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# Techno-Economic Assessment of Chemical Looping Gasification of Biomass for Fischer–Tropsch Crude Production with Net-Negative $CO_2$ Emissions: Part 2

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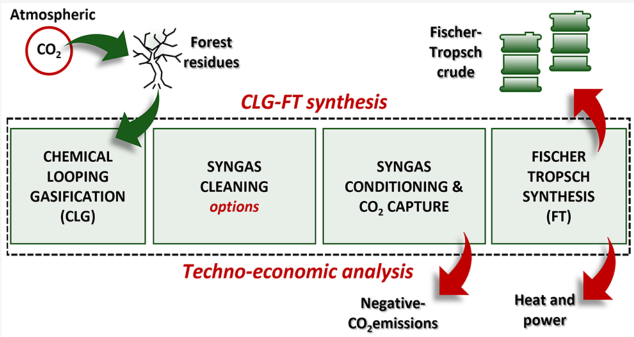
**ABSTRACT:** This work presents a techno-economic analysis of a novel gasification system, chemical looping gasification (CLG), used as the primary gasification process for biofuel production through Fischer–Tropsch synthesis (FTS). Two different gas cleaning process configurations, cold-gas cleanup and hot-gas cleanup process trains, are explored, along with off-gas utilization possibilities, to study their influence on the process economics of an integrated CLG–FT process plant. Off-gas recirculation to increase Fischer–Tropsch (FT) crude production has a significant influence on reducing the levelized production costs for FT crude. The results indicate that the specific production cost estimated for a CLG–FT plant with a hot-gas cleanup train is roughly 10% lower than the case with a cold-gas cleanup train, while the total plant costs remain relatively the same for all plant configurations. In addition to this, the former has a considerably higher overall system energy efficiency of 63%, roughly 18% more than the latter, considering the co-production of FT crude, district heating, and electricity. The specific investment costs range from 1.5 to 1.7 $M€_{2018}/MW_{LHV}$, and the specific FT crude production cost ranges from 120 to 147 $€_{2018}/MWh_{FT}$. Roughly 60% of total carbon fed to the process is captured, enabling net-negative $CO_2$ emissions. A $CO_2$ price for negative emissions would significantly reduce the specific fuel production costs and would, hence, be competitive with fossil-based liquid fuels.




## 1. INTRODUCTION

Growing transportation sectors, such as the aviation and maritime sectors, are considered to be one of the most challenging sectors to decarbonize as these sectors are heavily locked-in to petroleum products and electrification is still technologically challenging. $CO_2$ emissions from the aviation sector, including commercial and freight operations, were estimated to be around 918 million tonnes in 2018,[1] which had peaked at 1027 million tonnes in 2019[2] prior to the pandemic, accounting for roughly 3% of the total global $CO_2$ emissions, while the $CO_2$ emissions from international shipping were roughly 707 million tonnes in 2019, which accounts for the majority of the emissions from the maritime sector.[3,4] A wide array of suggested measures, such as efficiency improvements, policy measures, and low-carbon fuels, are required to meet the sustainable development scenario.[3] Sustainable biofuels, derived from biomass, mainly from agricultural and forest waste streams, are one of the promising pathways to decarbonize these sectors. It requires little to no changes to the current fuel distribution network and could directly contribute to the replacement of conventional transport fuels, such as diesel and gasoline. In addition to this, biomass through waste streams avoids uncertainties, such as the impact of indirect land-use change, associated with the production of biofuel using food crops. These waste streams could be used to produce advanced biofuels, such as synthetic natural gas (SNG), or liquid biofuels, such as methanol and Fischer–Tropsch (FT) crude from the Fischer–Tropsch synthesis (FTS) process. Figure 1 shows a simple schematic of the chemical looping gasification (CLG) process that uses metal oxide particles that oxidize with air and reduce the fuel in the two interconnected circulating fluidized bed reactors, namely, the air reactor (AR) and the fuel reactor (FR), thereby restricting the dilution of the product gas from the FR with nitrogen from the air. In addition to this, $CO_2$ is expected to remain in the product gas at concentrations higher than indirect gasification (IG), enabling lower costs associated with $CO_2$ separation from the product gas. Thus, lower volumes of diluents and tar in the raw syngas can, therefore, reduce costs

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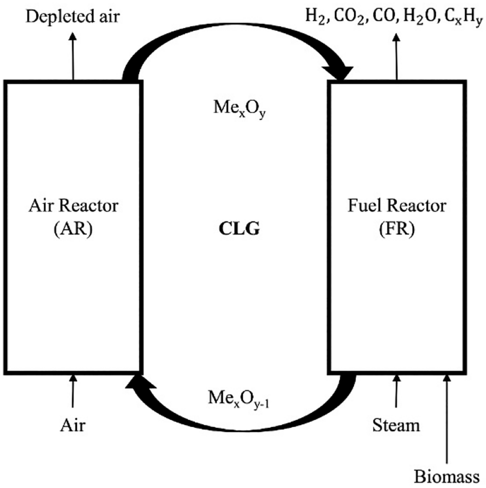


**Figure 1.** Schematic of CLG of biomass. This figure was reproduced from part 1 of this work[5] (10.1021/acs.energyfuels.2c00819). Copyright 2022 American Chemical Society.

associated with gas purification to meet the minimum required specifications for the synthesis of biofuels.

