Full Length Article

Pyrolysis of forest residues: An approach to techno-economics for bio-fuel production ☆

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ABSTRACT

The techno-economics for producing liquid fuels from Maine forest residues were determined from a combination of: (1) laboratory experiments at USDA-ARS's Eastern Regional Research Center using hog fuel (a secondary woody residue produced from mill byproducts such as sawdust, bark and shavings) as a feedstock for pyrolysis to establish product yields and composition, and (2) Aspen Plus® process simulation for a feed rate of 2000 dry metric tons per day to estimate energy requirements and equipment sizes.

The simulated plant includes feedstock sizing and drying, pyrolysis, hydrogen production and hydrotreatment of pyrolysis oils. The biomass is converted into bio-oil (61% yield), char (24%) and gases (15%) in the pyrolysis reactor, with an energy demand of 17%. The bio-oil is then hydrotreated to remove oxygen, thereby producing hydrocarbon fuels. The final mass yield of gasoline/diesel hydrocarbons is 16% with a 40% energy yield based on the dry biomass fed, this yield represents a fuel production of 51.9 gallons per dry metric ton of feedstock. A unique aspect of the process simulated herein is that pyrolysis char and gases are used as sources for both thermal energy and hydrogen, greatly decreasing the need to input fossil energy. The total capital investment for a grass-roots plant was estimated to be US$427 million with an annual operational cost of US$154 million. With a 30 year project life, a minimum fuel selling price was determined to be US$6.25 per gallon. The economic concerns are related to high capital costs, high feedstock costs and short hydrotreating catalyst lifetimes.

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1. Introduction

Petroleum and its derivatives have become the most important source of energy for many countries. The price volatility of petroleum feedstock and growing concerns over the environmental impacts associated with the production of petroleum derived fuels have created interest in finding renewable substitutes that can be used without major changes to the current fuel distribution and utilization infrastructure. Fast pyrolysis is an attractive primary process for converting lignocellulosic biomass to bio-fuel because of its simplicity as well as its flexibility in scale-up [1,2]. However, the high oxygen content and other characteristics of the oil produced via biomass fast pyrolysis limits its use in combustion engines. Thus, additional catalytic hydrotreating and hydrocracking of bio-oil are required in order to make it suitable as a transportation fuel. The technical feasibility and economic viability of renewable hydrocarbon fuel production via fast pyrolysis process are analyzed by a number of studies. The estimated total capital investment of these studies range from $287 MM to $700 MM for conversion of 2000 dry MT biomass feedstock per day to hydrocarbon fuel. The sensitivity analyses of these studies show strong impact of capital cost, fuel yield, and feedstock cost on the minimum fuel selling price [3–5].

Forest residues are considered as one of the potential biomass feedstock resources for the bioenergy production. The 'hog fuel', a forest residue, was used for the pyrolysis experiments in this study. Hog fuel is a secondary woody residue produced from mill byproducts such as sawdust, bark and shavings. Currently, most hog fuel is combusted for heat production. As a comparison, hog fuel has a similar ash content to corn stover, between 4% and

10% [6]; making it more akin to herbaceous feedstocks than to clean wood (hybrid poplar wood has just ≅2% ash). Ash percentage is an important factor in pyrolysis because the ash components have catalytic effects which can change the depolymerization mechanisms and the pyrolysis product yield distribution [7]. The total quantity of hog fuel processed annually by the Maine forest products industry is about 613,000 green tons [8], which has an energy content equivalent to almost 2 million barrels of oil.

In this study, fast pyrolysis of hog fuel was carried out in a pilot scale pyrolysis unit at an Agricultural Research Service (ARS) laboratory of U.S. Department of Agriculture (USDA) at Wyndmoor, PA. The experimental results were applied to develop a techno-economic model to assess the economic potential of hydrocarbon biofuel production from hog fuel. A complete process model for the hydrocarbon biofuel production was created by reviewing published literature on biomass drying, bio-oil upgrading, and hydrogen production strategies.

The pyrolysis system that the process model is based on was developed by the USDA together with the University of Pretoria (South Africa). The patent pending design consists of a dual fluidized bed combustion-reduction integrated pyrolysis (CRIPS) unit that efficiently integrates char combustion for heat integration. In the CRIPS, two reaction chambers are connected to convey hot sand from a combustor chamber to the pyrolysis chamber as shown in Fig. 1. Sand and a fraction of the char produced in the pyrolysis reaction are transferred to the combustor by an auger which mixes the sand with more char to combust and therefore transfer heat to the re-circulating sand. This technology has been proven at pilot plant scale and is under development at the USDA - ARS.

