

# Hydrothermal liquefaction of biomass for the production of diluents for bitumen transport

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## Abstract

This study explores the hydrothermal liquefaction (HTL) of wood chips to bio-crude followed by upgrading to diluents, which are used to transport bitumen through pipelines. In this study, we considered a 2000 dry t day<sup>-1</sup> plant capacity with two scenarios. The first scenario uses hydrogen for upgrading from the on-site hydrogen production plant (i.e., the hydrogen production scenario) and the other relies on procuring hydrogen from an external source (i.e., the hydrogen purchase scenario). We developed a data-intensive process model for HTL and used it to estimate plant capital costs. Project investment costs for the hydrogen production and hydrogen purchase scenarios are 559.67 and 429.13 M \$, respectively. The product values (PV) of the diluent from the two scenarios are  $0.98 \pm 0.03$  and  $0.79 \pm 0.03$  \$ L<sup>-1</sup>, respectively, at a 95% confidence interval. The sensitivity analysis shows that diluent yield and internal rate of return (IRR) have the highest impact on the PV of the diluent, followed by capital cost and biomass cost. The optimum plant size at which the cost of production is lowest is 4000 dry t day<sup>-1</sup> for PVs of 0.82 \$ L<sup>-1</sup> and 0.68 \$ L<sup>-1</sup> for the hydrogen production and purchase scenarios, respectively. This study offers insights into the techno-economic feasibility of producing diluents from HTL. The results of the study could help in the production of diluents for bitumen transportation for

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the oil sands industry and help reduce the overall greenhouse gas (GHG) footprint of the oil and gas sector.

**Keywords:** Biomass; hydrothermal liquefaction; cost; process modeling; techno-economic assessment; oil sands.

## **1 Introduction**

Environmental concerns over increased greenhouse gas emissions (GHG) have led to rigorous regulations on carbon emissions from the use of fossil fuels<sup>1-4</sup>. Biomass, a clean and renewable energy source, is being considered for the production of liquid fuels and chemicals through several biochemical and thermochemical processes<sup>5-9</sup>. Although biochemical conversion is used commercially, there are concerns with respect to its economic sustainability. The increase in fuel production leads to the increase in food prices and has triggered a debate over food versus fuel<sup>2</sup>. Although the commercial scale production of second-generation biofuels is now a reality, its implementation still has challenges<sup>10, 11</sup>. With biochemical conversion technology, investments costs are significantly higher for cellulosic ethanol than for cornstarch- or sugarcane-derived alcohol<sup>10</sup>. Moreover, there are technical bottlenecks for biochemical enzymatic hydrolysis because of the lower specific activity of existing commercial enzymes, high cost of enzymes, and poor understanding of enzyme biochemistry, structure, and mechanics<sup>12, 13</sup>. Other key factors impeding the development of biochemical conversion technology are lack of financial support and subsidiary and political incentives<sup>14</sup>.

In contrast to biochemical approaches, direct liquefaction is a simple conversion technique to produce liquid fuels from biomass. Known direct liquefaction approaches are fast pyrolysis and high-pressure hydrothermal liquefaction (HTL)<sup>15-20</sup>. The presence of water in biomass has a

negative effect on pyrolysis, as it needs the high heat of vaporization, which not only limits biomass options as a feedstock but also negatively affects the economy of the process. Usually, pyrolytic liquefaction processes liquefy biomass with a moisture content of less than 20%. However, water contents in other types of biomass such as tropical grasses, aquatic species like algae, and food wastes can be 80–85% or higher.

In order to fit biomass to pyrolysis applications, processes related to drying such as atmospheric drying, mechanical dehydration, and other techniques have been investigated; unfortunately, none has proven to be economical. Therefore, as mentioned earlier, as processes such as pyrolysis do not handle high moisture feedstock well<sup>19-21</sup>, HTL is seen as an alternative solution for handling wet biomass because the process eliminates the cost-intensive drying step<sup>19, 22, 23</sup>. Water as a reactant acts as a catalyst and acquires the properties for suitable liquefaction in terms of increased density, heat transfer capabilities, and quick decomposition. HTL processes biomass at 250 - 380 °C at 5 to 30 MPa and a residence time of 5 - 60 min<sup>19, 24-26</sup>. The main products obtained are bio-crude, gas, and an aqueous fraction<sup>19, 27</sup>. HTL bio-crude has a lower oxygen content (10-20 wt%) and a higher heating value (which is 35 MJ kg<sup>-1</sup><sup>27-29</sup>) than pyrolysis bio-crude<sup>30-32</sup>. HTL bio-crude's heating value is similar to the heating value of 40 - 45 MJ kg<sup>-1</sup> for conventional petroleum fuels<sup>16, 33</sup>. During the HTL process, 85% of the oxygen in biomass is removed as H<sub>2</sub>O and CO<sub>2</sub><sup>34</sup>. The oxygen is as little as 10 wt% in bio-crude, resulting in a higher caloric value than the biomass feed itself. The bio-crude from HTL can be upgraded further by removing oxygen through hydrotreating<sup>15, 35</sup>. Hydrotreating refers to the stabilization and selective removal of oxygen from raw bio-crude through its reaction with hydrogen over a catalyst (e.g., alumina-supported, sulfided CoMo/NiMo, or noble metal catalysts).

Although biomass gasification and pyrolysis have been used commercially<sup>27, 36</sup>, biomass HTL has only been demonstrated at a pilot scale. And although hydrotreating in heavy oil is a well-known refinery process<sup>37</sup>, the removal of oxygen from HTL bio-crude by hydrotreating has not been realized at a commercial scale. In addition, the design of a bio-refinery facility with a biomass processing capacity of 2000 dry t day<sup>-1</sup> or more usually takes at least four years before start-up<sup>38</sup> and is not easily modifiable once designed<sup>39</sup>. Thus a detailed techno-economic analysis is needed to understand process performance parameters and acceptable levels of risk at appropriate market conditions. Although there is published research on the techno-economics of thermochemical-based technologies such as gasification and fast pyrolysis<sup>30, 40-45</sup>, there has been little research on HTL-based techno-economic models<sup>34, 46-49</sup>. No studies focus on the techno-economic assessment of diluent production through the HTL of biomass feedstocks. The stabilized bio-crude can be used as a diluent to reduce the viscosity of bitumen such that the mixture of diluent and heavy hydrocarbons has an appropriate density and viscosity, subject to pipeline specifications<sup>50</sup>. As the production of biofuels from biomass HTL is yet to be done at a commercial scale, it is imperative to use a modeling and simulation tool to examine process performance and viability based on technical and economic parameters. There is a need to understand this, and that is a key gap in knowledge that this paper tries to address. The overall aim of this study is to conduct a comprehensive techno-economic assessment of production of diluent from biomass through HTL process. The specific objectives are to:

- Develop a process model for HTL followed by stabilization of bio-crude for the production of diluents;
- Develop the capital cost estimates for a 2000 t day<sup>-1</sup> diluent production plant using the process model;

- Develop scale factors for a HTL plant for the production of diluents;
- Estimate diluent production cost in two scenarios:
  - Scenario 1 - the hydrogen required for upgrading is generated on site through steam reforming using natural gas
  - Scenario 2 - hydrogen is purchased from an external source.
- Determine the optimum plant capacity for diluent production;
- Conduct a sensitivity analysis to understand the effects of parameters that affect the economics of the process;
- Conduct an uncertainty analysis using the Monte Carlo approach to assess the associated uncertainty on the PV of diluent;
- Conduct a case study for Canada.

## **2 Stabilized bio-crude as a diluent**

Because of alarming environmental concerns and limited conventional oil reserves, there is growing interest in unconventional oil reserves like those in the oil sands in Western Canada. Bitumen production is projected to reach 3.8 M barrels day<sup>-1</sup> by 2022. Strict regulations from policies such as the European Fuel Quality Directive, the Low Carbon Fuel Standard, and Alberta's Specified Gas Emitters Regulation mandate the lowering of greenhouse gas emissions<sup>51</sup>. Because bitumen is compositionally complex and highly viscous, it is difficult to transport and refine. In order for heavy, enriched oil sands to penetrate energy markets, processing technology development is needed. Pipelining offers a convenient and economical method of transporting heavy hydrocarbons over long distances; however, the high viscosity of the hydrocarbons could create operational difficulties like clogging and high-pressure drops. Heavy hydrocarbons are characterized by the presence of asphaltenes, salts, and minerals.

Known approaches to facilitate the transportation of heavy hydrocarbons include drag reduction, viscosity minimization, and in situ upgrading<sup>52</sup>.

In this study, we look for ways to produce agents from renewable feedstocks that reduce the viscosity of bitumen so that it can be easily transported by pipeline. The viscosity of bitumen is usually reduced with dilution agents. Dilution agents are natural gas condensates that normally consist of heavy oil fractions from lighter hydrocarbons but can include lighter crude oil fractions. Diluents help transport heavy hydrocarbons by pipeline and further facilitate dehydration and desalting downstream<sup>53</sup>. The use of diluents produced from natural gas condensates adds to the GHG emission load of the oil sands industry. Hence, there is a need to obtain diluents from an environmentally friendly resource such as biomass. This study develops a preliminary case for use of stabilized bio-crude as a diluent for bitumen transportation by pipeline. This research highlights the use of biomass for the production of stabilized bio-crude through HTL, which could be used as a diluent to lower viscosity in bitumen.

Diluent reduces the viscosity of bitumen by weakening the intermolecular forces<sup>54</sup>. Several studies have been done on the commercial use of liquid solvents as diluents to reduce the viscosity of bitumen<sup>55-58</sup>. The diluent has always less viscosity than the heavy hydrocarbon<sup>59</sup>. The ideal light solvent is the one that leads to required viscosity reduction but not to the precipitation of asphaltene fractions<sup>60</sup>. Some known solvents such as naphtha, toluene, tetrachloride, and benzene have been used in the Athabasca oil sands for oil recovery with and without steam<sup>57, 61</sup>.

