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Renewable hydrocarbon fuels from hydrothermal liquefaction: A techno-economic analysis

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Abstract: This study demonstrates the economic feasibility of producing renewable transportation drop-in fuels from lignocellulosic biomass through hydrothermal liquefaction and upgrading. An Aspen Plus[®] process model is developed based on extensive experimental data to document a techno-economic assessment of a hydrothermal liquefaction process scheme. Based on a 1000 tonnes organic matter per day plant size capacity, three different scenarios are analyzed to identify key economic parameters and minimum fuel selling prices (MFSP). Scenario I, the baseline scenario, is based on wood-glycerol co-liquefaction, followed by thermal cracking and hydroprocessing. Results show that a minimum fuel selling price (MFSP) of 1.14 \$ per liter of gasoline equivalent (LGE) can be obtained. In Scenario II, only wood is used as feedstock, which reduces the MFSP to 0.82 \$/LGE. Scenario III is also based on a pure wood feedstock, but investigates a full saturation situation (a maximum hydrogen consumption scenario), resulting in a slightly higher MFSP of 0.94 \$/LGE. A sensitivity analysis is performed identifying biocrude yield, hydrogen, and feedstock prices as key cost factors affecting the MFSP. In conclusion, the study shows that renewable fuels, via HTL and upgrading, can be highly cost competitive to other alternative fuel processes. © 2017 The Authors. *Biofuels, Bioproducts and Biorefining* published by Society of Chemical Industry and John Wiley & Sons, Ltd

Keywords: biomass; biofuels; HTL; drop-in fuels

Introduction

Biomass is the most important renewable carbon source with the ability to replace current fossil transportation fuels. One way of converting highly diverse, low-cost biomass into drop-in liquid fuels is by hydrothermal liquefaction (HTL). In HTL, biomass is mixed with water and processed at temperatures between 250 and 450 °C and pressures between 5 and 35 MPa.^{1–3}

The severe conditions improve water solvent properties, enhancing the production of liquid biocrude, a water phase with dissolved organics and a gas phase.⁴ One of the key benefits of the HTL process is the low oxygen content of the biocrude, where around 85% of the original inherent oxygen of the biomass is extracted mainly as CO₂, water, or as water soluble organics.⁵ The biocrude from the HTL process can be further upgraded to equivalent hydrocarbons via, for example, hydrotreatment.⁶

The first conceptual high-pressure HTL process was developed in the 1970s at the Pittsburgh Energy Research Center (PERC) and demonstrated at the Albany Biomass Liquefaction Experimental Facility, Oregon, with a processing capacity of 100 kg/h.⁷ Since then, many different organic materials have been processed, from lignocellulosic matter to algae, at several continuous processing facilities.^{3,8–11}

Even though HTL has been proven technically viable, the economic perspectives of the process have only been superficially addressed. The Pacific Northwest National Laboratory (PNNL) established an economic frame of reference for the viability of fast pyrolysis and hydrothermal liquefaction followed by upgrading. Fuel costs of 0.82 and 0.53 \$/liter of gasoline equivalent (LGE) were estimated for fast pyrolysis and HTL, respectively. The study considered a large-scale production facility, processing 2000 metric tons per day of dry wood (474 MW).⁶ The assessment was evaluated based on an 8% dry matter content feedstock, which is rather low compared to other experimental studies that have demonstrated the processing of feedstock having dry matter contents much higher.^{10,12,13} In another study by Zhu *et al.*, based on a similar process but in a current 'state-of-technology' scenario, the cost of producing renewable fuels from woody biomass (15 wt.% dry biomass content) using HTL was estimated at 0.74 \$/LGE.⁵ In the same study a 'goal case' scenario was evaluated, projecting a minimum fuel selling price (MFSP) of 0.67 \$/LGE.

In a more recent study, de Jong *et al.* compared the production costs of several renewable processes for jet fuel production from various feedstock. From the comparison, HTL emerged as a best performing technology with estimated jet fuel production costs in the range of 21–29 €/GJ (~0.79–1.09 \$/L).¹⁴ Other techno-economic studies carried out by PNNL include the conversion of algae and municipal waste water.^{8,15} For the two feedstocks, base case scenarios were calculated yielding a MFSP of 1.19 and 1.29 \$/LGE for algae and municipal waste water, respectively.