In general, limited by the biomass composition, the bio-oil produced has low hydrogen content (from 5 wt% to 7 wt%); this, combined with a high oxygen content (around 32 wt%), results in a fuel with poor energy density compared with hydrocarbon fuels. Upgrading the bio-oil represents a challenge in finding a feasible path that involves proven technologies. The goal of performing an upgrading step is to convert the bio-oil into a physically and chemically stable product with zero oxygen content and about 13% hydrogen by mass, i.e. nearly chemically equivalent to petroleum [9–11]. The resulting mixture of hydrocarbons would therefore, ideally be a "bio-crude" oil that can be refined following the same path used by conventional petroleum refineries.
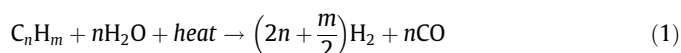
Taking these considerations as a design strategy for the upgrading, Jones et al. [4] has proposed hydrotreating of bio-oil under mild conditions followed by very high temperature hydroprocessing for the complete deoxygenation of bio-oil. It is known that low temperature hydrogenation is helpful to prevent reactor plugging and coke formation during hydrotreating under severe conditions and to reduce the frequency of catalyst replacement or regeneration. Jones et al. [12] presents a 60-day catalyst life as the current state of technology. Considering this, the process modeled herein utilizes an additional reactor for hydrotreating under severe conditions. The additional reactor is used when the other has been loaded with a new catalyst. This strategy increases the capital investment of an important component in the plant but ensures uninterrupted processing. A heavy fraction of hydrotreated bio-oil further undergo catalytic hydrocracking to produce additional hydrocarbon fuel [4]. Jones et al. [4] has reported the hydrogen consumption across all the stages of hydrotreating as 5.8 lb per 100 lb of dry pyrolysis oil.

One limitation in bio-oil upgrading is how to generate the hydrogen. Jones et al. [4] has proposed an in-situ hydrogen generation using steam reforming of natural gas which favorably affects the process economics because of the currently low price of natural gas. However, the energy content of the natural gas used for upgrading in this report, is on the order of 20% of the energy content of the final product, and natural gas is not a renewable resource.

The current study focuses attention on different methods to produce hydrogen without utilizing natural gas or other fossil sources. Wright et al. [3] evaluates the economics of producing hydrogen from bio-oil reforming, the strategy is not always economical because the main product yield and revenue are utilized. Alternatively, by-products from the subprocess are potential hydrogen sources. Yan et al. [13] describes the potential of using pyrolysis char to generate a hydrogen rich gas, this strategy takes advantage of the higher char yield from pyrolysis of forest residue. The path to generate hydrogen from char consists of steam gasification of char under certain conditions that maximize the carbon conversion and dry gas yield. These conditions include a steam flow rate of 0.165 g/min per gram of char at 850 °C. The composition of the gas product on a dry basis and it properties are presented in Tables 1 and 2.

Table 2 shows a H_2 yield potential based on additional water-gas shift. The hydrogen producing reactions are:

$$C_nH_m + nH_2O + heat \rightarrow \left(2n + \frac{m}{2}\right)H_2 + nCO \tag{1}$$

$$CO + H_2O \rightarrow H_2 + CO_2 + heat \tag{2}$$

Steam reforming of syngas is additionally used because the hydrogen requirements cannot be met by char gasification alone. This process is similar to that used in the current hydrogen production industry, except that instead of natural gas, gases are sourced from gasification of char and non-condensable pyrolysis gases. Mann [14] presented a techno-economic analysis of hydrogen production by steam reforming of syngas. A primary reformer uses catalyst-filled tubes immersed in a gas furnace. It converts methane and higher hydrocarbons to CO and hydrogen (reaction (1)). Two further reactors, operating at high temperature and low temperature, convert the remaining CO via reaction (2). Then a
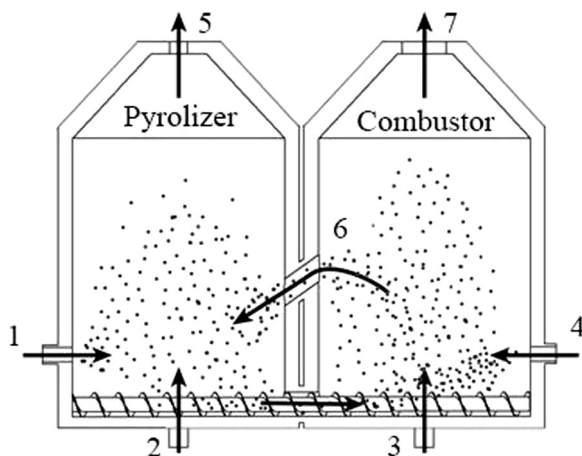

Fig. 1. Combustion reduction integrated pyrolysis system (CRIPS). 1 biomass feeding, 2 fluidization gas, 3 air, 4 char, 5 pyrolysis gas, 6 hot Sand, 7 flue gas.