This study aims to conduct a techno-economic assessment of a CLG plant that uses forest residues as the primary feedstock with a FTS plant for conversion of syngas to FT crude. The analysis includes a comparison between two different gas cleaning process configurations, i.e., the cold-gas cleanup and hot-gas cleanup process trains. In addition to this, valorization of hydrocarbon off-gas from the FT synthesis plant for an additional heat supply is explored, and heat and power production potential are examined. The economic analysis accounts for the net-negative $CO_2$ emissions achieved from the integrated CLG−FT process. The analysis in this study is based on Aspen Plus simulations of the CLG process, modeled in part 1[5] of this work (10.1021/acs.energyfuels.2c00819), which was validated with experiments[6] performed with Linz−Donawitz (LD) slag as the primary oxygen carrier. Furthermore, sensitivity analysis is performed on the economic assessment with parameters that significantly influence the overall plant economics. Finally, the various trade-offs in each suggested process configuration are identified and compared.

## 2. BACKGROUND

**2.1. Model Description.** The techno-economic assessment is based on the CLG−FT process models developed in Aspen Plus, described in part 1[5] of this study (10.1021/acs.energyfuels.2c00819), which includes all modeling details and other associated process parameters. Figure 2 shows a simple schematic of two modeled cases that excludes the feedstock pretreatment and FT crude upgrading step. The gas cleanup process train is the major difference between the two process configurations, i.e., cases CG and HG. Case CG employs a more conventional and mature process cleaning train that includes an amine-based scrubber for the acid gas removal (AGR) stage, whereas case HG employs a sour-gas shift, followed by AGR and $CO_2$ capture through a single Rectisol[7] unit, prior to the FT reactor.

**2.2. Case Description.** In both the cases, shown in Figure 2, the purge or off-gas emanating from the FT reactor is recycled back to the process after the hydrocarbon (HC) recovery block. However, this off-gas could be combusted to supply additional heat for steam superheating in heat recovery steam generators (HRSGs). A marginal penalty on the negative emissions from these processes is expected, resulting in lower negative emissions when compared to cases where it is recycled back to the process. This has been accounted for in the economic assessment. Cases with off-gas combustion, expectedly, tend to have lower chemical efficiency or biomass-to-FT crude production efficiency, as defined in part 1[5] of this work (10.1021/acs.energyfuels.2c00819), compared to its counterpart with off-gas recycling. The investigated cases in the techno-economic assessment are listed in Table 1, where the CLG−FT models estimate the highest chemical efficiency