Diluent consists mainly of paraffinic liquid hydrocarbons<sup>60, 62, 63</sup>. Recently, a process involving liquid pyrolysis oil as a bio-diluent obtained from biomass pyrolysis was patented<sup>64</sup>.

Pollard et al. indicated that pipeline transport specifications for diluted bitumen required appropriate density and viscosity for better performance. Another study reports a patent describing the methods of extracting oil from bitumen using Fischer-Tropsch liquid fuel as diluent<sup>65</sup>. The research determined the potential use of Fischer-Tropsch liquid fuel to reduce the viscosity of bitumen to facilitate transport. The stabilized bio-crude produced through HTL is highly deoxygenated (< 2% O<sub>2</sub>) and could be used as a synthetic crude product<sup>66</sup>. In general, the extraction power of solvent is significantly influenced by physical properties such as density, viscosity and solubility<sup>67</sup>. The lower the diluent viscosity, the lower the viscosity of the blended heavy hydrocarbon-diluent mixture<sup>68</sup>. Table 1 provides the comparison of the physical and chemical properties of stabilized bio-crude from HTL with petrochemical diluents.

It is obvious that the stabilized bio-crude, as a synthetic crude product, is promising as a diluent because its properties show considerable parallels with known diluents for viscosity reduction in the pipeline transport of bitumen. The specific gravity of bio-crude is in the same range as that of known diluents. Both stabilized bio-crude and diluents have low viscosities, which will make heavy hydrocarbons such as bitumen flowable. In addition, the total acid number (TAN) of stabilized bio-crude is significantly low; the low TAN will help avoid corrosion during bitumen pipeline transportation. The properties of stabilized bio-crude from HTL and of diluents are summarized in Table 1<sup>58, 69-81</sup>.

## **Table 1**

### **3 Materials and methods**

The techno-economic analysis was conducted for a process that includes biomass input, HTL, and hydrotreater with or without a hydrogen production plant on site. The process modeling and simulation were done in Aspen Plus<sup>82</sup>. The economic analysis was done through the Aspen Icarus Process Evaluator, which allows for inputs for investment calculations. Available data (or information), wherever necessary, with respect to the operating parameters, were obtained from the literature.

#### **4 Feedstock**

The feedstock considered in this study is whole tree wood chips, which are assumed to have a 50% initial moisture content. A feeding rate of 2000 dry t day<sup>-1</sup> is also assumed. The resulting biomass slurry with 8.2 wt% (dry) is used as input to the HTL reactor. The proximate and ultimate values of the feedstock based on dry matter, as obtained from literature, are shown in Table 2<sup>83</sup>.

### **Table 2**

#### **5 Process modeling**

The simplified block diagram in Figure 1(a) depicts hydrothermal liquefaction-based upgrading. The process has three parts, feedstock preparation, hydrothermal liquefaction, and hydrotreating with or without an on-site hydrogen plant. Initially, wood is crushed to fine particles and mixed with water to form biomass-water slurry. The resulting slurry is then pumped to a high pressure HTL system where it is preheated by the incoming hot effluent through the HTL reactor. Solid residues are filtered as ash from the hot effluent passing through the HTL reaction vessel. The products from the HTL reactors are bio-crude, an aqueous stream with dissolved organics, and gas. The hot effluent, devoid of solid particles, is cooled and separated into two liquid streams and a gas phase. The first liquid stream is the bio-crude and the other is wastewater (aqueous)



with dissolved organics. A part of the aqueous phase is recycled back for feedstock preparation while another part is fed to the wastewater treatment facility. The resulting raw bio-crude passes through the upgrading system, which includes hydrotreating in the presence of catalyst (Ru/C) to reduce its oxygen content<sup>84</sup>. Fast pyrolysis oil, which usually produces more oxygen than oil through HTL<sup>66</sup>, requires a two-step hydrotreating process, and single hydrotreating is assumed to be sufficient for HTL oil. Once upgraded, the bio-crude is passed through a debutanizer column and a set of distillation columns to produce value-added hydrocarbons such as stabilized bio-crude. The HTL and hydrotreating model developed for biomass in this study was validated through experiments, as reported in the literature<sup>83</sup>. The process model developed for the production of diluent from biomass through HTL is shown in Figure 1 (b).

## **Figure 1**

### *6 Hydrothermal liquefaction (HTL)*

The biomass slurry is pumped to an operating pressure of 18 MPa, then run through a softener, where it comes in contact with hot recycle water from the HTL reactor and turns into a paste-like material<sup>34</sup>. The slurry then passes through a heat exchanger, where it is preheated by hot incoming liquid effluent from the HTL reactor. During this process, the water is slightly below a supercritical state, which makes dissolution of organics easier. The incoming effluent then goes through the HTL reactor at 355 °C and 20.3 MPa. The reactor is a plug flow type rather than a continuous stirred tank reactor, because plug flow reactors are highly economical<sup>49</sup>. The residence time in the reactor is usually between 2 and 100 min<sup>34</sup>. The effluent then passes through a filter to separate fine solid residues in the form of ash, which we assume is disposed as solid waste. The mass yield distribution of HTL product components is used to simulate the HTL reactor. The experimental results obtained from gas chromatography, mass spectrometry, and

high-performance liquid chromatography help identify the major components needed to simulate HTL products. The products require minor adjustments to the mass yields in order to facilitate closure of mass balance, which is determined by comparing the density and boiling point curve of the simulated oil with the density and simulated distillation (Simdist) of actual experimental testing results<sup>49</sup>. The filtered effluent passes back through a heat exchanger where heat from the filtered effluent is passed to the incoming biomass stream, and the filtered effluent drops to 148 °C.

The cold effluent is depressurized to 0.25 MPa and then split into three streams: an aqueous phase, an organic phase with bio-crude, and a gas phase. The HTL oil, a mainstream product, makes up roughly 45-50 wt% of the biomass on a dry basis. Approximately 90% of the aqueous phase is recycled back to the softener, where it comes in contact with the biomass so as to obtain the desired wt% of biomass into the incoming stream. A part of the aqueous phase comprising water with small amounts of dissolved organics is directed to a wastewater treatment facility, as the aqueous phase does not have enough organics present for further recovery, from an economics point of view<sup>85</sup>. The wastewater facility is assumed to use anaerobic digestion to convert aqueous organics into carbon dioxide and methane.

The gas components, comprising carbon dioxide, hydrogen, methane, and light alkanes, are assumed to be used in the hydrogen generation plant for steam reforming in the hydrogen production scenario. The resulting bio-crude is fed to the downstream hydrotreating unit where it undergoes deoxygenation in the presence of a catalyst<sup>86</sup>. The design specification of HTL is shown in Table 3.

## 7 *Upgrading*

Crude bio-crude is a heavy organic liquid high in oxygen. The oxygen can be reduced by stabilization through hydrotreating. Hydrotreating is a well-defined process in petroleum refineries. It involves treating HTL bio-crude with hydrogen at a temperature of approximately 400 °C. Very few publications discuss the hydroprocessing of bio-crude from HTL<sup>15, 83, 87, 88</sup>. There is little detailed analysis on upgrading HTL bio-crude; however, it is assumed to be similar to fast pyrolysis upgrading<sup>83</sup>.

For upgrading, the HTL bio-crude is pressurized to 14 MPa before it comes in contact with the compressed hydrogen, depending on the scenario. The effluent from the upgrading process goes through a separation unit into an aqueous phase, off-gases, and upgraded oil. The aqueous phase water can be sent to a wastewater treatment facility while the off-gases could be redirected to the hydrogen production plant. The off-gases from the hydrotreating units are routed to a pressure swing adsorption column to recover unused hydrogen. The hydrogen is then recycled back for use in the upgrading process. The resulting tail off-gases from the column are combined with gases from the HTL system. The hydrotreated bio-crude is debutanized to stabilize it by removing light components including butane in a lights-removal column. The overhead off-gases from the column are directed to the hydrogen plant for their use in hydrogen production. The stabilized bio-crude is assumed to have the same properties as those of a diluent or synthetic crude product. The design specifications for the upgrading section are given in Table 3.

### **Table 3**

## 8 *Hydrogen production plant*

Supplemental natural gas is used to produce hydrogen through steam reforming. The resulting off-gases from all processing areas including hydrothermal liquefaction and hydrotreating are not sufficient to meet the hydrogen requirement in the upgrading unit. The composition of the natural gas used for this purpose is obtained from the literature<sup>89</sup>. The natural gas is initially compressed to 2 MPa before hydrodesulfurization and then mixed with superheated steam at 335 °C under 4.5 MPa, with a steam/carbon molar ratio of 3.5<sup>89, 90</sup>. A portion of off-gases from HTL and hydrotreating is compressed and then fed along with natural gas for hydrogen production. Superheated steam at 370 °C under 4.5 MPa comes into contact with makeup natural gas and off-gases from different units in the steam reformer. The resulting product is converted to syngas in the steam reformer operating at 850 °C under 2 MPa. The conversion is followed by a high temperature water-gas-shift to increase the hydrogen content in syngas, and there water and carbon monoxide are converted into carbon dioxide and hydrogen. The water condenses, and the recovery of hydrogen is assumed to be 80%<sup>91</sup>. The saturated steam requirement is met by obtaining heat from the reformer unit and the power demand is assumed to be met through electricity purchased from the grid.

## 9 **Techno-economic analysis**

### 10 *Capital and operating cost estimation*

Once the steady-state process flowsheet model is developed for biomass HTL and the upgrading system in Aspen Plus with input parameters and corresponding mass and energy balances for sizing process equipment, the results are loaded into the economic analysis simulator engine in the Aspen Icarus platform<sup>92</sup>. The model simulates the process equipment connected by mass, energy, or work streams. The model generates energy performance, efficiency, and product

yields. The process model was further used to estimate equipment costs<sup>92-94</sup>. The model equipment was mapped and sized based on the design parameters. The economic analysis in this study is based on an “nth” plant design, which does not take financing, longer startup times, and special financial needs into consideration<sup>95</sup>. This means that we used a capital investment estimate for technology that is mature and commercially available. This hypothetical plant is assumed to be located in Western Canada, and wage rates for labor and supervision are those used in region. The capital cost estimates for standard equipment such as vessels, pumps, heat exchangers, reactors, and compressors were estimated using the process model.