Table 1
Gas composition of steam gasification of char.

Gas	V%/dry basis
H_2	52.41
CO	14.03
CO_2	27.60
CH_4	1.74
C_2H_4	4.01

Table 2
Gas proprieties from char gasification.

Propriety	Value	Unit
H_2/CO	3.74	(mol/mol)
LHV	8.3	(MJ/Nm^3)
H_2 yield	57.07	(mol/kg)
H_2 yield potential	87.83	(mol/kg)
Dry gas yield	2.44	(Nm^3/kg)
C-conversion	95.78	(wt%)

pressure swing adsorption (PSA) system purifies the hydrogen and recovers some of the hydrogen not consumed in the two bio-oil upgrading steps.

We developed a model using Aspen Plus® to calculate mass and energy balance on reactors, heat exchangers, separators and pressure changers units for the pyrolysis hog fuel and upgrading of the produced bio-oil of using in situ generated hydrogen to produce hydrocarbon bio-fuels. This is the baseline to understand key factors in the techno-economic assessment. Mass flows are used for equipment sizing and capital cost estimations. Potential heat integration was considered before determining plant energy requirements.

2. Methods

2.1. Feedstock

Hog fuel comes in varied sizes and shapes, with chips being typically 5 mm thick and 30 mm long. The hog fuel was ground in a wood grinder using a screen (1.8 mm) and a dust collector, and dried to reduce moisture content from 44% to 5.9%. Table 3 provides an elemental analysis of the dried and ground particles.

The higher heating value (HHV) of the hog fuel was calculated to be 18.3 ± 0.9 MJ/kg using a correlation of C, H, O content presented by Sheng and Azevedo [15]. The moisture content of the hog fuel fed into the pyrolizer was 5.9%.

2.2. Pyrolysis experiments

Fig. 2 shows a schematic of the laboratory-scale pyrolysis unit at the USDA-ARS laboratory, which was used for the pyrolysis experiments. The ground biomass was fed from a 30 L hopper into the reactor at rate of 1 kg/h, using an injection auger. The pyrolizer is a fluidized bed reactor consisting of a vertical stainless steel vessel, 7.8 cm in diameter and 52 cm in height, which operates at about 500 °C. The fluidized bed medium is quartz sand which transfers heat into the biomass particles. Preheated nitrogen was used as the fluidization media at flow rate of 105 L/min. The pyrolysis products were sent to a cyclone where char was separated and collected in a glass container. Four condensers were used to separate the condensable fraction of the solid-free gas/vapor as well as to collect a fraction of the bio-oil. After the condensers, an electrostatic precipitator was employed to collect the remaining fraction of the bio-oil, which is more homogeneous and has lower water
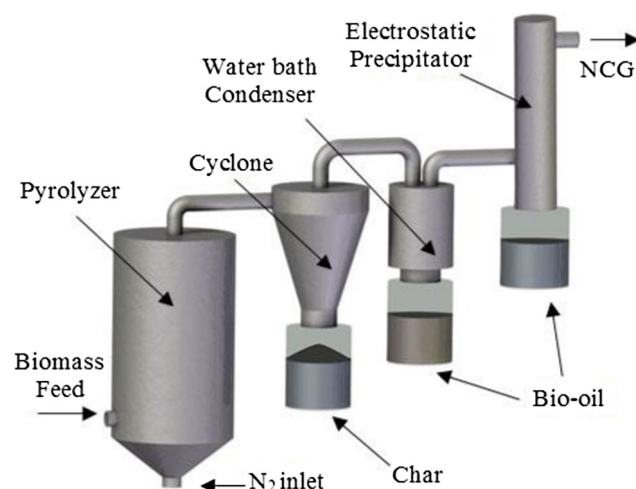
Table 3
Analysis of hog fuel biomass.

Component	Weight % (dry basis)
C	45.6
H	5.59
N	0.75
S	0.00
O	44.1
Ash	3.96

Fig. 2. Fast pyrolysis experimental setup. Bio-oil from condensers has higher water content than from the Electrostatic precipitator. (NCG = non-condensable gases).

content. Tar formation can occur inside the pipes that connect the reactor, the cyclone, and the condensers, if there is insufficient thermal insulation.

2.3. Model of pyrolysis oil in Aspen Plus®

The bio-oil was modeled using a mixture of compounds that are known to be present in both the bio-oil and in the Aspen Plus® databases. The mass fraction of each compound was determined by matching the elemental analysis and HHV of the bio-oil to satisfy the atom and energy balances of the unit operations. Table 4 presents the selected model compounds and their mass fractions.

2.4. Process topology design

Fig. 3 presents an overview of the process topology for the conversion of hog fuel into biofuel. The bio-energy plant used in this model processes 2000 dry MT/day of hog fuel, which is equivalent to amount produced by a small Maine pulp mill. The process has 5 steps: biomass pretreatment, pyrolysis, hydrogen production, bio-oil upgrading, and energy generation. These sub-processes are explained in detail below.

