figure 3(b), the biomass feedstock cost, about $54.3 million, is the largest (46%) contributor to annual operating costs. This is due to the high cost to collect and transport corn stover.

As commonly used in literature, an IRR of 10% is assumed in this analysis as a base line (Chau, et al. 2009; Swanson, et al. 2010). Based on the estimated capital costs, operating costs and IRR, an MSP of 5.59 $/GGE is calculated for the bio-oil gasification pathway.

This techno-economic analysis shows higher capital investment and MSP compared to the previous techno-economic studies on thermochemical production pathways (shown in Table 5). This is mainly due to the conservative assumptions on fuel yields and installation factor. Additionally, capital and operational costs are all adjusted to cost year 2013, therefore, the economic feasibility is affected by the rapid escalation in construction and equipment costs in recent years.

Figure 4 summarizes the variation in facility IRRs and MSP at different capacities. If we fixed the IRR to be 10%, the MSP will drop rapidly from 14.96 \$/GGE to 5.59 \$/GGE as facility size increases from 500 to 1900 t d^{-1} and drop at a slower rate beyond 1900 t d^{-1}. When the facility size is 5000 t d^{-1}, the MSP is about 3.5 \$/GGE, which is the average gasoline price for the next 20 years according to EIA's prediction (EIA 2014). On the other hand, if we assume the selling price of biofuel to be \$3.5/gal, we can analyze the relationship between the facility IRR and facility capacity. In this case, a minimum facility capacity of 1900 t d^{-1} is necessary for a positive IRR. Larger capacities are in favor due to the economies of scale.

{Insert Figure 4 here}

4.3. Uncertainty Analysis

The sensitivity analysis is presented in Figure 5 to demonstrate the sensitivity of MSP to changes in the parameters. The parameters under investigation are IRR, fuel yield, feedstock cost, fixed capital cost, catalyst cost, balance of plant (BOP), and availability operating hours. The analysis finds that MSP is most sensitive to IRR, fuel yield, feedstock cost, and fixed capital cost. IRR is influential because it affects the entire cash flow. At this stage, it is projected that there is room for improvement in the fuel yield, which will make this pathway more competitive since a 20% increase in fuel yield will lead the MSP from 5.59 \$/GGE to 4.66 \$/GGE. As a significant portion of operating costs, feedstock price is a highly sensitive parameter. The fixed capital cost affects the capital depreciation and average income tax, a ±20% range in fixed capital cost results an MSP in a range of 5.02 \$/GGE to 6.17 \$/GGE.

{Insert Figure 5 here}

The sensitivity analysis considers the influence of one parameter on the MSP at a time by assuming other parameters hold constant, while in reality, all these parameters may change simultaneously. For the uncertainty analysis, Monte-Carlo (MC) simulations are conducted to understand the effect of all key parameters interacting simultaneously. Distributions for key parameters are determined based on literature and prior knowledge. Simulation data for key parameters are generated from their distributions. These simulated data then serve as the input to analyze the empirical distribution of MSP and to quantify the uncertainty of economic feasibility of the bio-oil gasification pathway. R software is employed to conduct the MC simulation and analyze the results. The iterations number for the MC simulation is set to be 5000.

IRR, fuel yield, fixed capital cost, and biomass cost are treated as key parameters (changing variables) since these parameters are shown by the sensitivity analysis to have the most significant impact on MSP. Due to data availability limitation, all of these variables are assumed to follow triangular distributions with the same variation ranges used in the sensitivity analysis as suggested in literature (Zhang, et al. 2013a; Zhang, et al. 2013c). In order to analyze the impact of distribution selection in the MC simulation, a second scenario where these key parameters are assumed to follow normal distribution with means equal to their base level and variances equal to one sixth of the range of their triangular distributions respectively (Thilakaratne, et al. 2014).

Figure 6(a) details the probability density function of MSP from MC simulation. It can be observed that the distributions of key parameters have a significant influence on the distribution of MSP. The normal distributions case results in a larger mean (5.46 $/GGE to 6.23 $/GGE) and the distribution of MSP is shifted to right by about one dollar than the triangular distributions case. Both cases show that the probability density functions of MSP are skewed to right a little, which indicates extremely high MSP has a very low probability.

{Insert Figure 6 here}

As shown in Figure 6(b), the empirical cumulative distribution of MSP in triangular distribution case shows that about 31.32% of runs in the analysis have MSP less than $5/gal and 23.62% of the runs will have MSP exceeding 6 $/GGE. On the other hand, in the normal distribution case, only 5.94% of runs in the analysis have MSP less than 5 $/GGE and 58.06% of the runs will have MSP exceeding 6 $/GGE. These results indicate that this pathway is likely to be economically infeasible for a facility capacity of 2000 t d^{-1}. Larger capacity for the bio-oil gasification facility is necessary for this pathway to be economically feasible. Suitable assumptions for distribution of key parameters are essential for uncertainty analysis.

This techno-economic analysis is subject to limitations which suggest future research directions. First, more experimental data is needed to adjust the model parameters to improve the model accuracy. Another limitation for traditional TEA work is lack of economic evaluation based on supply chain (Gnansounou and Dauriat 2010). Thus, practical logistic configuration and constraints should be considered to have a more realistic analysis of commercial scale (Zhang, et al. 2014a), since this pathway could adapt to have decentralized logistic configuration to reduce feedstock costs. It is reasonable to study the supply chain including various decentralized pyrolysis facilities coupled with a central bio-oil gasification facility for a region. Moreover, comparative study of similar pathway considering TEA and logistic configuration could be conducted to assess the economic superiority.

5. Conclusions

In this paper, a detailed process modeling is presented and results of techno-economic analysis and uncertainty analysis of this fast pyrolysis and bio-oil gasification pathway are conducted to assess the economic feasibility. Base on a facility capacity of 2000 t d^{-1}, the results of the TEA study show a capital investment of 510 million dollar and MSP of 5.59 $/GGE. The sensitivity analysis illustrates that MSP is most sensitive to IRR, feedstock cost, and fixed capital cost. The MC simulation results also indicate that a larger facility capacity is necessary to make this pathway economically feasible.

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Figure captions

Figure 1 Generalized process flow diagram for bio-oil gasification pathway

Figure 2 Process flow diagram for bio-oil gasification process (a) and FT synthesis process (b)

Figure 3 Equipment cost and installed cost for each area (a) and annual itemized operating costs (b) ($ Million)

Figure 4 Variation of MSP and facility IRR with facility size

Figure 5 Sensitivity analysis for MSP in 2013 $/GGE

Figure 6 Probability density function (a) and empirical cumulative distribution (b) of MSP from MC simulation