In general, unit operations from Aspen Plus is mapped and purchased equipment costs are obtained from vendor quotes<sup>96</sup>. Figures 2 (a) and (b) show the breakdown of direct and indirect costs associated with a plant’s investment<sup>97</sup>. An installation factor for the purchased equipment costs provided by the model only takes into account costs for piping, electrical, and other installations. Such cost estimates are considerably lower than those reported in the literature<sup>98</sup>. Hence, for typical solid-liquid plants as studied in this work, an overall installation factor of 3.02 is assumed. After an estimate of the total purchased equipment cost (TPEC) has been calculated, the total project investment (TPI) is obtained using factors developed by Peters et al. as shown in Table 4<sup>97</sup>. The plant in our study is assumed to have a process efficiency of 70% in the first year, 80% in the second year, and 85% in the third year and beyond. In addition, the construction phase of the project allocates 20%, 35%, and 45% of the project capital cost in the first, second, and third year of construction, respectively. Costs associated with electrical fittings, piping, instrumentation, and contingency, are estimated as a percentage of total purchased equipment costs. With respect to the plant’s location in a certain jurisdiction, a location cost factor of 10% is considered in this economic study. A camping cost of 5% for raw material is included. The

product value (PV) is determined based on the discounted cash flow rate of return analysis (DCFROR) spreadsheet<sup>40</sup>. For the DCFROR, investment parameters of a 10% rate of return and 20 years of economic life are assumed and the PV is estimated at a net present value of zero. A stream factor of 90%, is assumed in this analysis<sup>40</sup>.

## Figure 2

### Table

The variable annual operating costs include raw material cost, catalyst and chemicals costs, utility cost, labor and maintenance costs, operating charges, general and administrative (G & A) costs, and plant overhead. The raw materials take into account the costs of biomass, catalysts, hydrogen, and natural gas, depending on the scenario. The biomass feedstock is estimated to be 45.17 \$ dry t<sup>-1</sup>, which includes the costs associated with transportation, harvesting, silviculture, road construction, nutrient spreading, and the premium paid to land owners<sup>99-101</sup>. The costs of electricity, natural gas, and hydrogen were taken to be 0.067 \$ kWh<sup>-1</sup> <sup>102</sup>, 1.583 \$ GJ<sup>-1</sup> <sup>86</sup>, and 0.743 \$ kg<sup>-1</sup> <sup>103</sup>, respectively. For the hydrogen production scenario, the catalysts costs are obtained from the literature<sup>66</sup>. For the hydrogen purchase scenario, the catalyst cost for hydrotreating was derived from Tews and Elliott<sup>104</sup> and that cost assumed a one-year catalyst lifetime.

The costs for disposal of solids and wastewater produced during HTL processing of biomass are taken from Dutta et al.<sup>40</sup> In general, the cost of wastewater treatment is likely to be influenced by appropriate selection of treatment technologies and the amount of dissolved organics in the HTL aqueous phase. The nature and amount of dissolved organics in the aqueous stream is influenced by HTL reaction phenomena and the efficiency of three-phase separator<sup>105</sup>. The “nth plant” assumed in this study implies future improvements with mature technologies for

a commercial HTL plant, producing higher yields of bio-crude<sup>106, 107</sup>. Therefore, this analysis assumes that there is less organics loss to the aqueous stream, which decreases the wastewater treatment cost<sup>49</sup>. Hence, it is reasonable to assume that the disposal cost of wastewater obtained from HTL processing technology would most likely not be significantly different from that of conventional-based processes. However, a sensitivity analysis with respect to disposal costs was also undertaken to study their impacts on product value.

To operate a 2000 t day<sup>-1</sup> plant, 13 personnel (12 operators and 1 supervisor) per shift are required and three shifts per day are assumed for the labor cost analysis. Plant overhead includes plant security, safety, payroll overall and benefits, janitorial services, phone, and plant communications<sup>108</sup>. 2016 Alberta (i.e., Western Canada)-specific wage rates were used for labor and supervision<sup>95</sup>. The values of the other operating parameters are provided in Table 5. Unless otherwise stated, all cost estimates are in 2016 US dollars; the costs in Canadian dollars were converted based on the Bank of Canada rate of 0.75 CAN on 9<sup>th</sup> March, 2016.

## **Table 5**

### ***11 Sensitivity and uncertainty analyses***

Given the uncertainties associated with cost estimates, a sensitivity analysis was done by selecting key variables to study their influence on the diluent PV. Capital cost is a key sensitivity variable as there can be uncertainties in the cost estimate because of lack of data for and lack of knowledge of HTL design. Moreover, the upgrading platform of raw bio-crude to diluent is still in the development phase and thus limited public information is available. The yields from an upgrading system vary depending on catalyst lifetime and bio-crude separation performance. The

yield of diluent can significantly affect the economics of the process, as we underscore in this study. To account for uncertainty, a Monte Carlo simulation was then performed.

## **12 Results and discussion**

### **13 Cost analysis**

The cost analysis was based on an elaborative techno-economic assessment of HTL technology involving hydrogen production and purchase scenarios for 2000 dry t day<sup>-1</sup> of biomass. Figure 3 shows that HTL, hydrotreating, and hydrogen production account for 66.8% at 72.4 M \$, 10.88% at 11.8 M \$, and 22.32% at 24.2 M \$, respectively, of the purchased equipment cost. The HTL unit has the highest investment costs followed by the hydrogen plant, as also reported previously<sup>66</sup>. In another study, hydroprocessing with dual stage hydrotreating units accounted for 17% of the capital costs<sup>85</sup>. Compared to other existing thermochemical technologies such as pyrolysis, the capital cost of HTL is higher because of the shell and tube design of the HTL reactor, which operates at high temperature and pressure<sup>40, 86</sup>. The single hydrotreating is assumed to be the only upgrading step and thus single-step stabilized bio-crude hydrotreating costs less than two-step hydrotreating.

### **Figure 3**

Table 6 shows the total project investment costs incurred in both scenarios. The total project investment (TPI) for a hydrogen production scenario is estimated to be 559.7 M \$, which corresponds to a PV of 0.987 \$ L<sup>-1</sup>, and the TPI for the hydrogen purchase scenario is estimated to be 429 M \$, leading to a PV of 0.799 \$ L<sup>-1</sup>. The installed costs for a 2000 dry t day<sup>-1</sup> plant for the hydrogen production and purchase scenarios are estimated to be 327.5 M \$ and 251.1 M \$, respectively. The hydrogen production scenario has higher capital costs than the hydrogen



purchase scenario because of the extra equipment required for hydrogen production for hydrotreating. The PV for the hydrogen purchase scenario is 19% lower than the hydrogen production scenario because of the low capital cost in the hydrogen purchase scenario. The annual plant operating cost for the hydrogen production cost is estimated to be 129.5 M \$, which includes raw material cost, operating labor cost, maintenance cost, operating charges, general and administrative costs, plant overhead, utilities costs, and costs associated with solids and wastewater disposal. The product yield in both scenarios is 199.27 M L yr<sup>-1</sup>. The hydrogen generation plant relies on natural gas for the steam reforming process. In both scenarios, all the raw bio-crude obtained from HTL is used for hydrotreating, thus the product yield in both scenarios is almost the same.

#### **Table 6**

The breakdown of operating costs is shown in Figure 4. In both scenarios, the raw material cost is the highest portion of the operating costs, 34% in the hydrogen production scenario and 39% in the purchase scenario. The raw material costs include the costs of biomass feedstock, catalyst, and chemicals. The disposal costs, that is, the costs of solids disposal and wastewater treatment, comprise 3% of the variable operating costs. Wastewater treatment costs can be minimized either by reducing the loss of organics into the aqueous stream through an efficient three-phase separation process or by using a lower-cost wastewater treatment facility, both of which, in turn, would improve the stabilized bio-crude yield<sup>49</sup>. An increase in yield would offset the costs associated with hydrotreating and at the hydrogen production plant. With continuing improvements and efforts in commercialization, the process will likely become economically feasible in the foreseeable future.

#### **Figure 4**

Figure 5 shows the breakdown of various operating costs on the PV of stabilized bio-crude in both scenarios. The raw material accounts for a significant portion of the PV: 33.39 cents L<sup>-1</sup> and 31.36 cents L<sup>-1</sup> for the hydrogen production and purchase scenarios, respectively. Utilities costs contribute 5.26 cents L<sup>-1</sup> and 4.23 cents L<sup>-1</sup> to the cost of the fuel in the hydrogen production and purchase scenarios, respectively. Wright et al.<sup>108</sup> reported electricity costs of 4.33 cents L<sup>-1</sup> for a 2000 t day<sup>-1</sup> corn stover pyrolysis and bio-crude upgrading pathway. The effect of the key parameters is further discussed in the sensitivity analysis.

## **Figure 5**

### ***14 Sensitivity analysis***

#### *15 Plant capacity profile*

For this analysis, HTL technology was chosen for both the hydrogen production and the hydrogen purchase scenarios. The PV of the diluent was estimated by changing the plant capacity from 500 to 5000 dry t day<sup>-1</sup>. Figure 6 shows the variation of PV versus plant capacity. With increasing plant capacity, the PV declines due to benefits of economies of scale in the capital cost, i.e., the capital cost per unit output decreases as the plant size increases. The PV further falls as the decrease in capital costs per unit output is offset by the increase in biomass delivery costs<sup>109</sup>. The biomass delivery cost increases with the increase in plant size because biomass is transported longer distances. The profiles show that the lowest optimum PV is reached at a plant capacity of 4000 dry t day<sup>-1</sup> of biomass; beyond this capacity, the PV shows a slight increase and flattens out with further increases in capacity. This is because of the sharp increase in biomass delivery costs with increasing plant capacity. The plant capacity varies proportionally with the area from which biomass is acquired, and the transportation distance increases with the square root of the area<sup>109</sup>.