2.4.1. Biomass pretreatment
Collected hog fuel has a moisture content of 44% by mass, but covered storage could dry it to around 30%. Hog fuel chips were ground to 2 mm using a high efficiency grinder with an energy requirement of 14 kWh/green MT. The dryer consists of a super-heater steam dryer which is considered as a fire safe alternative for drying biomass compared to using hot flue gas. To obtain 10%

Table 4
Aspen Plus® model compounds and mass fractions to represent hog fuel Bio-oil.

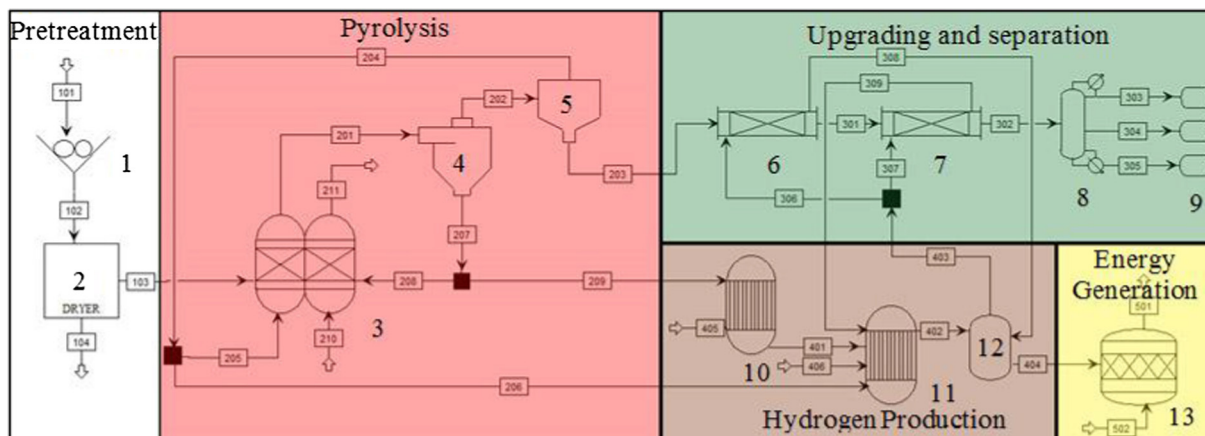| Compound	Weight fraction
Acetic acid	1.45×10^{-1}
Guaiacol	2.60×10^{-2}
Syringol	4.33×10^{-2}
Formic acid	1.01×10^{-1}
Propyl benzoate	2.71×10^{-1}
Phenol	2.78×10^{-2}
Toluene	6.29×10^{-2}
Furfural	3.02×10^{-1}
Benzene	2.15×10^{-2}

Fig. 3. Overall topology. 1 grinder, 2 dryer, 3 CRIPS, 4 cyclone, 5 condenser, 6 hydrotreating reactors, 7 hydrocracker, 8 distillation tower, 9 storage tanks, 10 char gasifier, 11 steam reformer, 12 PSA, 13 gas combustor.

moisture content the energy requirement is approximately 136 kWh/green MT of hog fuel at 30% moisture.

2.4.2. Pyrolysis

After pretreatment, the biomass was fed to the pyrolysis chamber of the CRIPS, where it mixes with hot fluidized sand and reacts to form pyrolysis gases and char. The char was separated by a cyclone and a fraction was fed to the combustion chamber where the recirculating sand was heated to a temperature of 680 °C. The remaining char was combusted for process energy. The bio-oil was condensed and non-condensable gases were used for pyrolysis fluidization.

2.4.3. Bio-oil upgrading

Upgrading consists of two separate reactions: hydrotreating and hydrocracking [4]. A near complete removal of oxygen content of pyrolysis oil can be accomplished using three catalytic stages of hydrotreating [4]. For the first two reactor stages, ruthenium based catalysts were employed [4]. The first catalytic reactor was operated at mild conditions of 81 atm and 140 °C, and the second catalytic reactor was operated at 136 atm and 252 °C [4]. A molybdenum based catalyst was used in the final hydrotreating stage, which operates at severe conditions of 136 atm and 400 °C [4].

After hydrotreting, the oil was processed over a conventional sulfided hydrocracking catalyst for the complete deoxygenation of hydrotreated bio-oil, as well as to produce additional fuel by creacking the heavy fractions of hydrotreated bio-oil [4]. Space velocities of hydrotreting and hydrocracking catalysts were obtained from Jones et al. [12]. For the modeling purpose, the life of ruthenium and molybdenum based catalysts were assumed to be one year and 60 days, respectively [12].