Recently, Pedersen *et al.* demonstrated successful, continuous bench-scale operation of an aspen wood-glycerol co-feedstock in which an aqueous phase recirculation procedure was incorporated.¹⁰ In a later study, the biocrude from the experimental study was fully characterized, showing that a significant fraction (~70 %) of the total biocrude can be turned into gasoline equivalents through (i) biocrude distillation, (ii) thermal cracking of the heavy fraction, and finally (iii) a co-hydroprocessing the lighter distillation fraction and obtained cracking product.¹⁶

This study uses the experimental data of Pedersen *et al.*^{10,16,17} to establish the technoeconomic basis for a conceptual industrial scale HTL plant using wood,

alone or co-liquefied with glycerol, to produce gasoline equivalents. The assessment intends to highlight potential improvements of the system and to perform economic evaluation of the key parameters affecting production costs. The final objective is to estimate the MFSP of renewable gasoline equivalents derived from HTL on wood. The following three scenarios are investigated:

1. Scenario I: A mixture of glycerol and wood is co-fed into the HTL process. The biocrude is distilled into a volatile and a non-volatile fraction. The non-volatile fraction is thermally cracked, and the cracking products are co-hydroprocessed with the volatile fraction into gasoline equivalents.
2. Scenario II: Similar to Scenario I, but the feedstock is assumed to consist only of wood.
3. Scenario III: Like Scenario II, but in this scenario a complete hydrogenation of all compounds is assumed to evaluate a *maximum hydrogen consumption* scenario.

Materials and methods

This study is based on the development of an Aspen Plus[®] model of a conceptual industrial scale plant with a processing capacity of 1000 tonnes of organic matter per day. The input parameters for the Aspen Plus[®] model are based on previous experimental work, where a 50/50 wood-glycerol co-feedstock was processed continuously, producing more than 40 L of biocrude,^{10,17} which was extensively characterized in a later work by Pedersen *et al.*¹⁶ The co-feedstock was processed at 400°C, 300 bar, and with a mass flow rate of approximately 14 kg/hr. Biocrude yields in the order of 20–30 % were obtained. Although carboxymethyl cellulose (CMC) was used in the experimental runs for stabilizing the feed slurries, its presence has been excluded from the economic assessment since it is unnecessary in large scale operation. The departure point for the biocrude characterization was a fractional distillation yielding a mass fraction of 49% volatile compounds (boiling point below 350°C (atmospheric equivalent temperature)), and a 51% mass fraction of non-volatile compounds (distillation residue). 150 chemical compounds were identified in the volatile fractions. The volatile fraction was then hydrotreated, reducing the oxygen content from 14.5% to 4.6%. The hydrotreated product was characterized and some 30 compounds identified. More information about the catalyst, reaction parameters, identified compounds etc. can be found in Pedersen *et al.*¹⁶ The residue was characterized by pyrolysis GCxGC-MS, identifying more than 40 chemical compounds in the thermally cracked

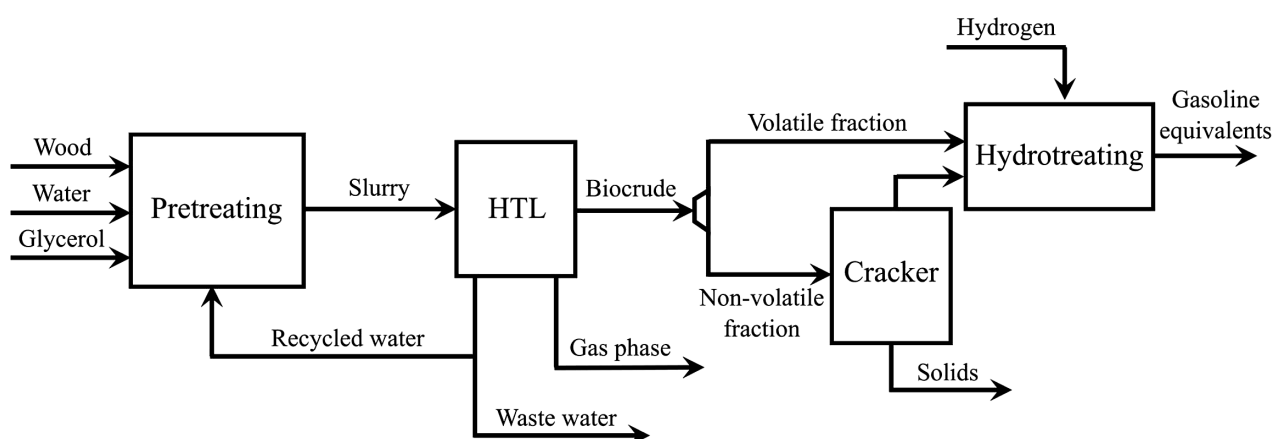


Figure 1. Overall HTL system diagram including pretreatment, upgrading, and distillation.

product.¹⁶ The chemical compounds identified are used to establish reactor models (HTL reactor, thermal cracker, and hydrotreater) in the present study.

Figure 1 shows the main process scheme considered in this work. Starting from the pre-treatment, wood is grinded and mixed with glycerol and water, constituting the initial slurry. This mixture is pressurized to 300 bar, then heated to 400°C, prior to entering the HTL reactor. Within the reactor, the organic macromolecules are decomposed into smaller compounds. Afterwards, the product is depressurized, cooled down and separated into three phases: water phase, gas, and biocrude.

The gas phase, which mainly consists of CO₂,¹⁶ can be extracted for further utilization either as a combustible source or for example, for CO₂ or H₂ recovery. In the current study, the gas phase is used as a combustible source for process energy. The gas composition can be seen in Table 2.