In the hydrogen production scenario, the PV decreases from 3.04 to 0.83 \$ L<sup>-1</sup>, and the optimized PV is estimated to be approximately 0.83 \$ L<sup>-1</sup> at a capacity of 4000 dry t day<sup>-1</sup>. In the hydrogen purchase scenario, the PV decreases from 1.09 to 0.69 \$ L<sup>-1</sup>, and the optimized PV is estimated to be 0.68 \$ L<sup>-1</sup> at a capacity of 4000 dry t day<sup>-1</sup>. The optimum plant capacity is a reflection of the potential trade-off between plant capital cost and biomass transportation cost.

## **Figure 6**

### *16 Exploration of key parameters*

The effect of many cost parameters on the PV of diluent was studied through of a sensitivity analysis. The sensitivity analysis was conducted by varying cost factors by  $\pm 20\%$ . The results are shown in Figure 7. The most sensitive parameter is the diluent. Therefore, it is imperative to optimize the design and operation of the hydrothermal liquefaction process to obtain high product yields.

The PVs of the diluent for the hydrogen production and purchase scenarios range from 0.90 to 1.09 \$ L<sup>-1</sup> and 0.73 - 0.89 \$ L<sup>-1</sup>, respectively, when bio-crude yield is changed by  $\pm 20\%$ . Thus, a slight improvement in bio-crude upgrading and separation performance could lower the PV considerably. The other key influential parameters in order of decreasing sensitivity are the IRR, capital cost, and the biomass cost. In the hydrogen production scenario, the PV of the stabilized bio-crude ranges from 0.92 to 1.06 \$ L<sup>-1</sup> with a  $\pm 20\%$  change in IRR, and in the hydrogen purchase scenario, the PV ranges from 0.75 to 0.85 \$ L<sup>-1</sup> with the same changes in IRR. Hence, the PV increases with an increase in IRR in both cases.

The capital cost is another key parameter; it has a profound effect on both the capital investment and the return on investment. As the capital cost is changed by  $\pm 20\%$ , the product

value changes by approximately  $\pm 6.43\%$ . Biomass cost is another sensitive parameter because the cost of procuring biomass can vary depending on location and climatic conditions. When biomass cost is changed from 36.14 to 54.21 \$ dry t<sup>-1</sup>, the PV runs from 0.95 to 1.02 \$ L<sup>-1</sup> for the hydrogen production scenario and 0.76 to 0.83 \$ L<sup>-1</sup> for the hydrogen purchase scenario.

## **Figure 7**

### ***17 Uncertainty analysis***

The lack of representative field data and knowledge for advanced technologies like HTL hinders our ability to estimate costs. Both this lack of accurate and available data and the use of assumptions create uncertainties. The sensitivity analysis showed the effect on PV of changing a single parameter at a time. A Monte Carlo simulation was performed on a 2000 dry t day<sup>-1</sup> plant. We used Model Risk software<sup>104</sup> to run 10,000 iterations on chosen random values from all given parameters influencing the PV. The uncertainty analysis is relevant to costs pertaining to capital, biomass, maintenance, labor, utilities, operating charges, G & A, and plant overhead. The uncertainties were carried out on the cost parameters in the range of 80 - 125% based on the sensitivity analysis results. The PVs from the uncertainty analysis are shown in Figure 8. The simulation for the hydrogen production and purchase scenarios for the woody biomass result in PVs of  $0.98 \pm 0.03$  \$ L<sup>-1</sup> and  $0.79 \pm 0.03$  \$ L<sup>-1</sup> at 95% confidence, respectively.

## **Figure 8**

### ***18 Performance results and comparison with previous studies***

This techno-economic study was carried out for a proposed location in Western Canada. Although some studies have focused on producing value-added fuels through an HTL upgrading platform in other jurisdictions, to the best of our knowledge, none of them has a Canadian

context. The stabilized bio-crude from HTL can be further upgraded into gasoline and diesel-blended fuels, which most published studies have stated. One study highlights the production of stabilized bio-crude through a HTL process wherein the bio-crude is assumed to be sold as a synthetic crude with a liquid product cost of 2.28 \$ gallon<sup>-1</sup> <sup>66</sup>. Another study has looked into producing gasoline and diesel, a further processing product of stabilized bio-crude, with a minimum fuel selling price (MFSP) of 2.40-4.77 \$ gallon<sup>66, 95</sup>. Ou et al.<sup>85</sup> reported an MFSP of approximately 0.68 \$ L<sup>-1</sup> for gasoline and diesel-blended products from defatted microalgae through hydrothermal liquefaction and hydroprocessing. Recently, a study showed that the present state of technology (SOT) for a woody biomass HTL platform results in an MFSP of 1.29 \$ L<sup>-149</sup>. However, the goal case, which assumes commercially employed mature technologies with considerable future advancements, results in an MFSP of 0.74 \$ L<sup>-1</sup>. The goal case assumed that there is less organics loss to the aqueous phase and a single reactor is employed for hydrotreating. Consequently, the cost of hydrotreating for the goal case was ~30% lower than for the SOT case. A similar study involving the techno-economics of HTL of lipid-extracted algae resulted in an MFSP of 0.75 \$ L<sup>-1</sup> <sup>105</sup>. The authors showed that the key factor influencing the cost was biomass feedstock cost followed by product yield.

Most existing studies are focused on the production of gasoline and diesel-blended fuels through the HTL route. Recently, there has been increasing interest in the production of diluents through thermochemical approaches such as pyrolysis<sup>64, 65</sup>. The present study highlights the potential of HTL, a key thermochemical conversion technology, for diluent production through a comprehensive techno-economic assessment. This study is pertinent to the Canadian context as the diluent is widely used in bitumen recovery and processing<sup>110</sup>. The present analysis shows that the PV of diluent is  $0.79 \pm 0.03$  and  $0.98 \pm 0.03$  \$ L<sup>-1</sup> for the hydrogen purchase and hydrogen

production scenarios, respectively. However, at an optimum plant size of 4000 dry t day<sup>-1</sup>, the PV reaches 0.82 \$ L<sup>-1</sup> (3.10 \$ gallon<sup>-1</sup>) and 0.68 \$ L<sup>-1</sup> (2.57 \$ gallon<sup>-1</sup>) for the hydrogen production and purchase scenarios, respectively. These results show that the PV of diluent through HTL is in good agreement with the values reported previously<sup>49, 66</sup>.

During the period 2007-2014, the US gasoline wholesale price was 1.77-2.93 \$ gallon<sup>-1</sup><sup>111</sup>. Based on our estimate, the liquid product via HTL at a base scale of 2000 dry t day<sup>-1</sup> is not yet cost effective when compared with the price of conventional petroleum-derived gasoline products. However, at a larger scale, the diluent production through HTL for hydrogen production scenario becomes competitive. With incentives and subsidies from the government in the form of financial support and appropriate tax credits towards commercial deployment of novel HTL-based technologies, the economic feasibility and competitiveness of these technologies would improve. Furthermore, if the HTL facility is co-located with a refinery, the costs of diluent production through HTL can further be reduced<sup>112, 113</sup>. A hydrogen production plant would not be required in an HTL processing facility if process off-gas from HTL can be directed to the refinery for H<sub>2</sub> production. Moreover, the bio-crude from HTL can also be directed to the refinery for further upgrading. Such an approach would remove the upgrading facility from the HTL, thereby further reducing product costs. Thus, the installation of an HTL facility close to a refinery and government aids would help reduce costs and make the technology more appealing.

### ***19 Challenges and key insights***

Though the technology looks promising at a larger scale, there are risks and uncertainties associated with financing due to the lack of experience in such advanced technology projects.

The relative uncertainties can be overcome through R&D and by performing techno-economic evaluations.

The present analysis considers the technological systems likely to be implemented in the future in Canada. Considerable research and development efforts are required to describe HTL process systems including upgrading and catalyst stability<sup>114</sup>. Key aspects of process development in terms of catalyst performance and stability should also be investigated<sup>86</sup>. Ultimately, direct liquefaction to improve the quality of bio-oil requires less upgrading and would reduce diluent costs<sup>115-117</sup>.

In this study, the model assumes solid contents of ~8.2% in the slurry; this would need a larger reactor volume to handle water during the HTL process. Hence, further research needs to consider scale-up and the feasibility of pumping slurry with high solids content. In addition, there are technical constraints with reactor design in terms of its ability to withstand high temperature and pressure. The present model is based on the assumption that all the ash is removed as solids. However, separation may not be efficient, and some minerals will be left behind in the aqueous phase. The deposition of minerals in the reactor may cause corrosion and fouling, which hinder the long-term operation of HTL process. Furthermore, this model considered the quality of the aqueous phase sufficient for wastewater treatment and the potential fouling effects unimportant. The difficulty lies in developing an efficient separation system to improve bio-crude/water separation performance to improve yield of the bio-crude and subsequent wastewater treatment costs. Furthermore, the analysis of bio-crude components is challenging and cumbersome using standard methods, and a more detailed characterization and fractionation of HTL bio-crude is required to better represent major groups of compounds in bio-crude. Similarly, the stability and quality of bio-oil play an important role in identifying possible

options for bio-crude transport off site in the event that the upgrading facility is not co-located with the HTL plant.

