**From laboratory scale to innovative spruce-based biorefinery. Note II: Preliminary techno-economic assessment**

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Abstract

This work shows a preliminary techno-economic assessment (TEA) of a biorefinery co-producing ethanol, pyrolysis oil and char from Norway spruce (*Picea abies*) via steam explosion and enzymatic saccharification followed by anaerobic fermentation of the sugars, and fast pyrolysis of the lignin-rich saccharification residue. The capacity of the modelled biorefinery was set to treat 100,000 tons of dry wood per year. The input for the assessment of the facility was retrieved from process simulations carried out in COFE v3.6 (AmsterCHEM), see Note I. Here, we propose different strategies to minimize the total costs of the biorefinery. The results indicate that the studied biorefinery process is economically feasible, while its profitability depends considerably on the source of hydrogen, oscillation in the market price of the generated products and utility costs.

**Keywords**: biorefinery, techno-economic assessment, energy integration, hydrogen, bio-oil and bioethanol production, biorefinery optimization

* 1. Introduction

The need for decarbonization drives a shift from fossil fuel- to bio-based economy. In this scenario, bio-products (bulk biochemicals such as methanol, ethanol, and biofuels) are expected to replace the conventional ones produced from crude oil (Cherubini et al., 2010). While the concept of biorefinery is not new, designing a process that is economically feasible at a commercial scale is a challenge, partly due to the low cost of fossil resources, the high O-content of biomass and the high complexity of operations (Bisotti et al., 2023; Gilardi et al., 2023). Despite the sustainability aspect, the marginal profitability, competition against highly optimized traditional oil refineries and suboptimal use of on-site resources exacerbate the struggle of biorefineries for competitiveness (Cherubini, 2010). Here, we provide a preliminary techno-economic assessment (TEA) of spruce-to-fuel biorefinery that is based on a combined biochemical and thermochemical conversion to produce ethanol and bio-oil-derived fuel fractions, as described in Note I. We adjusted the process layout to minimize the energy demand, waste generation and carbon footprint. Furthermore, we considered using alternative hydrogen sources (i.e., green, blue and grey). The TEA shows that a profitable implementation of the proposed biorefinery is realistic; however, the costs of utilities, raw materials and source of hydrogen deeply affect the breakeven point and pay-back time.

* 1. Improvements to the biorefinery baseline

To improve the feasibility of the spruce-to-fuel biorefinery process depicted in Figure 1, we introduced the following improvements (also highlighted in green in Figure 1).

A diagram of a process

Description automatically generated

**Figure 1**: Biorefinery block flow diagram (only main streams are reported).

***On-site enzyme production:*** Since enzymes (when purchased from external producers) can make up about one-third of the total operating costs, we included enzyme production on-site based on the setup by Humbird et al. (2011) using cheap raw materials, e.g., corn steep liquor (CSL, C-source), ammonia (N-source), SO2 (S-source), and a cellulase-producing strain.

***Fermentation:*** Regarding bioethanol production, the fermenting strain was added together in solution with corn steep liquor (CSL) and diammonium phosphate (DAP) as suggested by Humbird et al. (2011).

***Heat integration:*** Biochar, the side product of pyrolysis, has many potential applications. As the market is not stable, we chose to recover its high heating value by burning it. We compared oxy-combustion, which requires pure oxygen, and air combustion as alternatives. For the oxy-combustion, we adjusted the CO2 recycle to achieve the desired flame temperature (1100°C). To achieve the same temperature in the air combustion, we regulated the airflow to the combustor. The resulting hot flue gas supplies thermal heat for the pyrolysis reactions and, subsequently, to a steam generation unit. This unit generates steam by burning the light tail gas recovered from the top of the distillation column fractionating the upgraded crude bio-oil. The total produced steam could cover the steam demand for biomass pretreatment (i.e., steam explosion) completely and ethanol purification partially. The difference was covered with natural gas combustion.

* 1. Materials and methods

After closing the energy and material balances in Note I, here, we prepared a techno-economic assessment (TEA) for the spruce-to-fuel biorefinery sketched in Figure 1 (process design) and Note I (operating conditions). For that, we made the following assumptions (Section 3.1) and analysed the impact of the hydrogen source on process feasibility and profitability (Section 3.2).