The water phase is rich in organics, and in the case of lignocellulose processing, the organic content of the water phase represents a significant share of the organic output (> 50%). In this study, it is assumed that 90% of the water phase is recovered by recirculating it back to the system input. Other means of water phase utilization has been proposed, such as anaerobic digestion for biogas production.⁶ The remaining 10% of the water phase is lost to a water phase cleaning process. After the HTL core process, the biocrude is separated into a volatile and a non-volatile fraction (residue) by distillation according to Pedersen *et al.*¹⁶ The residue is then thermally cracked and mixed with the volatile stream prior to co-hydrogenation in a hydrotreater. The obtained liquids are referred to as *gasoline equivalents*.

Modeling of core processes

The total system is divided into four core blocks, which are explained in the following.

Table 1. Ultimate and proximate analysis of aspen wood and distillation residue. Obtained from Pedersen *et al.*^{10,17}

Proximate	[%]	Ultimate	[%]
<i>Aspen wood</i>			
Moisture	0	C	50.49
FC	20.61	H	6.19
VM	77.04	O	43.32
ASH	2.35	HHV	18.8 MJ/kg
<i>Residue</i>			
Moisture	0	C	50.49
FC	42.06	H	6.19
VM	57.00	O	43.32
ASH	0.94	HHV	35.2 MJ/kg

Feedstock properties and pre-treatment

The system is fed with a mixture of wood, water and glycerol as shown in Table 2. For modeling purposes, the aspen wood (dry) is defined as a non-conventional solid in Aspen Plus®, for which its properties (e.g. enthalpy and density) are calculated based on the proximate and ultimate analysis shown in Table 1.

HTL reactor

The HTL reactor operating conditions are set to 400°C and 300 bar, obtaining a 30% biocrude yield. The HTL reactor is modelled as a yield-type reactor (RYield), where the output biocrude is modeled using 81 different chemical compounds, representing the whole biocrude, adopted from experimental data.¹⁶ Although the overall mass balance between the inlet and the outlet of the reactor is maintained, this procedure does not automatically conserve the atomic distribution across the reactor. Therefore, the individual

mass fractions of the 81 compounds have been manually adjusted to minimize atomic discrepancies across the reactor to obtain better energy consumption predictions.

The biocrude is split into a volatile and a non-volatile fraction, based on a maximum, atmospheric equivalent temperature of 350°C.¹⁶ The weight distribution of the volatile and non-volatile fractions is 43.6% and 51.8%, respectively. The remaining 4.6% corresponds to non-condensable compounds under the given experimental conditions and experimental mismatch. The splitter is a 'dummy component' used to imitate distillation, but without any heating or cooling requirements.

The gas phase represents 13 wt.% of the organic inlet. The remaining 57 wt.% of the organics leave the HTL reactor as dissolved organics in the aqueous phase; a phenomenon that should be studied further, as it represents a significant percentage of the total organic inlet. Regarding the water content, it has been assumed that it remains constant within the HTL reactor, with 90 % recirculation rate. The remaining 10% is discharged from the system as waste water.

Thermal cracking

The thermal cracker operation conditions are set to a pressure of 4 bar and a temperature of 600°C, with a cracking efficiency of 70 %.¹⁶ Like for the aspen wood, the input non-volatile fraction is characterized as a non-conventional solid, for which its properties are calculated based on the proximate and ultimate analysis also shown in Table 1. The cracking products are defined as conventional compounds based on the identified compounds of the cracked product. In total, the cracking product is defined by 32 chemical compounds from the Aspen Plus[®] library based on data from Pedersen *et al.*¹⁶ The remaining solids are discharged from the system and are no further investigated.

Hydrotreating

The conditions for the hydrotreater are set to 360°C and 77.5 bar,¹⁶ constituting an exothermic reaction. It is assumed that the hydrotreating proceeds with a 100% carbon efficiency with only water as a by-product by means of the hydrodeoxygenation reaction shown in Eqn (1):



The final product has been modeled by 23 compounds from the Aspen Plus[®] library based on data from Pedersen *et al.*¹⁶ The liquid yield of the hydrogenation is then 80.25%, mainly hydrocarbons and residual oxygenates, and 19.75% water on a mass basis. From the performed atomic balance, the gasoline equivalents in Scenarios I and

II contains 3.3% oxygen (0 % oxygen in Scenario III), with a hydrogen consumption of 0.036 g H₂ consumption for Scenarios I and II, 0.0482 g H₂/g biocrude in Scenario III.

The SRK-KD equation of state is used, which uses the Kabadi-Danner mixing rules to account for the water-hydrocarbon immiscibility.¹⁸ This method has been designed to be implemented in accordance to the NBS steam tables to obtain accurate results.

To summarize the different process conditions, Table 2 collects the main properties and assumptions of the system.

Table 2. Main system properties and assumptions for Scenario I.