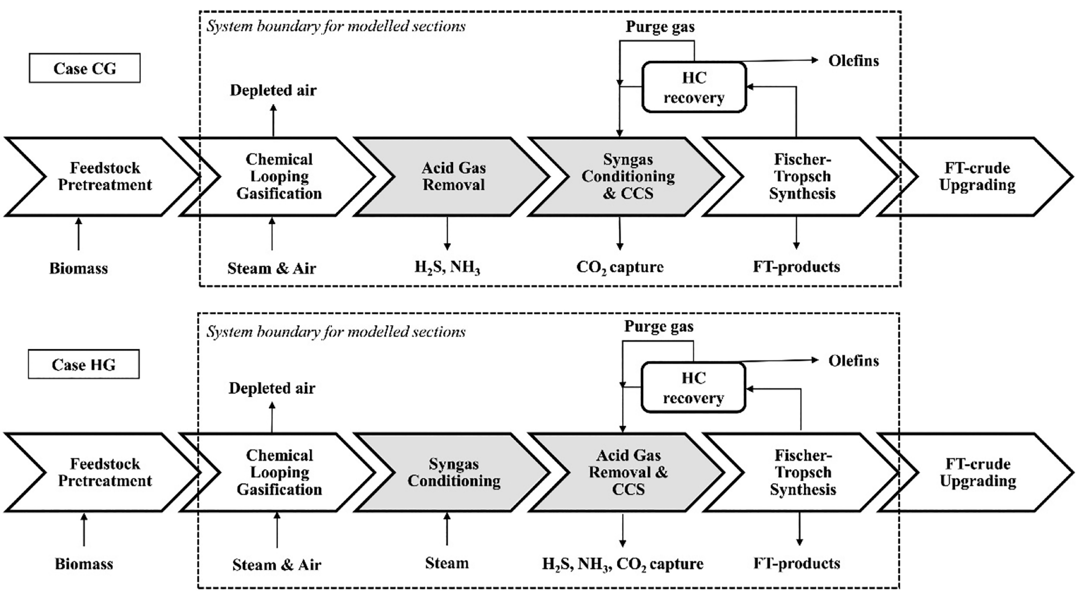


**Figure 2.** Simple schematic of cases CG and HG in the CLG−FT plant that excludes the pretreatment and product upgrading steps; the shaded stage indicates the differences between the two cases. This figure was reproduced from part 1 of this work[5] (10.1021/acs.energyfuels.2c00819). Copyright 2022 American Chemical Society.

for case CG1 (40.3%), followed by HG1 (38%), CG2 (32.3%), and HG2 (29.2%).

**Table 1. Cases Investigated in the Heat Recovery and Economic Assessment**

| cases based on gas cleaning process train | cases investigated | description |
|---|---|---|
| case CG | CG1 | off-gas recycled |
| | CG2 | off-gas combusted |
| case HG | HG1 | off-gas recycled |
| | HG2 | off-gas combusted |

## 3. PROCESS INTEGRATION

The CLG−FT process train contains multiple process heating and cooling demands at different temperature levels. Pinch analysis tools are applied to estimate the heat and power integration opportunities from the overall CLG−FT process chain, using the stream data extracted from the simulation models. An overall minimum temperature difference ($\Delta T_{min}$) of 10 °C is chosen for all cases studied in this work. The overall system energy efficiency or the thermal efficiency [lower heating value (LHV) basis] of the CLG−FT process, shown in eq 1, is defined as the ratio of the useful energy output, i.e., the sum of the heat delivered as district heating, net electricity produced or consumed, and the FT crude produced from the process to its required energy input, i.e., biomass fuel fed to the CLG plant.

$$\eta_{sys} = \frac{Q_{FT,LHV} + Q_{DH} \pm W_{el,net}}{Q_{biomass,LHV}} \quad (1)$$

Grand composite curves (GCCs) and foreground−background (FGBG) analysis or otherwise known as split GCC analysis are applied for energy targeting. The GCCs graphically represent the heat recovery potential of a process at different temperature levels, including the minimum heat and cold utility demand of the process.[8] In this work, FGBG analysis is applied to estimate the co-generation potential from the CLG−FT plant (background process) with the integration of a heat recovery steam cycle (foreground process). ProPI, a Microsoft Excel add-in developed at Chalmers University of Technology, is used for generating the GCCs for all of the cases. A heat recovery steam cycle (HRSC) is considered for integration with the CLG−FT plant for heat and power production. Steam generated in the HRSG is used in an extraction back-pressure turbine to satisfy the on-site steam demand and use the surplus heat available in the CLG−FT process. The steam conditions at the HRSG outlet are assumed to be at 100 bar pressure and 450 °C. The superheated steam temperature is limited to 450 °C, owing to corrosion issues associated with raw syngas cooling with steam, reported elsewhere.[9] Turbine isentropic efficiency is assumed to be 85%. Pump and generator efficiency is assumed to be 90 and 98%, respectively. The steam extraction pressure levels are adjusted accordingly to match the heat demands of the CLG−FT process and, subsequently, activate at least one pinch point between the two process GCCs. In all cases, the extraction back-pressure turbine, a low-pressure exhaust steam pressure of 1 bar, is considered, which is suitable for a district heating (DH) condenser. The DH condenser is assumed to have a return and supply temperature of 65 and 90 °C, respectively.