This upgrading process incorporates an energy integration design that reclaims heat from the exothermic hydrotreatment reactions and from the energy generation and uses it to preheat each incoming flow as shown in Fig. 4. The gases from the mild hydrotreating step (stream 308) are sent to the PSA system to recover part of the unused hydrogen. The gases from the hydrocracking step (stream 309) are rich in methane, hence they are sent to the hydrogen generation step.

2.4.4. Hydrogen production

Char was gasified to generate syngas which was then used to produce hydrogen. The process yields were scaled to an industrial level from a technology assessment by Worley and Yale [16] for

gasification of biomass. This technology uses a direct-heated bubbling fluidized bed gasifier operating at 850 °C and 1 atm, which burns fuel gas from the PSA off-gases to perform the gasification reactions. Syngas obtained was then treated in a series of reactions and purification stages to obtain pure hydrogen.

A primary reformer followed by two subsequent water-gas shift reactors were used to produce hydrogen from syngas. The primary reformer was operated at 850 °C and 35.5 atm. In the primary reformer, the reactants and products were circulated inside tubes filled with a commercial nickel-based catalyst. The heat necessary for this endothermic reaction was provided from the off-gas of the PSA unit. A syngas compressor was utilized to provide pressure for this reaction. Due to the pressure drop across the primary reformer, subsequent water-gas shift reactors were operated at 27.5 atm and 24.5 atm, respectively. In the primary reformer 60 mol% of the CH_4 and 22 mol% of the CO were converted. In two water-gas shift reactors operating at temperatures of 370 °C and 200 °C, remaining CO was converted to a mixture of 64% H_2, 32% CO_2, 3% CH_4 and 1% CH_4.

2.4.5. Energy generation

Fuel gases from the PSA off-gas were combusted to generate heat. The composition of this fuel gas by volume was H_2 30%, CO_2 57%, H_2O 6%, CH_4 5% and C_2H_4 2% with a lower heating value (LHV) of 6.5 MJ/Nm^3. The energy produced has several applications within the overall process. It was used to heat the partially upgraded oil from the first step of the hydrotreatment to 405 °C, the bio-oil to 340 °C and the recirculating steam in the super-heated steam dryer.

2.5. Baseline economics

We used the module costing technique to estimate the capital cost or fixed capital investment of the plant. This technique uses the purchased cost of the equipment at base condition, which is using common material of fabrication and operating near ambient pressure, and modifies it considering the equipment type, the specific system pressure and specific materials of construction. Capital costs were estimated by:

$$1.18 \cdot \Sigma BMC + 0.5 \cdot \Sigma BMC_0$$

where BMC and BMC_0 are the bare module cost at operating (the installed cost) and the base condition respectively, for all the equipment in the plant. For sizing and equipment costing of the pyrolysis and upgrading processes, related techno-economic studies were used as Refs. [4,5]. A list of equipment sizing and costing procedure
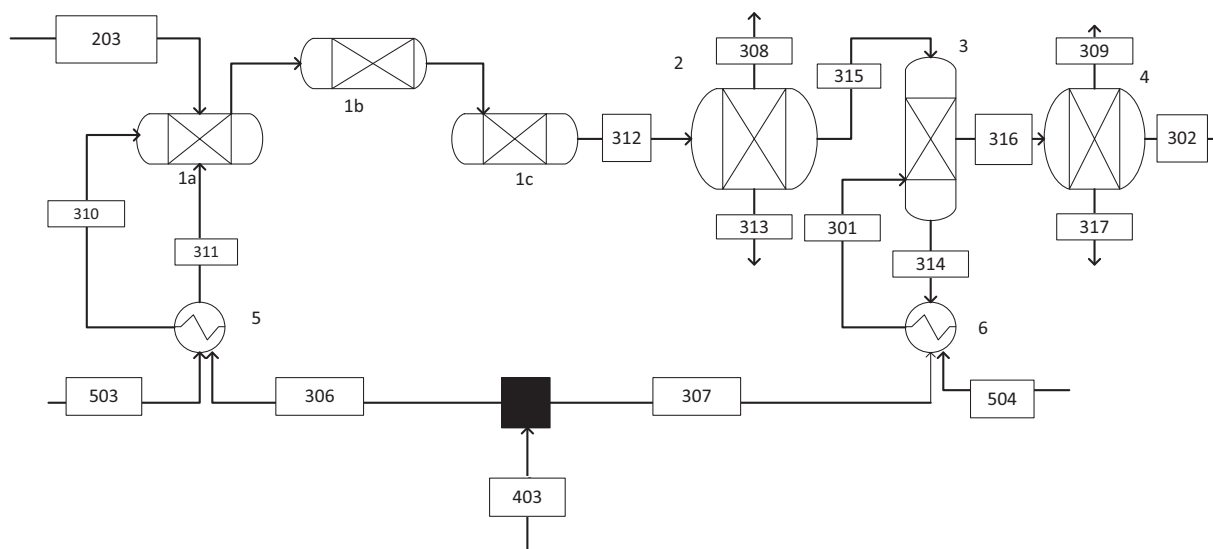
Fig. 4. Upgrading process and heat integration strategy. 1 three-stage hydrotreater (mild upgrading), 2 separator, 3 hydrocracking reactor, 4 separator, 5 and 6 heat exchangers.

is included in Electronic Supplementary Material. The profitability of the project was calculated under the economic assumptions presented in Table 5.