System parameters	Values
Inlet (Wood+Glycerol) tonne/ day	1000
Inlet Composition, wt%	
Dry wood	17.8
Glycerol	16.5
Water	65.7
HTL	
Temperature, °C	400
Pressure, bar	300
Yields, kg/100 kg dry wood	
Biocrude	30
Gas	13
Solids and dissolved organics wt%	57
Biocrude composition, wt%	
Volatile fraction	45.7
Non-volatile fraction	54.3
Gas phase composition, wt. %	
CO ₂	92.72
CO	2.79
CH ₄	2.43
H ₂	2.07
Aqueous phase composition, wt. %	
H ₂ O	77.05
Dissolved organics	22.95
Thermal cracking	
Temperature, °C	600
Pressure, bar	4
Conversion rate, wt. %	70
Hydrotreating	
Temperature, °C	360
Pressure, bar	77.5
H ₂ consumption, gH ₂ /g biocrude	0.036
Product distribution, wt. %	
Gasoline equivalents	80.25
Water	19.76
Boiler	
Efficiency, %	90

Results and discussion

Energy analysis

Energy is a major concern for an industrial plant, since an optimal configuration implies economical savings.

From the thermal requirements a pinch analysis was performed obtaining the minimum hot and cold utility demands. A minimum temperature approach of 20°C was used. The energy requirements presented are calculated per kilogram of organic matter fed to the system (kJ/kg).

The overall hot and cold utility demands, before incorporating any heat recovery, are 7404 and 5325 kJ/kg, respectively. Figure 2 illustrates the Grand Composite Curve (GCC), which provides the optimum hot and cold utility demands after heat integration. From this it can be seen that the minimum heating/cooling requirements are 3314 and 552 kJ/kg, for the optimum case, corresponding

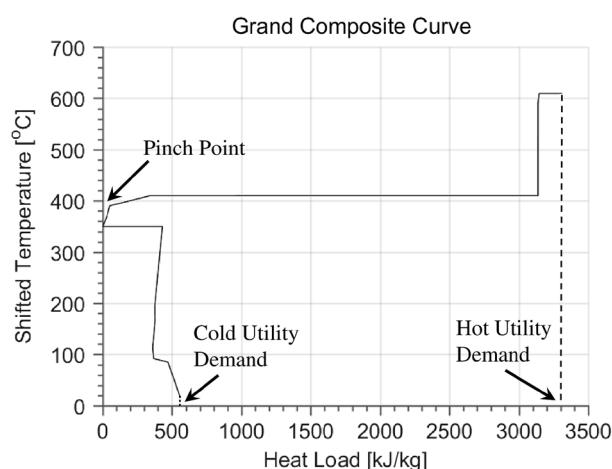


Figure 2. Grand composite curve for drop-in fuel production.

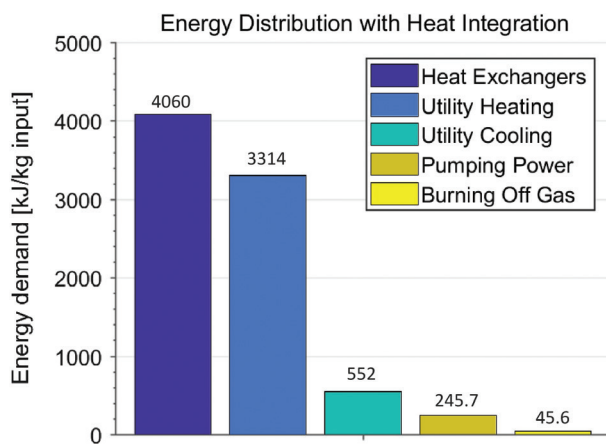


Figure 3. Heat duties, pump work, and combustion of the off-gas for drop-in fuel production.

to heating and cooling demand reductions of 55.22% and 89.6%, respectively.

Figure 3 illustrates the energy distribution in the system. It can be seen that the required energy is 3559.7 kJ per kg of organic matter fed to the system, this value is only accounting for the heating and pumping demand. After integrating the heat recovery and combustion of the off-gas, the utility heat demand was reduced to 44.1% of the original demand. The cooling demand seen in Fig. 3 is supplied by a cooling system which has not been further investigated.

Economic study

The economic analysis is divided into two sections: A market prices analysis of the different expenditures, and a sensitivity analysis of the main parameters affecting the MFSP. A plant lifetime of 20 years and an availability of 90% are assumed.

MFSP determination

Capital costs (CAPEX)

The determination of the plant cost is based in literature.^{5,6,8,19} The prices presented on these referential studies were adapted to fit the characteristics of the plant conceived on this work. For instance, the costs for water treatment and hydrogen production plants were excluded from the capital costs. Moreover, costs not well specified on the references were also excluded (e.g. missing equipment⁵). These modified prices have been scaled to a feed rate of 2915 tonne/day which considers the global load of the plant (wood, water, and other components).