* 1. *Assumptions*
     1. *Capital Investments (CAPEX)*

The CAPEX was estimated based on Guthrie’s work (1978) and updated to 2021 using the Chemical Engineering Plant Cost Index (CEPCI). This year was selected to evaluate feasibility under stable market conditions as it predates the war in Ukraine in 2022, which created turmoil for the economic indexes. The purchase base cost of each piece of equipment was turned into the corresponding bare module cost, where expenditure for materials, actual operating pressure, and installation were included. The size of reactors was based on the residence time and inlet volumetric flow. The diameter and height of separation columns and the number of trays were estimated for achieving adequate vapor–liquid contact (i.e., 70% flooding conditions) and global separation efficiency (70% efficiency). The reactors were designed as simple vessels since correlations for tube bundles are not available. The analysis neglects the costs of filters and cyclones. Furthermore, we assumed that combustion technology (air or oxy-combustion) does not affect investment costs for the waste heat boiler. This assumption considers that compact volumes for oxy-combustion compensate for the need for more expensive materials. The final value for the CAPEX reflects the grass-roots cost, meaning that the biorefinery is built as a greenfield plant. The lifetime of the plant was assumed to be 20 years.

* + 1. *Operating costs (OPEX)*

The OPEX (Costs of Manufacturing – COM) were calculated based on Turton’s work (2018), as given in Equation 1, where FCI is the total investment cost, COL is the operating labour cost, CUT, CRM and CWT are the total utility, raw material and waste treatment costs. For simplicity, the CWT term was neglected.

|  |  |
| --- | --- |
|  | (1) |

The costs of the consumables and utilities were retrieved from prior literature for hydrogen (Arcos and Santos, 2023) and part of the consumables (Bbosa et al., 2018). The remaining values were retrieved from databases (https://ceskdata.com/).

* + 1. *Cash flow and internal rate of return (IRR)*

For the discounted cash flow, we assumed a minimum acceptable target value of 12% for the internal rate of return (IRR) and 30% taxation on the revenues. IRR is a discount rate that makes the net present value (NPV) of all cash flows equal to zero.

* 1. *Sensitivity analysis*

A sensitivity analysis was performed to gain insight into the biorefinery’s performance under various case studies (CS). We defined the process with air combustion of biochar as the baseline scenario (CS1). For biochar combustion, either air or oxy-combustion was considered. For oxy-combustion, the pure O2 was either co-produced with green-H2, via water splitting with electrolysis, or obtained from an Air Separation Unit (ASU) when using grey- or blue-H2. When using green-H2, oxy-combustion was considered only since oxygen is a vented by-product of green-H2 and hence is free (0 $/kgO2). For the cases CS1 to CS5, the calculated profitability indexes were the IRR and payback time, considering a lifetime of 20 years for the biorefinery:

* CS1: baseline biorefinery (with air combustion and grey-H2)
* CS2: biorefinery with oxy-combustion (O2 from ASU) and grey-H2
* CS3: biorefinery with air combustion and blue-H2
* CS4: biorefinery with oxy-combustion (O2 from ASU) and blue-H2
* CS5: biorefinery with oxy-combustion and green-H2

Based on the biorefinery layout of CS5, we assessed the impact of the bioethanol selling price on the overall economics. CS6 to CS8 were set up to find the minimum ethanol selling price (MESP) to achieve an IRR of 12%.

* CS6: MESP for the biorefinery as in CS5
* CS7: MESP for a biorefinery as in CS5, but when green-H2 price equals blue-H2 one
* CS8: MESP for a biorefinery as in CS5, but when green-H2 price equals grey-H2 one

Finally, we assessed the impact of using green-H2 on the economics (CS9 to CS11). These last cases were designed to determine the maximum affordable green-H2 price (MGHP) considering market price oscillations and different countries, e.g., 0.60 in the US and 1.20 $/litre in Sweden; thus, 0.80 $/litre is a time-weighted average worldwide. The target value for the IRR was kept fixed at 12%.