3. Results and discussion

3.1. Pyrolysis of forest residues

Pyrolysis of the hog fuel produced 61 wt% bio-oil (including water), 15% gas and 24% char as shown in Fig. 5. Also shown is the energy distribution in the products, on the basis of their higher heating values weighted by each mass fraction. Carbon is distributed in the products with a mass balance closure of greater than 95%. Ash content in the hog fuel is used as a key component to obtain an accurate mass balance, assuming that all the ash is separated with the char. The chemical composition and heating values of the products are presented in Tables 6 and 7.

3.2. Energy and mass balance

The energy and mass yields for the process are summarized in Fig. 6 and Tables 8–10. Energy yields and demands are calculated using the higher heating values of the streams. This model assumes 2000 dry metric tons per day of biomass which has the potential to produce 421 MW. The overall mass yield is 16% and the energy yield is 40% (stream 4). The energy demand of the drying process (stream 8) is 5.6%.

About 50% of the pyrolysis char is combusted to supply energy for the pyrolysis process; this energy can be divided into heat to
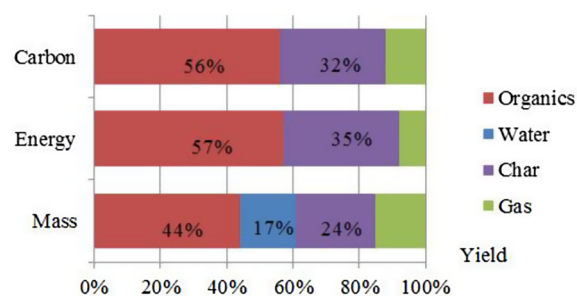

Fig. 5. Fast pyrolysis mass, energy and carbon yields.

Table 6
Elemental analysis and higher heating value (dry, ash-free basis) of pyrolysis products.

Element	Bio-oil Organics wt%	Char wt%
C	61.2	63.4
H	6.00	2.53
N	0.60	0.97
O	32.2	19.4
Ash	0	13.6
HHV MJ/kg	24.9	31.2

Table 7
Chemical composition and higher heating value of non-condensable gases.

NCG	Volume %
CO_2	30.6
CO	41.6
CH_4	14.0
H_2	13.8
HHV MJ/kg	9.77

increase the temperature of the wet biomass from 25 °C to 500 °C corresponding to 1.2 MJ/kg and the energy for the pyrolysis reaction corresponding to 0.1 MJ/kg. The upgrading process requires 3.3% of the energy generated (stream 13), from the exothermic hydrotreating reactions. Half of the hydrogen gas (stream 11) is consumed by the upgrading and the remainder is returned via stream 12 with the mild upgrading gas and cracking off-gas to be recovered in the PSA unit.

Table 5
Economic parameters.

Plant size	2000 DMTPD
Annual interest rate	10%
MACRS depreciation	7 years
Taxation rate	40%
Stream factor	90%
Cost of land	6% of Purchased Equipment Cost
Salvage value	10% of FCI_L
Project life	30 years
Construction period	2.5 years
Working capital	5% of Total Capital Investment

Fig. 6. Energy and mass flows in the process.

Table 8
Mass and energy inputs in the process.

Inputs	1	9	14	19	Totals
	Biomass/water	Air	Water	Air	
Mass [MT/d]	2000/858	1853	492	1514	6717
Energy [MW]	421/0	0	0	0	421

Table 9
Mass and energy outputs in the process.

Outputs	4	5	6	7	17	18	18	Totals
	Fuel	Storage/Vapor	Flue gas/Ash/Storage/Cooling	Storage/Cooling /Water	Flue gas	Char/Ash	Coke formation	
Mass [t/d]	328	0/636	2070/42 0/0	0/0/852	2506	32/37.2	214	6717
Energy [MW]	168	4.4/19	3.0 /0 5.4 /62.6	14/38/0	14.4	10.5	81.4	421

Table 10
Internal mass and energy flows.