The prices were updated to the present value based on a specific cost index for machinery and equipment,²⁰ as shown in Eqn (2). Then prices were scaled to the designed pressure, temperature, and plant capacity using a $b=0.8$ scaling exponent, as shown in Eqn (3).

$$\text{Cost}_{\text{ref plant 2017}} = \text{Cost}_{\text{ref year}} \left(\frac{\text{Cost-index}_{2017}}{\text{Cost-index}_{\text{Ref year}}} \right) \quad (2)$$

$$\text{Cost}_{\text{plant}} = \text{Cost}_{\text{ref plant 2017}} \left(\frac{\text{Plant capacity}}{\text{Ref P capacity}} \right)^b \cdot \left(\frac{f_{\text{TNEW}}}{f_{\text{Tref}}} \right) \cdot \left(\frac{f_{\text{PNEW}}}{f_{\text{Pref}}} \right) \quad (3)$$

The correction factors used for pressure and temperature were taken from Smith.²¹ The plant cost was calculated to 174.5 million USD, based on the mean value of the scaled and updated prices of the reference cases, shown in Table 3. The price in Tews *et al.*⁶ is considerably lower than for the other studies, which is probably due to the scaling method

Table 3. Original and updated value of similar plants.

Plant capacity [tonne/day]	Wood rate	Appraisal year	Original capital cost [M\$]	Scaled and updated capital cost [M\$]
25000	8	2011	360.3 (6)	92.1
13333	15	2007	471.5 (5)	178.2
6695	20	2011	264.1 (8)	191.8
6667	30	2011	395.1 (19)	235.7
Base Plant Cost				174.5

Table 4. Parameters used in the base cost calculation and sensitivity analysis for Scenario I.

General Assumptions		Min.	Base cost	Max.	Sources
Plant lifetime [years]			20		
Plant Capacity (Wood + Glycerol)[tonne/day]			1000		
Operative plant time [days/year]		295.65	328.5	361.35	
Drop-in fuel yield, [kg/100 kg organic matter input]		23.05	20.95	18.86	
Parameters Variation		Min.	Base cost	Max.	
CAPEX	Plant cost [Million USD]	119.3	225.8	305.3	(5, 6, 8)
VOC	Wood + Transportation [USD/tonne]	37.00	41.50	70.00	(5, 22)
	Glycerol [USD/tonne]	190.00	230.00	440.00	(25)
	H ₂ cost [USD/kg]	1.90	3.51	5.12	(23, 24)
	Electricity cost [USD/kWh]	0.116	0.118	0.121	(26)
	Natural Gas cost [USD/kWh]	0.03	0.036	0.041	(27)
	Wood grinding energy [kWh/tonne]	20.00	130.00	240.00	(28)
	Water disposal	2.5 % of VOC			(4)
FOC	Fixed Operating Cost	17.5 % of VOC			(5, 6)
Bio-fuel production cost, USD/LGE		0.83	1.14	1.87	
Fossil fuel production cost, USD/L		0.44	-	0.64	(29)

used, accounting for the total plant capacity and not considering the wood feed rate which is lower for the mentioned study. Including a 5% annual interest loan with a payback time of 10 years, the total plant costs were calculated to 225.8 million USD.

Operating costs (OPEX)

OPEX are directly related with the plant production, accounting for the necessary expenses to run the plant. The OC can be divided into two categories:

- Fixed operating costs (FOC): Account mainly for operation and maintenance labor (O&M) for the plant. In this work fixed at 17.5 % of the VOC.^{5,6}
- Variable Operating Costs (VOC): Depend on the amount of processed material and energy consumed for the process.

The FOC includes e.g. the price of the hydrotreatment catalyst (a CoMo/F-Al₂O₃ catalyst in a two-stage hydro-treater) based on Zhu *et al.*⁵ and Tews *et al.*⁶

Fuel production cost

The drop-in fuel production cost is calculated with the parameters presented in Table 4 for the base case. Furthermore, the best and worst scenario are calculated based on market prices and yield variation. The presented costs are on basis of liter gasoline equivalent (LGE).

It should be noted that the price of some parameters depends on the production technology. Therefore, the hydrogen can be produced from steam reforming with a low price 1.9 USD/kg, but it increases to 5.12 USD/kg if it is obtained from electrolysis. It is worth mentioning that the glycerol price ranging between 190 and 440 USD/tonne, does not include refined glycerol which has a higher price. The cost variation for this feedstock could be related to the purity of the product. The price of the crude product will be used on the subsequent economic analysis. It is expected that the unrefined material will behave as well as the pure glycerol for the process.

The selected wood feedstock is sawn timber residue. The price range of the timber residue has been set to 37 and

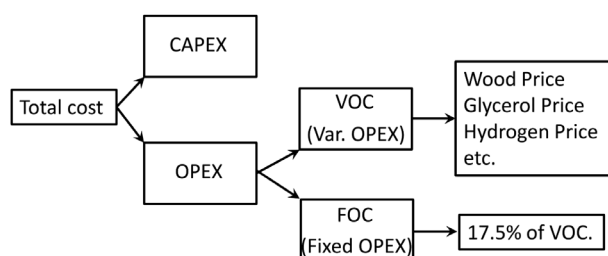


Figure 4. Representation of the total cost distribution.