* CS9: MGHP at the current market price of bioethanol (0.80 $/litre, fuel-grade spec)
* CS10: MGHP when bioethanol price is set at 0.60 $/litre (i.e., US price)
* CS11: MGHP when bioethanol price is set at 1.20 $/litre (i.e., Sweden price)
  1. Results

Table 1 highlights the contributions of the single-unit operations to the investment cost (CAPEX) of the biorefinery. Table 2 gathers the prices of raw materials, utilities and end products used for the calculation of operating costs and revenues reported in Table 3.

**Table 1**: CAPEX for biorefinery plant: characteristic size, base purchase cost, and bare module cost of each piece of equipment. Costs are reported in millions of USD (M$).

| **Unit** | **Size** | **Purchase cost** | **Bare module cost** |
| --- | --- | --- | --- |
| Steam explosion reactor chamber | 8 m3 | 0.02 | 0.60 |
| Saccharification reactor chamber | 600.3 m3 | 0.67 | 2.71 |
| Pyrolysis chamber | 455.6 m3 | 0.50 | 2.04 |
| Fermenter (all units) | 689 m3 | 1.64 | 6.67 |
| HDO mild | 91.8 m3 | 0.11 | 0.45 |
| HDO severe | 15.7 m3 | 0.03 | 0.11 |
| Enzyme production | 524.5 m3 | 0.58 | 2.35 |
| Ethanol purification columns (total) | 37.0 m3 | 0.07 | 0.27 |
| Bio-oil distillation columns (total) | 14.5 m3 | 0.02 | 0.10 |
| Steam boiler | 6290 kW | 3.83 | 9.38 |
| Char combustion boiler | 5160 kW | 2.85 | 6.97 |
| Compressors (all 4 units) | 650 kW | 0.58 | 1.25 |
| Heat recovery exchanger-1 | 85.1 m2 | 0.16 | 0.34 |
| Heat recovery exchanger-2 | 258.2 m2 | 0.28 | 0.60 |
| Heat recovery exchanger-3 | 113.1 m2 | 0.18 | 0.39 |
| Water coolers (all 8 units) | 276.7 m2 | 1.12 | 3.69 |

**Table 2**: Costs of consumables (light blue), utilities (red) and products (grey background).

|  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- |
| **Chemical** | **Price ($/ton)** | **Value (M$/year)** | **Chemical** | **Price** | **Value**  **(M$/year)** |
| Spruce chips | 43.23 | 4.34 | Oxygen (ASU) | 2.50 $/kg | 7.02 |
| 2-naphthol | 3500 | 6.25 | Grey H2  Blue H2  Green H2 | 1.50 $/kg  3.50 $/kg  7.25 $/kg | 2.64  6.16  12.76 |
| NH3 | 410 | 1.48 | Natural gas for steam generation | 20 $/MWh | 0.99 (oxy-comb.)  1.09 (air comb.) |
| SO2 | 276 | 1.46 | Electricity | 80 $/MWh | 0.77 |
| Sodium acetate | 20 | 0.99 | Cooling water | 0.02 $/m3 | 0.08 |
| Fermen strain | 100 | 0.45 | Bioethanol | 0.80 $/litre | 17.1 |
| DAP | 895 | 1.44 | Gasoline | 2.10 $/litre | 17.9 |
| CSL | 50 | 1.10 | Light diesel | 1.72 $/litre | 25.9 |
| Ethylene glycol | 860 | 0.01 | Heavy diesel | 0.37 $/litre | 2.0 |

In addition to CAPEX, OPEX and revenues, Table 3 indicates the profitability indexes, i.e., the internal rate of return (IRR), payback time (PBT) and MESP/MGHP, resulting from the sensitivity analysis described in Section 3.2. For PBT, “Never” means that the investment cannot be recovered by 20 years (i.e., the assumed biorefinery’s lifetime).