Internals	2	3	8	10	11	12	13	15	16
	Biomass/Water	Bio-oil/Water	Drying	NCG Char/Ash	H_2	HDO/Cracking gases	Energy to Upgrading	Energy to H_2 prod.	PSA Off gas
Mass [t/d]	2000/222	870/556	0	310/192	83.2	60.4/52	0	0	990
Energy [MW]	422/0	250/0	23.4	35/69.2/0	137	44.8/50.0	14	12.8	64

Coke formation refers to waste carbon formation on the catalyst surface inside the upgrading reactors. The energy content of this coke represents 19% of the biomass energy (considering a HHV of 32.8 MJ/kg for coke). This energy is potentially recoverable by an appropriate reactor design.

The energy and mass yields for hog fuel pyrolysis and upgrading are presented in Table 11. A comparison is made between the current study and the 2015 State of Technology [12] which used yields for clean sawdust and produced hydrogen using natural gas [12]. The higher heating value of forest residues (18 MJ/kg) was used as the starting point for the conversion process. The upgrading increases the energy density to 45 MJ/kg. Note that the current process is approaching the upper limit of process yield without the use of fossil energy, a concept discussed in [4].

The main characteristic of hog fuel pyrolysis as compared with other biomass pyrolysis is its low yield of organic liquid. For example, the yield is about 27% lower than yield reported for corn stover or aspen poplar wood chips. The lower bio-oil yield is balanced with a higher char yield from the pyrolysis reaction. This characteristic can be taken as an advantage to create a self-sustained process due to the possibility of utilizing char to produce a hydrogen-rich gas used in upgrading via hydrotreatment.

3.3. Economics

Table 12 shows the sum of bare module cost of the equipment for each sub-process of the plant. Including contingency, fees and

Table 11
Mass and energy yields.

Product from:	Mass yield		Energy Yield	
	Present study	Jones et al. 2015	Present study	Jones et al. 2015
Pyrolysis (25 MJ/kg)	44	62	57	70
Pyrolysis + Upgrading (45 MJ/kg)	16	26	40	64

auxiliary facilities costs give the total grass root cost for average 2015 chemical engineering plant cost index (CEPCI = 556.8).

The capital investment of the plant is highly affected by the upgrading process followed by hydrogen production and pyrolysis processes. The production of bio-oil via pyrolysis of biomass is currently at a commercial level, however the hydrogen production and the upgrading strategies presented are at lab-scale development. This means that the capital investment of these sub-processes can be expected to decrease over time. Table 13 summarizes the operating costs of the process. Cost per dry metric ton of forest residues is based on estimations presented above. Annual catalyst cost includes hydrotreating, hydrocracking and reforming. Waste removal cost corresponds to water treatment and ash disposal. Electricity cost corresponds principally to compressors.

The production of hydrocarbons from hog fuel, using the technology presented in this study, gives an annual production of 34.1 million of gallons. A selling price of $6.25 per gallon is necessary in order to have a zero net present value, over the project life. These results plus manufacturing costs associated with the process are presented in Tables 14 and 15.

The MSP sensitivity to economic assumptions is measured for a ±33.3% change in the total capital investment, in the hydrogen plant installed cost, in the CRIPS reactor installed cost, in the upgrading capital cost, and in the annual costs associated with the replacement of all process catalysts. In this study, catalyst life of 60 days is assumed for the molybdenum based catalyst. It has been reported that the target catalyst life of at least one year is necessary for commercial relevance [12]. There is a possibility of shortfall in catalyst performance when it is used at commercial scale production as the feedstock impurities might poison the catalyst and reduce the catalyst life achieved at pilot scale. Thus, in this study we measured variation of the MSP for an increase in the catalyst life to one year and for decrease in the catalyst life to 30 days. Fig. 7 presents the results of the sensitivity analysis.

Total capital investment of hydrocarbon fuel production is the largest economic effect. The next large contributors are feedstock cost and the catalyst life. A ±33.3% change in the total capital investment results in a 17% variation in the MSP of hydrocarbon fuel. Feedstock price will have to compete with a price that represents the current usage of hog fuel as an energy source in pulp mills. Possible variations on its price modify the final fuel price in a range between $5.75 and $6.93. Increasing the catalyst life

Table 12
Summary of capital cost and fees.

Sub-process	Value [$ MM]	Contribution [%]
Pretreatment	13	4.5
Pyrolysis	67	23
Energy generation	7.4	2.6
Upgrading and Separation	121	43
Hydrogen production	75	26
Storage & water cooling	2.4	0.9
Total installed equipment	285.8	100
Contingency	15% of BMC	
Fees	3% of BMC	
Auxiliary facilities	50% of BMC_0	
Total grass root cost	US$427 MM	

Table 13
Summary of operating cost.