## **20 Environmental impacts**

The generation of fuels via HTL technology results in less greenhouse gas emissions compared to fuel generation from fossil fuels<sup>47, 118</sup>. The emissions from the HTL process are from solid residues and wastewater<sup>49</sup>. Though off-gas produced during hydrothermal liquefaction and upgrading is used in the hydrotreating section, the consumption of hydrogen depends on the amount of oxygen in bio-crude. If needed, the hydrogen requirements are met by the hydrogen plant, which uses natural gas as input feed<sup>49</sup>. This involves a technology that relies on fossil fuel. However, the considerable improvements in process efficiency will improve product yields, thereby minimizing emissions and wastes and leading to the production of sustainable biofuels<sup>119</sup>. What is required are methods to reduce the consumption of petroleum-based diluents in order to further reduce greenhouse gas emissions during oil sands processing. A hydrothermal liquefaction-based technology that converts biomass feedstock into renewable bio-crude fractions will ameliorate the environmental sustainability of oil sands processing when integrated with current infrastructure and technology. Further aspects of environmental considerations of HTL technology are beyond the scope of this study and will be assessed in future work.

## **21 Conclusion**

With decreasing conventional oil reserves, there is an increase in the use of heavy hydrocarbons such as bitumen from oil sands in petroleum refineries. As these hydrocarbons are geographically remote from a refinery plant, they require pipeline transportation to a refinery. The most commonly used technique for pipelining heavy hydrocarbons (i.e., bitumen) is to mix them with a diluent such as naphtha or a natural gas condensate. As stabilized bio-crude and



petrochemical diluents have close similarities, stabilized bio-crude can potentially be used as a diluent in the oil and gas processing industry. This study examined the techno-economic feasibility of producing diluent from woody biomass using hydrothermal liquefaction (HTL) technology through the development of a process model. Two scenarios were considered: HTL coupled with a hydrogen production plant and HTL using hydrogen from an external source. Our analysis shows that diluent can be produced with PVs of  $0.79 \pm 0.03$  and  $0.98 \pm 0.03$  \$ L<sup>-1</sup> for the hydrogen purchase and the hydrogen production scenarios, respectively. Hydrothermal liquefaction technology is still in a very early stage of development. The sensitivity analysis for both scenarios showed diluent yield to be the most sensitive parameter followed by IRR, capital cost, and biomass cost. With changes of  $\pm 20\%$ , other parameters show a small effect on the diluent PV. The optimum plant size beyond which there was no appreciable reduction in the PV of the stabilized bio-crude is 4000 dry t day<sup>-1</sup> of biomass for both scenarios.

For the hydrogen production scenario, the hydrogen plant contributed the second most significant portion to the total purchased costs after the hydrothermal liquefaction (HTL) unit. An HTL-upgrading platform coupled with a refinery that takes in off-gas from different process areas for hydrogen production could reduce process costs. The integration of HTL-upgrading technology would not only reduce the need to purchase hydrogen from an external source but also allow the option to procure hydrogen, if needed, from the refinery at a lower cost. The results of this study will help readers understand the cost structure for producing diluent from biomass, which has a low greenhouse gas footprint. The results could be used by decision makers in industry and government to make investment decisions and formulate policy.

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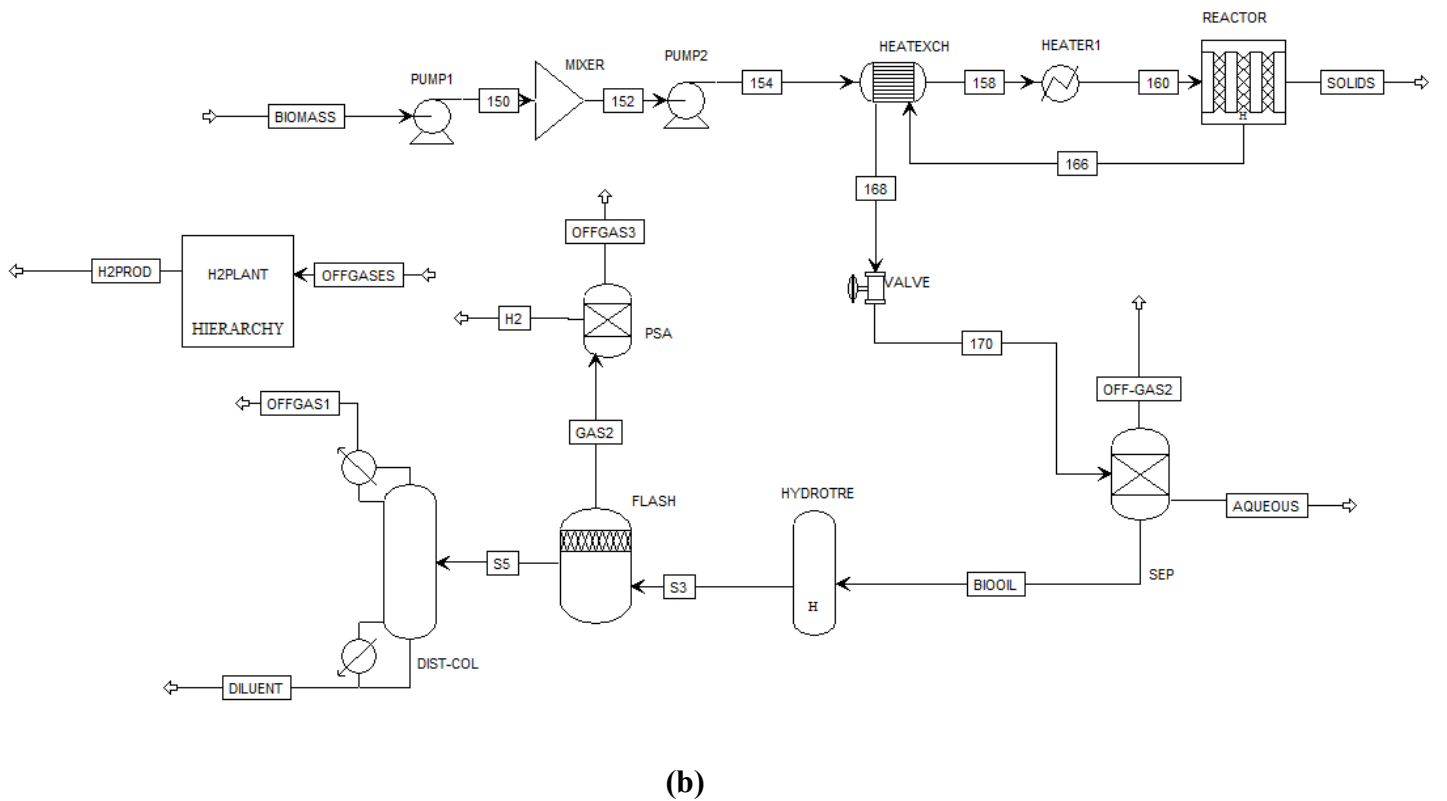
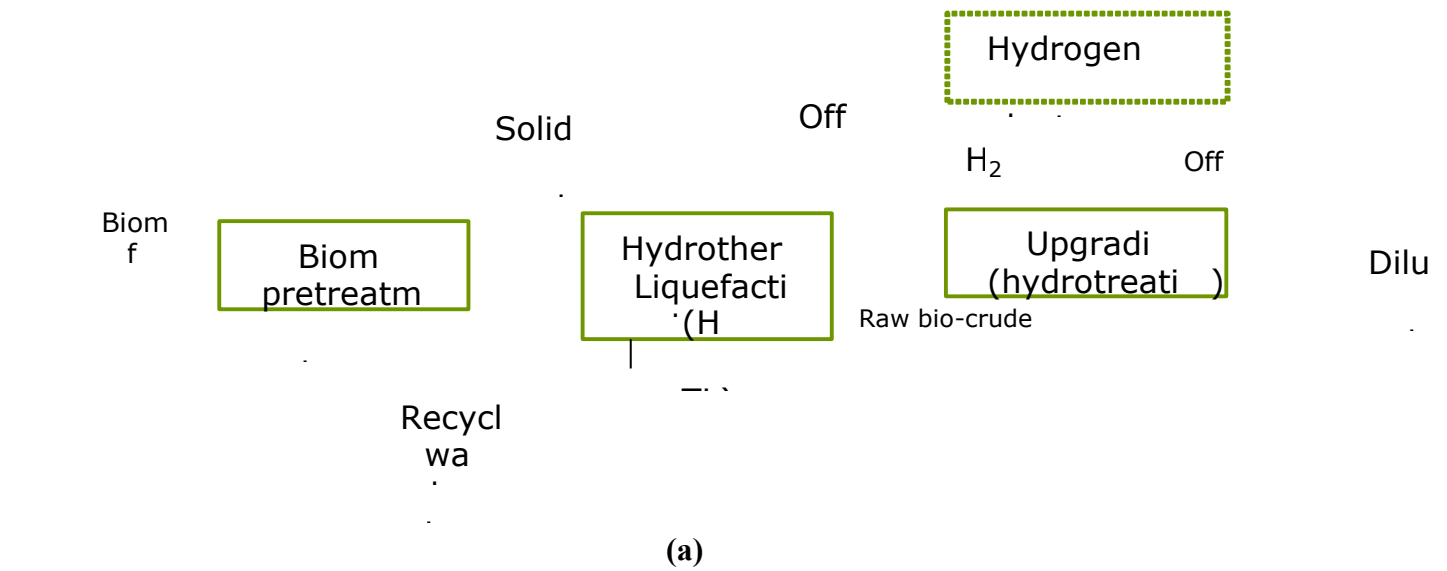
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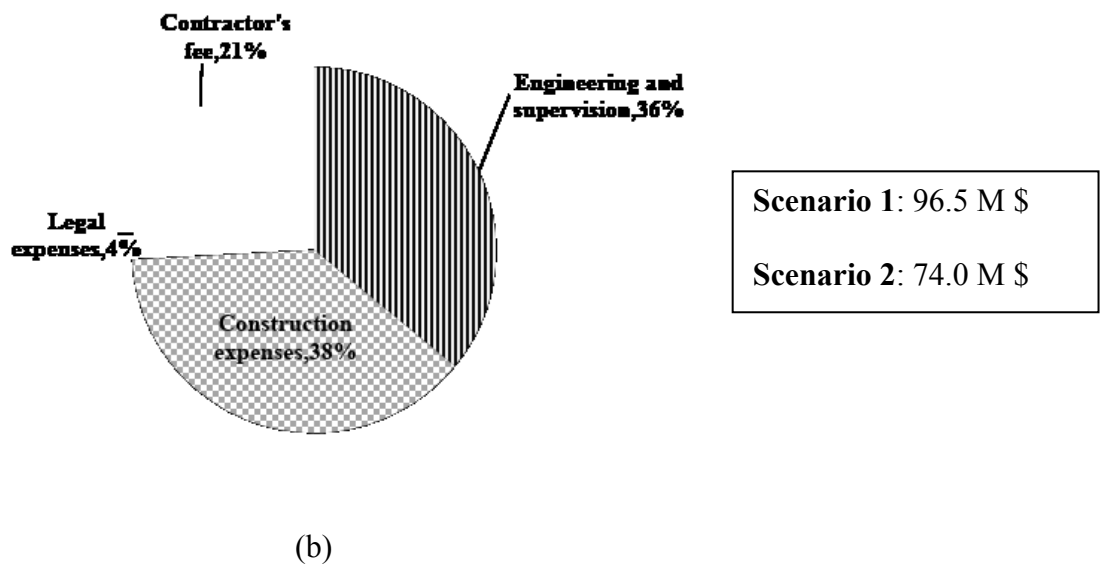
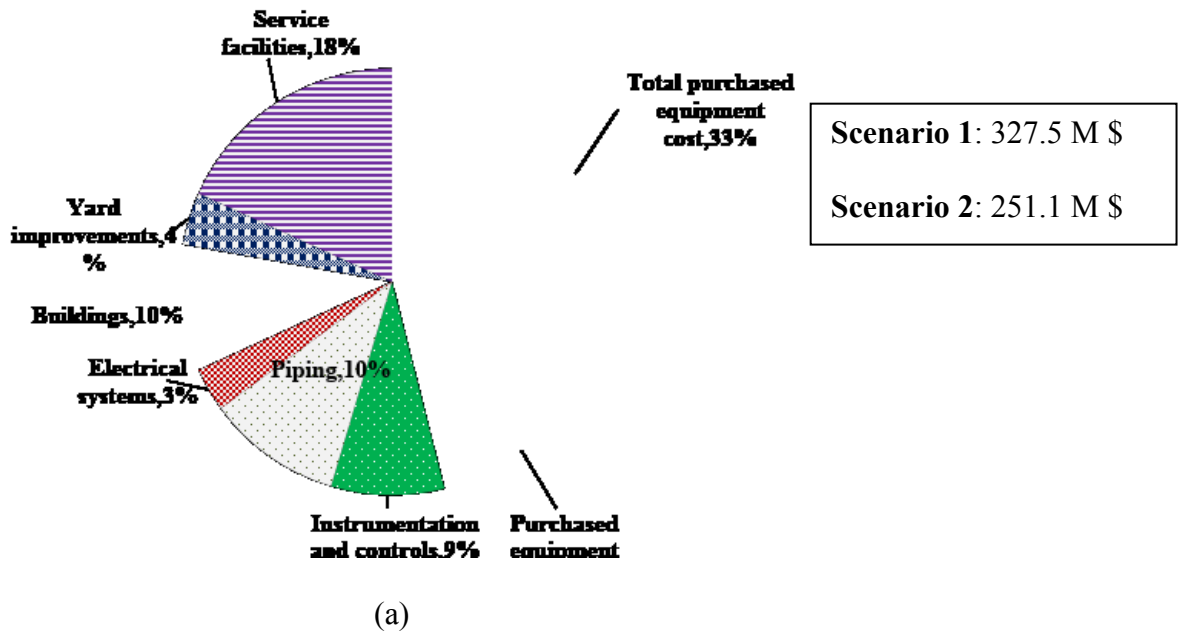
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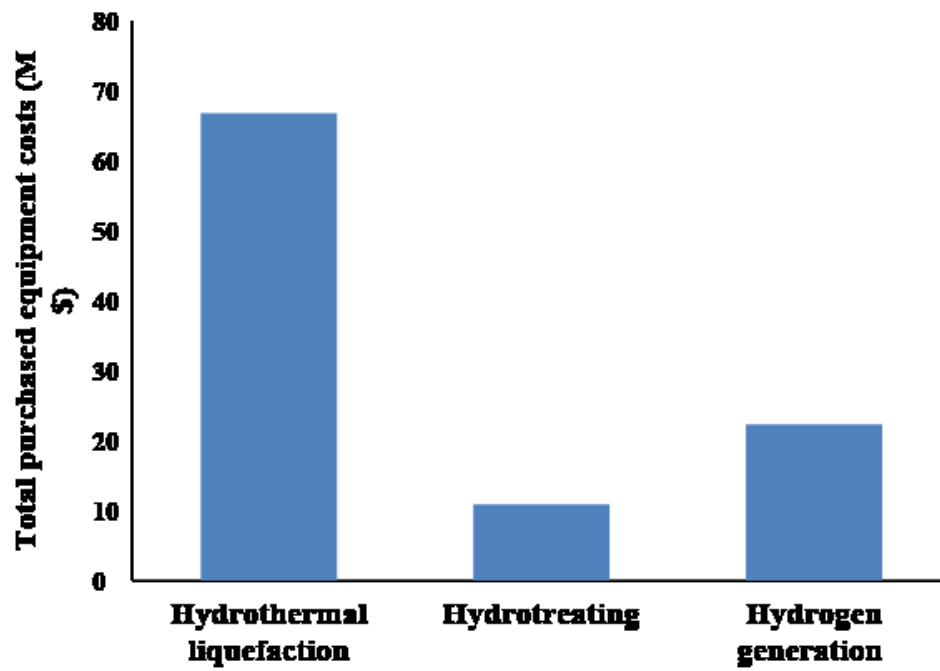