46 USD/dry tonne,²² which can increase up to 70 USD/dry tonne⁵ in the least optimistic case. This wide range reflects the different wood qualities and even geographical location for the wood extraction, with the lowest prices given by a US study.²² The other study⁵ does not detail the material origins, but is included due to the importance of this feedstock on the further economic analysis.

Sensitivity analysis

The sensitivity analysis is an economic tool designed to identify the parameters responsible for major cost variations.

Three different scenarios have been analyzed. In Scenario I the organic matter fed to the system is constituted by a mixture of 50/50 biomass-glycerol. For the second case, the glycerol has been substituted by water, therefore the inlet is reduced to 500 tonne/day of wood, but the overall mass flow is maintained. In Scenario III, the biocrude is totally deoxygenated and saturated. The prices influencing the MFSP for the three scenarios is shown in Table 5. Most of these prices will be used as starting values for the sensitivity analysis.

Table 5. Cost calculations for Scenarios I, II, and III.

	I	II	III
Cost parameters	[USD/LGE]	[USD/LGE]	[USD/LGE]
Wood residue	0.079	0.119	0.119
Crude glycerol	0.439	0.000	0.000
Hydrogen	0.121	0.101	0.197
Electricity	0.031	0.046	0.046
Thermal energy	0.141	0.214	0.214
Wood grinding	0.035	0.052	0.052
Plant cost	0.131	0.191	0.191
Water disposal	0.021	0.013	0.016
Fixed operating cost	0.147	0.092	0.109
Gasoline equivalents cost	1.141	0.821	0.935

Scenario I

Figure 6 shows the economic sensitivity of the biocrude production for the co-liquefaction scenario.

The white and grey bars represent the fuel cost change from the base case. The individual parameters are subjected to a 10% variation, in order to be comparable. The lines across the bars illustrate the variation of the fuel cost within the price ranges established in Table 4. Notice that the yield is not subject to price variations, thereby is only represented by a bar.

The most likely fuel production cost of Scenario I is 1.14 USD/LGE, approx. three times higher than fossil gasoline.

The HTL yield is the most sensitive parameter from the system, producing a cost change of 0.13 USD within a 10% variation.

Moreover, its unequal deviance at each side of the base cost shows that a reduction in the yield has a higher impact on the cost per liter than an equal yield increase.

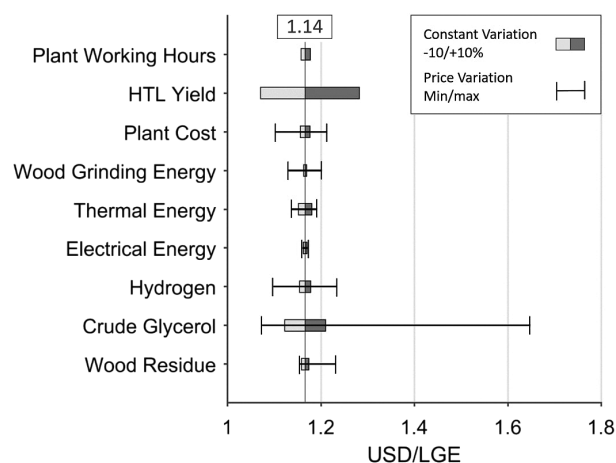


Figure 5. Cost sensitivity analysis for Scenario I.

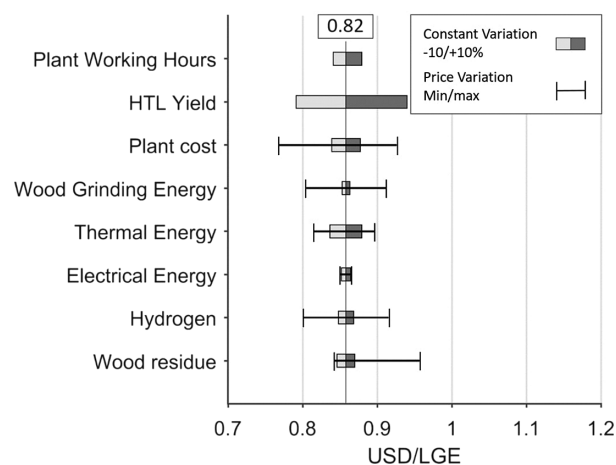


Figure 6. Cost sensitivity analysis for Scenario II.

Regarding the glycerol, the wide price variation represents a problem to the fuel cost stability. Furthermore, glycerol has also a wide bar that confirms the large effect that it has on the overall cost.

Hydrogen has the third major effect on the cost variation, with a price variation which depends on the production technology, where electrolysis represents the upper limit.

Despite the high consumption, wood has a small impact on the production cost due to its low cost. Therefore, it represents a small percentage in the total cost share.

The remaining parameters have a minor impact on the cost, where the electricity and thermal energy have a low impact on the system with a low-cost reduction potential.

As stated, glycerol represents a key parameter for the process viability due to its high cost and price range. Thus, it is necessary to study the economic behavior of the system using only wood as organic feedstock.