	Cost	Unit
Raw material		
Hog fuel	$69	dry t
Catalyst		
CoMo	$30[a]	kg
Ni/ZSM-5	$5[b]	kg
Ru/C	$60[c]	kg
Utilities[d]		
Cooling water	$0.27	t
Electricity	$0.06	kWh
Ash disposal	$53	t
Waste water	$0.34	m^3
Operating labor	$26.9[e]	h

[a] Ref. [14].
[b] http://www.alibaba.com.
[c] Ref. [4].
[d] Ref. [16]. Reference adjusted to CEPCI 2015.
[e] Bureau of labor and statistics.

Table 14
Annual revenue and expenses.

Annual revenue and expenses	Value (MM$)
Revenue from sales	**213**
Raw materials	45.3
Catalyst	22
Utilities	25.2
Waste treatment	1.66
Operating labor	1.64
Maintenance and overheads	32.2
Distribution & selling	7.48
Others	18.2
Total expenses (cost of manufacturing)	**154**

Table 15
Minimum selling price (MSP) breakdown.

	$/gal of product	% Contribution to MSP
Feedstock	1.33	21.3
Maintenance and overheads	0.94	15
Utilities	0.74	11.8
Average IRR	0.95	15.3
Capital depreciation	0.42	6.7
Average income tax	0.35	5.6
Catalyst	0.65	10.3
Distribution & selling	0.23	3.6
Operating labor	0.05	0.8
Others	0.54	8.6
Waste treatment	0.05	0.8
MSP	6.25	100

to one year will result in a 5% reduction in the fuel MSP. On the other hand reducing the catalyst life to 30 days causes the fuel MSP to rise by 8%. Such a rise is a result of increase of number of catalyst regeneration cycles when the catalyst life is 30 days - the number of catalyst regeneration cycles per year is 5 when the catalyst life is 60 days and this number is increased to 11 for the catalyst life of 30 days. In this analysis, we assumed that the
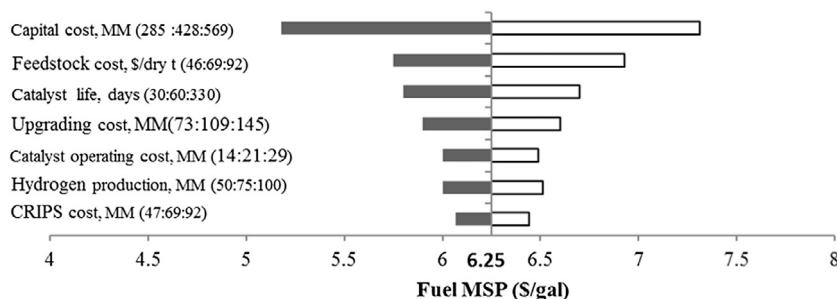
Fig. 7. Economic Sensitivities (±33.3% except catalyst life).

spent catalyst is treated at a hazardous waste facility, and the transportation and the tipping fee is assumed to be $56 per ton of spent catalyst.

The annual costs associated with the replacement of all process catalysts and capital costs of hydrogen production and bio-oil upgrading processes show significant impact on the hydrocarbon fuel MSP. In this study, we assumed that ruthenium catalyst used in the bio-oil hydrotreating can be recovered and regenerated. The annual operating cost for regenerating the ruthenium catalyst has been assumed to be 5% of the initial purchase cost according to personal communications with Chevron Phillips. The sensitivity analysis shows that the impact of CRIPS reactor installed cost on fuel MSP is less significant than the other above mentioned costs.

4. Conclusions

This work estimated the MSP for bio-fuel produced from forest residues in a 2000 MT/d plant. The conversion process is carried out by pyrolysis of ground biomass and subsequent two-stage hydrotreating of the pyrolysis oil. Experimental results of forest residue pyrolysis yield 24% char by weight; this material stream can be used to produce hydrogen for the hydrotreatment without requiring external natural gas. Similarly, non-condensable gases are used as energy source for the process. The bio-fuel yields were estimated around 16% by mass and 40% by energy. This study demonstrates that the lower oil yields for forest residues, relative to those achieved by pyrolysis of clean, bark-free wood, and can be compensated for by use of byproducts.

A strategy of three stages of catalytic hydrotreating followed by catalytic hydrocracking was used to produce hydrocarbon fuel from pyrolysis oil. Pyrolysis char was gasified to generate syngas employing a technology assessment from Worley and Yale 2012 from NREL [16] for gasification of biomass. Syngas obtained was then used in a series of gas reforming and water shifting reactions and purification to obtain hydrogen.

The total capital investment for the plant is US$427 MM with an annual cost of manufacturing of US$154 MM. Assuming a 30 year project life, the minimum fuel selling price would be US$6.25 per gallon. Economic sensitivities show that the selling price is most sensitive to capital cost followed by biomass cost and the catalyst life.

Acknowledgments

Appendix A. Techno-economic data

Supplementary data associated with this article can be found, in the online version, at http://dx.doi.org/10.1016/j.fuel.2016.12.063.

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