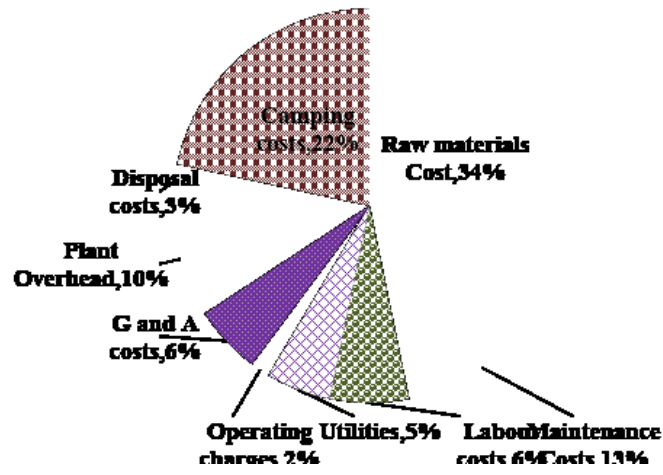
**Figure 1.** Schematic (a) and process flow model (b) for the production of diluent from biomass via hydrothermal liquefaction



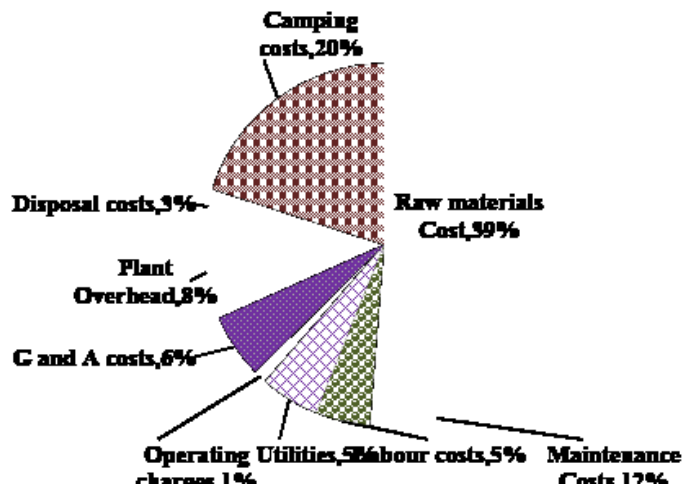
**Figure 2.** Percentage breakdown of direct (a) and indirect (b) costs for scenario 1 (hydrogen production) and scenario 2 (hydrogen purchase)



**Figure 3.** Proportion of purchased equipment cost for hydrothermal liquefaction, hydrotreating, and the hydrogen generation system

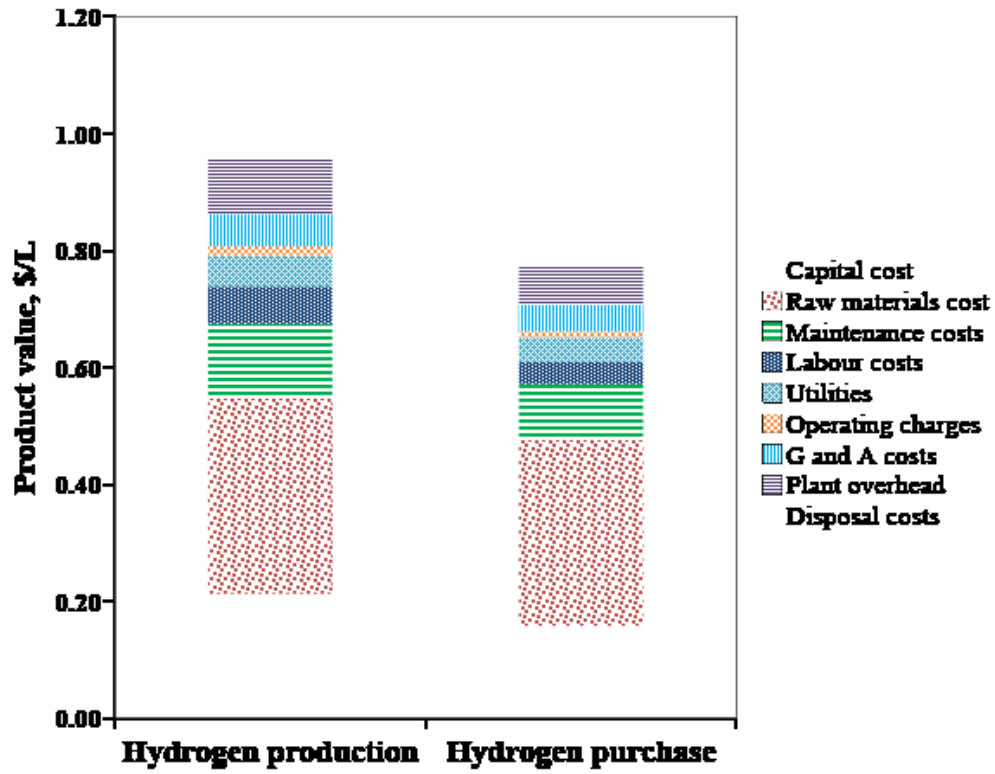


(a)