A minimum fuel selling price (MFSP) of drop-in fuels, accounting only for the production cost, is calculated based on the lower and higher market prices. The range obtained is 0.83–1.87 USD/LGE.

4.2 Scenario II

In this scenario, the glycerol is excluded from the feedstock but the wood feed rate remains unchanged. The system performance is unaffected except for the yield, which increases the conversion performance of the HTL by 10% due to the lack of glycerol.¹⁰ The changes made between Scenarios II and III are shown in Table 6. It should be noted that char formation increases by 25% when glycerol is excluded,¹⁰ which could generate other variations in the process.

Figure 7 represents the economical sensitivity of different parameters in the system. In this case, the production cost has been reduced to 0.82 USD/LGE, which represents a 28% decrease compared to Scenario I. The total mass flow of drop-in fuels is lower than the previous case, as the system is fed only with 500 tonne/day of wood.

Table 6. Updated parameters used in Scenarios II and III.

Parameter	II	III
Plant Capacity (Wood) [tonne/day]	500	500
Glycerol content	0 %	0 %
Biocrude yield of input wood [%]	40 %	40 %
Gasoline equivalents yield of input wood [%]	27.8 %	27.8 %
H ₂ consumption [g H ₂ /g of biocrude]	0.0356	0.0482
Gasoline equivalents cost [USD/LGE]	0.821	0.935

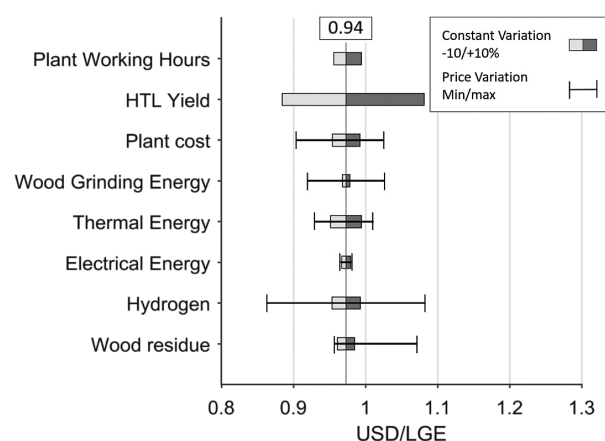


Figure 7. Cost sensitivity analysis for Scenario III.

The HTL yield is still the most dominant factor, maintaining the same behavior described in the Scenario I. The hydrogen consumption is proportional to the amount of biocrude processed, thus this parameter is not affected by the exclusion of glycerol from the system, but due to the share, it becomes the most sensitive parameter after the yield.

On the other hand, the price of wood accounts for a larger effect on the cost variation, as constitutes the only organic inlet of the system.

The sensitivity bar for the wood is equal in size to the thermal energy and plant cost, meaning that a variation on these parameters equally affects the fuel cost.

The rest of the parameters increase proportionally their share of the production cost, but still having a minor effect on the price contribution.

The beneficial effects of the glycerol in the HTL system must be further studied as it represents a major cost of the bio-fuels production. Therefore, it must be evaluated if the reduction of the char formation compensates the reduction in the biocrude yield in views of the relative high price of the glycerol, that could be six times higher than wood.

The MFSP ranges between 0.56 and 1.16 USD/LGE.

Scenario III

As for the previous scenario, glycerol is not included in Scenario III either. It is assumed that complete deoxygenation and saturation of the biocrude, consequently the hydrogen consumption increases to 0.0482 kg/kg of biocrude. The complete deoxygenation is the only change between Scenarios II and III as shown in Table 6. It is expected an increase of the HHV due to de lower oxygen content of the drop-in fuels. Figure 7 presents the sensitivity analysis of the third case.

Table 7. Fuel production price for the three scenarios.

Cases	Base cost [USD/LGE]	Cost ranges [USD/LGE]
Scenario I	1.14	0.83–1.87
Scenario II	0.82	0.56–1.16
Scenario III	0.94	0.64–1.31

The additional hydrogen requirement implies a 16% increase of the production cost, which leads to a final price of 0.94 USD/LGE. Therefore, hydrogen represents a sensitive parameter in this case. The MFSP ranges between 0.64 and 1.31 USD/LGE.

Table 7 summarizes the base cost calculated, and a range of production cost built with the lowest and highest prices found in literature for the three cases.

Discussion

In a recent study, de Jong *et al.*¹⁴ carried out an economic assessment on the MFSP of jet fuel derived from several advanced technologies based on fixed-price calculations, i.e., without any sensitivity analysis. From this assessment, they concluded that the MFSP was approximately 1081 USD/tonne, which is about 7.5% lower than the base cost of Scenario III (1169 USD/tonne, 0.94 USD/LGE), the scenario most similar in the current work.

Investigating the cost parameters used by de Jong *et al.*¹⁴ and the present study, clear differences appear. Therefore, in the following the sensitivity of three essential external parameters, namely the electricity, hydrogen, and wood prices, by using different sources for calculating the MFSP of Scenario III, will be discussed.