**Table 3**: Results of the sensitivity analysis. Colored cell(s) report the main outcome(s) of the corresponding case study (CS). Different colors cluster the cases as reported in Section 3.2.

|  |  |  |  |  |  |  |  |
| --- | --- | --- | --- | --- | --- | --- | --- |
| **Case** | **IRR**  **(%)** | **MESP**  **($/litre)** | **MGHP**  **($/kgH2)** | **CAPEX**  **(M$)** | **OPEX**  **(M$)** | **Revenues**  **(M$)** | **PBT**  **(year)** |
| CS1 | 17.5 | 0.801 | 1.501 | 71.62 | 49.24 | 67.90 | 9.5 |
| CS2 | 9.40 | 0.801 | 1.501 | 71.62 | 57.75 | 67.90 | Never |
| CS3 | 12.7 | 0.801 | 3.501 | 71.62 | 53.57 | 67.90 | 17 |
| CS4 | 3.50 | 0.801 | 3.501 | 71.62 | 62.08 | 67.90 | Never |
| CS5 | 4.40 | 0.801 | 7.251 | 71.62 | 61.45 | 67.90 | Never |
| CS6 | 12.0 | 1.06 | 7.25 | 71.62 | 61.45 | 75.15 | 20 (fixed) |
| CS7 | 12.0 | 0.77 | 3.50 | 71.62 | 61.45 | 67.14 | 20 (fixed) |
| CS8 | 12.0 | 0.62 | 1.50 | 71.62 | 61.45 | 62.81 | 20 (fixed) |
| CS9 | 12.0 | 0.80 | 3.85 | 71.62 | 54.20 | 67.90 | 20 (fixed) |
| CS10 | 12.0 | 1.20 | 8.93 | 71.62 | 65.20 | 78.90 | 20 (fixed) |
| CS11 | 12.0 | 0.60 | 1.31 | 71.62 | 48.69 | 62.39 | 20 (fixed) |

1 Bioethanol and hydrogen average prices as in Table 2.

* 1. Conclusions

Here we report the TEA of a spruce-to-fuel biorefinery. Our sensitivity analysis estimates the biorefinery’s economics when accounting for the hydrogen source, strategies to burn the formed biochar (CS1 to CS5), the selling price of the fuel-grade bioethanol (CS6 to CS8) and the price for green-H2 (CS9 to CS11). Our calculations indicate that, with the current market price values, only CS1 and CS3 (using air for biochar combustion and grey- or blue-H2 for HDO of the crude bio-oil) are economically feasible for the proposed cases. The price of bioethanol (product) and hydrogen (consumable) strongly affects the profitability of the proposed biorefinery. In CS6 to CS11, we focused on green-H2 because of the growing importance of renewable energy with a reduced carbon footprint and of current regulations and fuel blending mandates requiring "green" alternative fuels with the lowest carbon footprint. If the price of green-H2 decreases from the current 7.25 (CS6) to 3.50 (blue-H2; CS7) or 1.50 $/kgH2 (grey-H2; CS8), the MESP will decrease correspondingly. CS6 reveals that the MESP should be at least 1.06 $/litre (against the current average of 0.80 $/litre) to make the biorefinery profitable when using green-H2. However, if the MESP rises to 1.20 $/litre (CS10), the biorefinery can reach a break-even point and it can tolerate a much higher MGHP of 8.93 $/kgH2. CS6 indicates that incentives on bioethanol of at least 0.26 $/litre, i.e., 33% of the current bioethanol average price (0.80 $/litre), or even higher, e.g., 0.40 $/litre as assumed in CS10 (Sweden case), can support the “greenest” option. At the current average selling price of bioethanol, the MGHP is expected to almost halve to get a profitable green-H2-basedprocess as shown in CS9. In the worst scenario (i.e., the US average price of 0.60 $/litre; CS11), the MGHP must drop to 1.31 $/kgH2, i.e., below the current price of grey-H2 (1.50 $/kgH2). Although to date this case is not feasible, technological developments will likely make the production of green-H2 competitive with that of grey-H2 in the next decade. The carbon footprint of the proposed case studies and the incorporation of a carbon capture system will be done in the next studies. Further strategies to reduce the costs, such as internal recycling of phenols as cation scavengers for feedstock pretreatment, will be assessed.

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