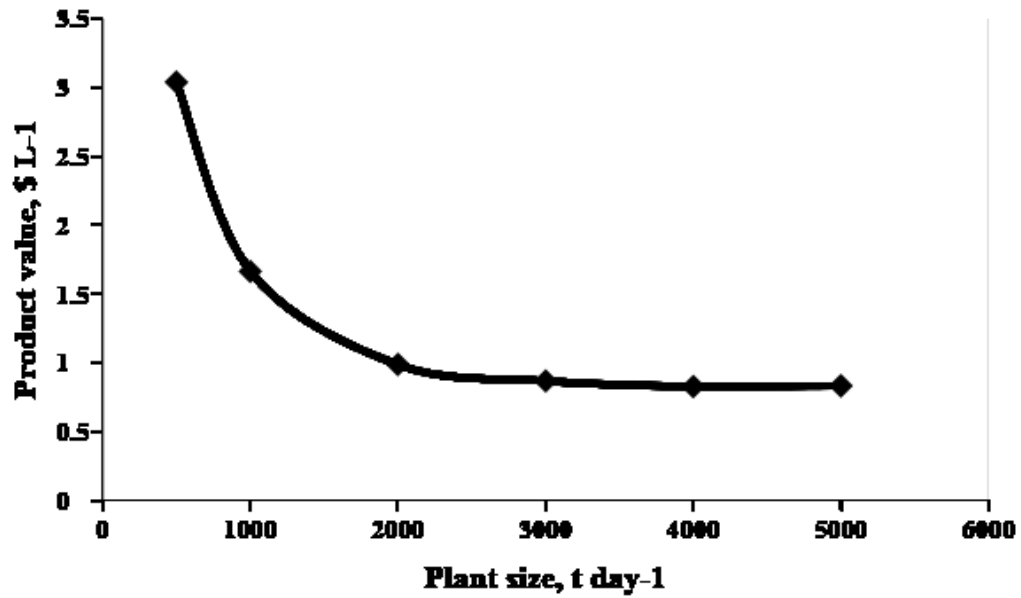


(b)

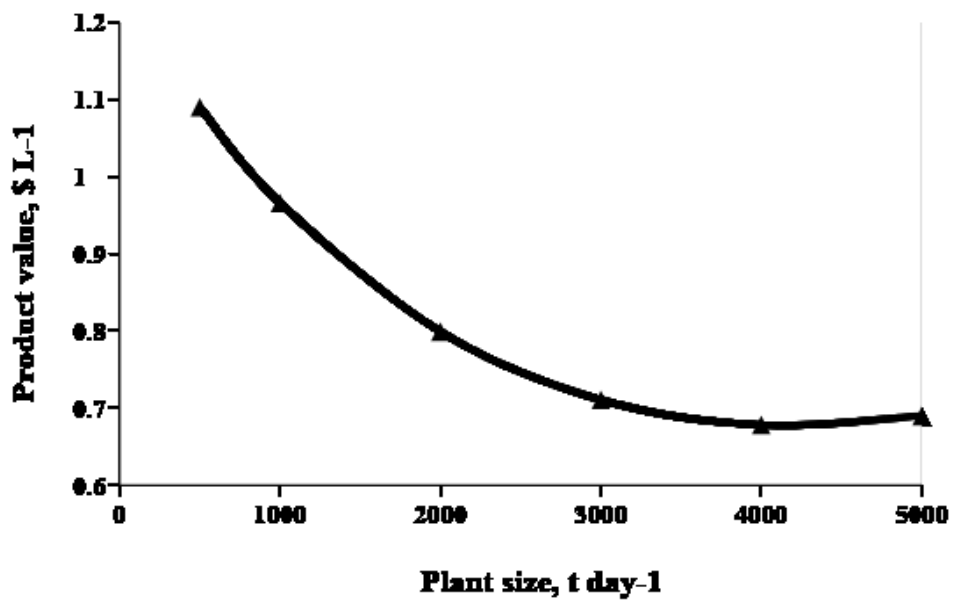
**Figure 4.** Breakdown of operating costs in the (a) hydrogen production and (b) hydrogen purchase scenarios



**Figure 5.** Contribution of various operating costs to the product value in the hydrogen production and the hydrogen purchase scenarios

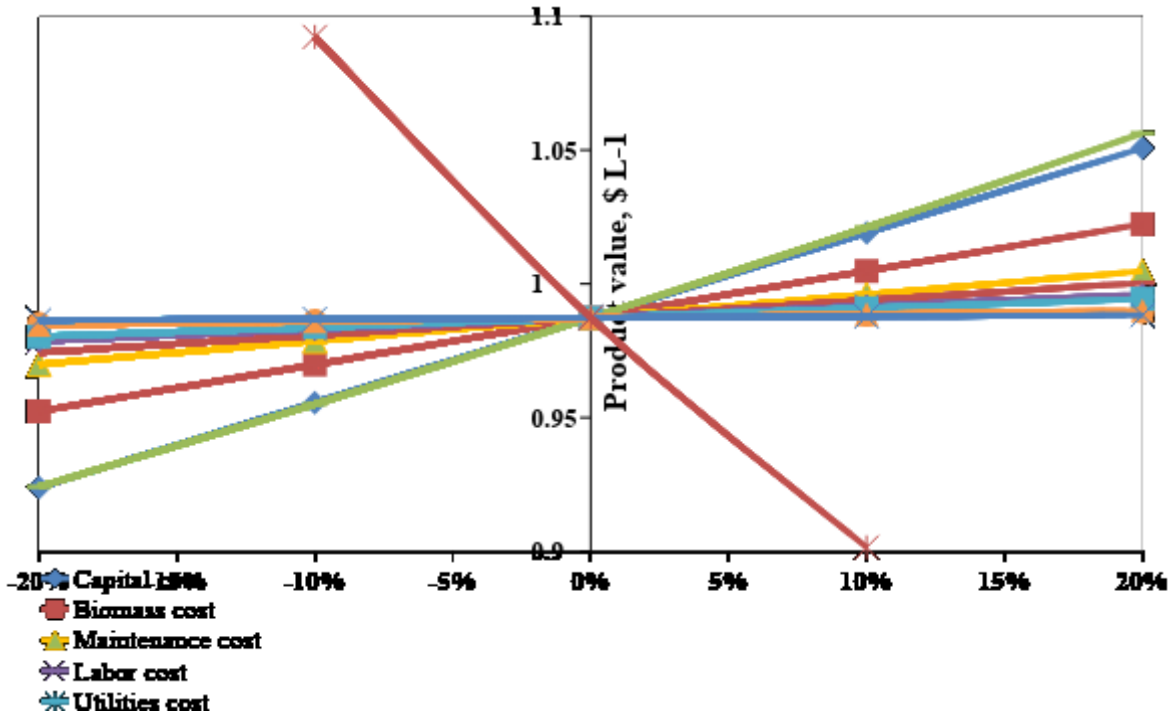


(a)

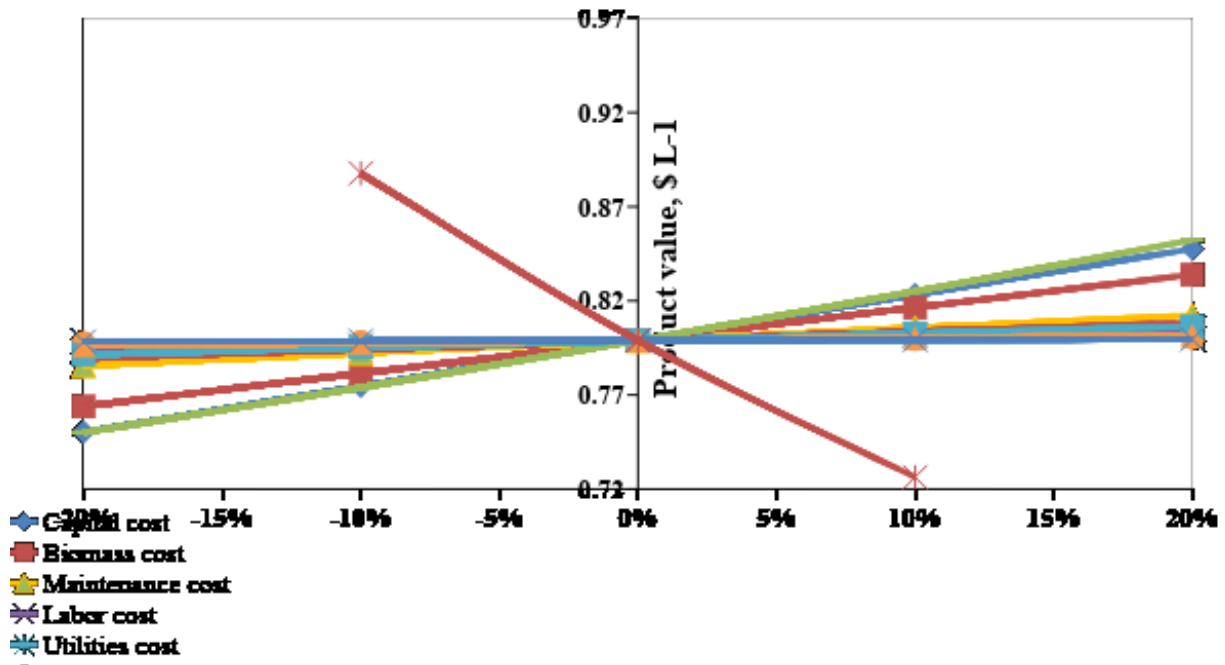


(b)