The first parameter to be discussed is the price of electricity. The base electricity prices used are 0.118 and 0.0784 USD/kWh, respectively, for the current study calculated from the mean EU price for the last 5 years and by de Jong *et al.*¹⁴ When evaluating the differences in electricity purchase price, the lower price is the result of a larger plant consumption and a tax-free price. Adjusting the electricity price used in the current study to that used by de Jong *et al.*¹⁴ Fig. 8 shows that the MFSP of Scenario III will drop just about 5% to a level very close to that of de Jong *et al.*

The second essential price difference appears for the cost of hydrogen. De Jong *et al.*¹⁴ use 1.00 USD/kWh compared to 3.51 USD/kWh in the current, with is calculated from the mean values from IEA²³ and Ainscough *et al.*²⁴ (1.9–5.12 USD/kg). Whereas it appears difficult to judge the reasoning behind the lower price level, using this lower price in Scenario III causes a substantial reduction in the

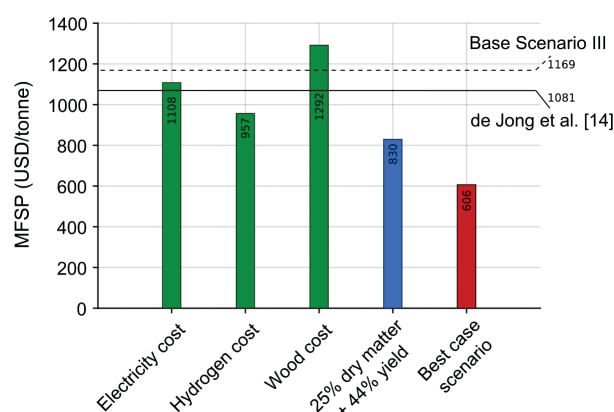


Figure 8. MFSP comparison illustrating the effects of using different external prices from Jong *et al.* on Scenario III (in green). Predicted MFSP by using more high-dry-matter content and yield (in blue).¹² MFSP of a best-case scenario using lowest external prices and high dry matter content and yield (in red).

MFSP by almost 20 % from 1169 USD/tonne to 957 USD/tonne. This sensitivity is also reflected in Fig. 8 and clearly demonstrates the need for robust sources for external parameter selection when carrying out such economic assessments.

The last parameter comparison is the cost of the wood. In the current study the price level is set rather optimistically (41.5 USD/tonne) as compared to the 70 USD/tonne used by de Jong *et al.*¹⁴ As Fig. 8 shows, using a higher wood price significantly affects the MFSP. Using a 70 USD/tonne target increases the MFSP of Scenario III by 10.5 % to 1292 USD/tonne. This comparison not only shows the sensitivity to this parameter from geographical effects, but also shows the need for low cost feedstock for a feasible business case.

In combination with external price parameters such as the three discussed above, internal process performance obviously also plays a role. In the current work, a coherent dataset from continuous HTL all the way through to upgraded fuel product has been used to ensure consistency. This sets this study a bit aside from most other techno-economic assessments, which use best available data from a variety of sources, and obviously predetermines such influential process parameters as dry matter content in the feedstock, and oil yield from the HTL stage, which in turn affect the calculated MFSP of the products. However, a recent study demonstrated 25% lignocellulosic dry matter in the feedstock for continuous HTL, as well as a biocrude yield of 44%.¹² Using such process data, all other parameters kept equal in Scenario III, the MFSP drops to 830 USD/tonne. This price should of course be qualified by using coherent data, but it highlights the

importance of an optimal performance of the HTL stage on the final MFSP.

By combining all the best prices from the discussed cases, an MFSP of 606 USD/tonne is obtained. The price illustrates the importance of a robust model and the need for a specific geographical location, to maintain stable and realistic prices.

Conclusion

This paper assessed the economic viability of gasoline equivalents production through hydrothermal liquefaction of aspen wood, by means of three different case scenarios.

A generic modeling platform was successfully established in Aspen Plus[®] using coherent experimental data to establish mass and energy balances for a subsequent economic assessment. The fixed MFSP for the three scenarios were in the range of 0.82–1.14 USD/LGE (1025–1425 USD/tonnes). From the sensitivity analysis it was found that the biocrude yield is the most sensitive parameter for the fuel production cost, which highlights the need to optimize the internal performance of the HTL process. Furthermore, it was found that the hydrogen consumption, feedstock price, and thermal energy consumption also represent key economic parameters. Based on current glycerol and crude glycerol prices, a co-liquefaction process scheme is unlikely a feasible scenario. The discussion illustrated that without a critical selection of coherent external parameters combined with verified experimental data, highly fluctuating MFSP can be obtained. Based on a best-case selection of internal and external parameters, an MFSP of 0.48 USD/LGE (606 USD/tonnes) can be predicted. Based on the findings of the current work, it is recommended that future work focuses on HTL process optimization, and that geographically dependent economic assessments are evaluated based on coherent experimental data. However, in conclusion this work presents a methodology on how to calculate a robust and preliminary estimate on the MFSP of gasoline equivalents from HTL of lignocellulosic biomass.

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