**Figure 6.** Variations in product value with increasing plant capacity in hydrogen production (a) and purchase (b) scenarios

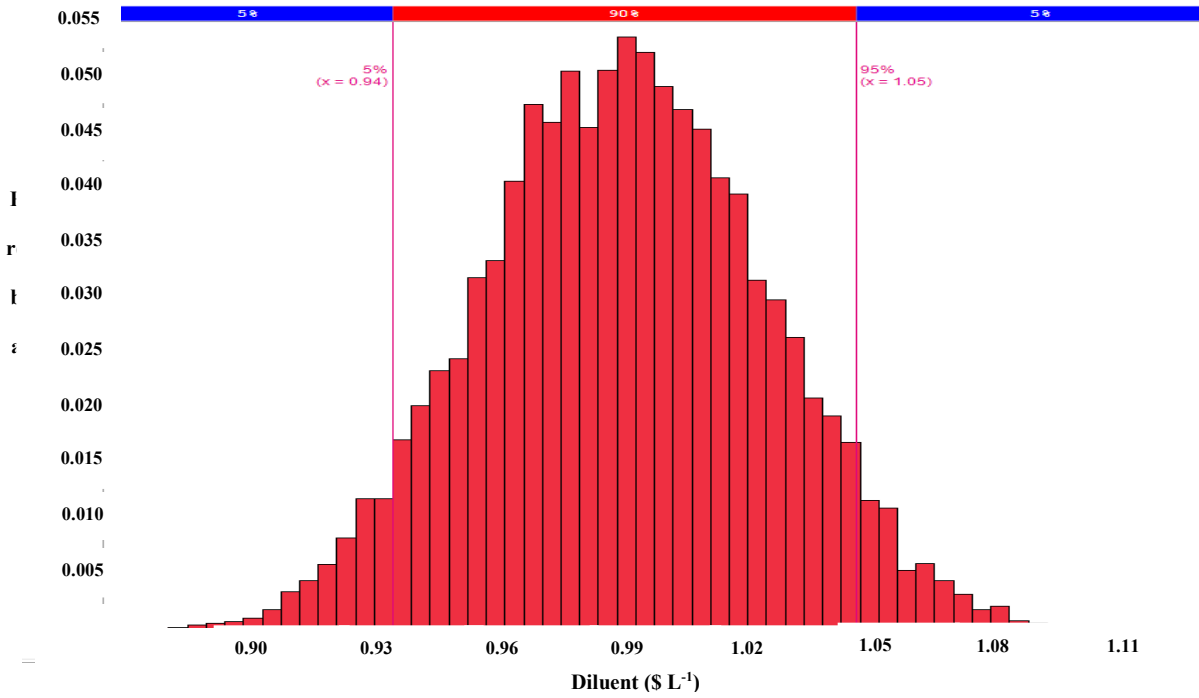


(a)

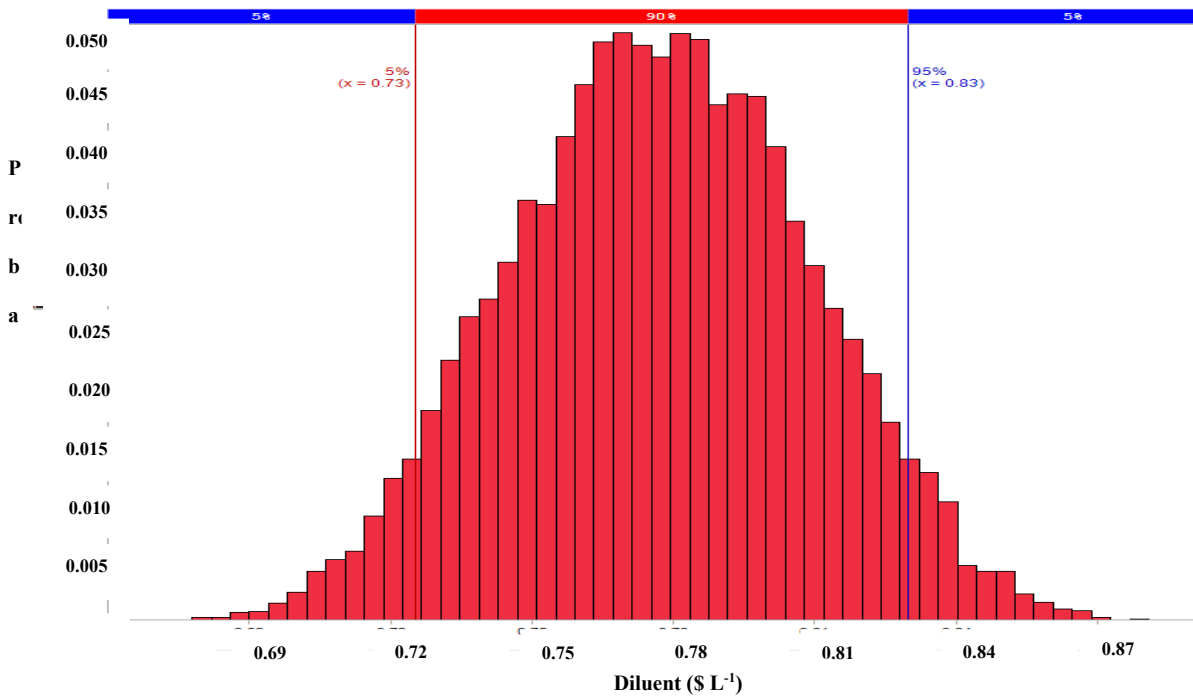


(b)

**Figure 7.** Sensitivity analysis of various parameters in the (a) hydrogen production and (b) hydrogen purchase scenarios



(a)



(b)

**Figure 8.** Product value uncertainty analyses for the (a) hydrogen production and (b) hydrogen purchase scenarios



**Table 1.** Properties of known diluents and stabilized bio-crude through hydrothermal liquefaction

<b>Product</b>	<b>Bio-crude<sup>70</sup></b>	<b>Bio-crude<sup>71</sup></b>	<b>Diluent<sup>58,69,72-74,76</sup></b>	<b>Naphtha<sup>58,73,75,77-81</sup></b>
Specific gravity	0.761	0.7747	0.675-0.861	0.7156-0.790
TAN (mg KOH g oil <sup>-1</sup> )	< 0.01	n/a	n/a	n/a
Viscosity (cP)	1.29	1.96	< 0.5 – 4.56	0.4482
Product, wt%				
C	83.4	84.2	83.4-87.2	84.0-87.25
H	13.5	13.9	10.6-12.8	11.20-16.0
O	0.07	1.7	0-0.4	0.03-0.22
N	< 0.05	0.1	0.0-1.6	0.23-0.94
S	< 0.005	0.0063	0.03-0.17	0.034-0.4

\*n/a: not available

**Table 2.** Proximate and ultimate analyses of the feedstock considered in this study

<b>Parameters</b>	<b>Feedstock</b>
Feedstock type	Wood chips
Initial moisture content, %	50
Particle size (mm)	1.5 - 2
Proximate analysis, wt% dry basis	
Fixed carbon	18.5

Volatile matter	80
Ash	1.5
Ultimate analysis, wt% dry basis	
C	50.9
H	6.0
N	0.3
S	0.03
O	41.3

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**Table 3.** HTL product characterization and process assumptions

Biomass flow rate, dry t day <sup>-1</sup>	2000
Biomass % (dry w w <sup>-1</sup> )	8.2
<b><i>Hydrothermal liquefaction</i></b> <sup>86</sup>	
Temperature, °C	350
Pressure, MPa	20.3
<b><i>Hydrotreating (Single-step)</i></b> <sup>87</sup>	
Temperature, °C	400
Pressure, MPa	10.5

H <sub>2</sub> , g g <sup>-1</sup> dry bio-oil	0.043
<i>PSA offgas recovery (%)</i> <sup>49</sup>	80

**Table 4.** Plant capital cost calculation factors

<b>Project investment cost factor estimates (in 2016 US dollars)</b>	
Installation factor	3.02
Total installed cost (TIC)	302% of TPEC
Indirect cost (IC)	89% of TPEC
Total direct and indirect costs (TDIC)	TIC + IC
Contingency	20% of TDIC

Fixed capital investment (FCI)	TDIC + contingency
Location cost	10% of FCI
Total project investment (TPI)	FCI + location cost

**Table 5.** Assumptions during economic analysis

<b>Items</b>	<b>Values</b>
Plant life (yr)	20
Cost year basis	2016
Capital cost distribution <sup>99,100</sup>	
Year 1 (%)	20
Year 2 (%)	35
Year 3 (%)	45
Production plant capacity factor <sup>99,100</sup>	

Year 1	0.7
Year 2	0.8
Year 3 and beyond	0.85
Internal rate of return (%)	10
Maintenance cost (\$)	3% of TPI
Operating charges (\$)	25% of operating labor cost
Plant overhead (\$)	50% of total operating labor and maintenance cost
Subtotal operating cost, SOC (\$)	Sum of all operating costs including raw material and utility cost
G & A cost (\$)	8% of SOC
Ash disposal cost <sup>40</sup> (\$ t <sup>-1</sup> )	43
Wastewater disposal cost <sup>40</sup> (\$ t <sup>-1</sup> )	0.63

**Table 6.** Main cost estimates for the hydrogen production and purchase scenarios

<b>Scenario</b>	<b>Hydrogen production</b>	<b>Hydrogen purchase</b>
Total purchased equipment cost (M \$)	108.4	83.1
Total project investment (M \$)	559.67	429.13
Operating cost (M \$)	129.5	107.12
Production cost (\$ L <sup>-1</sup> )	0.987	0.799