## 4. ECONOMIC ASSESSMENT

The economic analysis of the CLG−FT process is performed by calculating the levelized cost of fuel (LCOF) produced. The LCOF enables comparison to other process configurations and similar biomass-to-FT-crude production processes reported in the literature,[10−12] to some extent, depending upon the cost estimation methodology, economic assumptions, and scale of the plant. This work focuses on estimating the specific production cost of FT crude via CLG, which can be refined in an existing refinery to derive a wide range of biomass-based products, such as FT diesel, biokerosene, and bio-olefins. Thus, the costs associated with the upgradation of FT crude are not considered in this study primarily because product upgradation or refining requirements steps could vary widely depending upon the desired final product. The FT crude produced could either be refined to create biomass-based blendstocks that could be added to wide-ranging fossil-based refinery products to avoid risks associated with direct co-processing with crude oil in existing refineries.[13,14] Alternatively, the FT crude could be refined to maximize one specific fuel product with higher value, such as biomass-based aviation fuel. de Klerk[15] gives the example of the Sasol facility,[16] which relies on coal pyrolysis products to meet the minimum required specifications of aviation jet fuel, to emphasize that, although the FT-based refineries could be designed to maximize one specific biofuel product, it is rarely the case that refineries are designed to operate at such conditions. For these reasons, it was deemed reasonable to have biomass-based FT crude as the final product from the CLG−FT plant. Two different CLG−FT cases (cases CG and HG) with different gas cleaning process configurations are analyzed and compared to evaluate their influence on the overall plant economics. The LCOF produced is calculated as per eq 2, which is defined as a ratio of the sum of the annualized total capital costs ($TPC_{annualized}$), annual fixed operation and maintenance costs ($c_{O\&M,fix}$), and various operational and maintenance costs, such as the consumables excluding feedstock ($c_{O\&M,consumables}$) and annual feedstock costs ($c_{O\&M,feedstock}$), annual district heating production ($c_{net,DH}$), and annual electricity consumption ($c_{net,EL}$) to the annual production of FT crude (MWh) from the CLG−FT plant. Note that, in eq 2, the revenue from net annual production of district heat (M€) is taken as negative and the cost of net annual electricity consumption (M€) is taken as positive, because the plants in all cases are electricity-deficit, which will be discussed further in Section 5.1.

$$\text{LCOF } (\text{€/MWh}) = ((\text{TPC} \times \text{CCF}) + (c_{O\&M,fix} + c_{O\&M,consumables} + c_{O\&M,feedstock} + c_{net,EL} - c_{net,DH}))/(Q_{FT} \times \text{FLH}) \quad (2)$$

The capital charge factor (CCF), used in eq 2, is a factor used to represent the total plant costs (TPCs) distributed over the plant lifetime. The CCF is estimated here based on the cost estimation methodology provided by the National Energy Technology Laboratory (NETL)[17] for power plant performances with a set of global economic assumptions on financial and tax structure, loan, and debt interest rates that enables comparison between newly suggested plants and conventional plants. With assumptions relevant to Sweden, the CCF is estimated to be around 10.4%/year for this plant, with an effective tax rate of 25.7% and an assumed plant economic lifetime of 20 years. The estimated CCF lies in the lower range of CCF chosen for similar biomass-to-liquid fuel production


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plants, where it typically ranges from 10 to 15%.[10,18] Here, it is re-emphasized that techno-economic evaluations performed in this work are performed with conditions applicable to Sweden, and the economic assumptions used are listed in Table 2.

**Table 2. Basic Assumptions for Economic Calculations**

| parameter | value | unit |
|---|---|---|
| base year | 2018 | |
| CEPCI[19] | 603.1 | |
| U.S. Dollar-to-Euro conversion rate | 1.08 | US$/€ |
| SEK-to-Euro conversion rate | 10.59 | SEK/€ |
| forest residues LHV (moisture content of 6%) | 19.34 | MJ/kg |
| cost of ingoing fuel (forest residues, Sweden)[20] | 170 | SEK/MWh |
| availability of operating plant | 8000 | h/year |
| plant lifetime[20] | 20 | years |
| capacity during operation[20] | 100 | % |
| electricity price[10] | 50 | €/MWh |
| district heating[10] | 30 | €/MWh |
| location | Sweden | |

**4.1. Cost Estimation Methodology.** The scale of production is believed to favor a typical biomass-to-liquid fuel production plant; however, the availability of biomass significantly limits the maximum size of the plant. It was previously reported that a standalone advanced biofuel plant sized to 200–300 $MW_{th}$ of total biomass consumption is a feasible scale.[10,20] In this study, a plant capacity of 100 $MW_{th}$ LHV of biomass input is considered. It is assumed that the plant receives dried forest residue feedstock (with 6 wt % moisture), and there is no feedstock pretreatment and drying utility on site. As a result of the expected excess heat-generation on site, it is expected that there would be several opportunities with respect to drying feedstock on site. All cost estimates are made for a *n*th plant design. The cost estimation methodology of the total plant costs, shown in Table 3, is an established bottom-up approach (BUA), adapted from the European Benchmarking Task Force (EBTF) report on carbon capture and storage (CCS) in power plants.[21] This method was also applied in similar techno-economic studies[18,22] that evaluated different conceptual plant processes.

**Table 3. Cost Estimation Methodology**

| TPC calculation methodology | |
|---|---|
| plant equipment | cost |
| equipment a | A |
| equipment b | B |
| bare erected cost (BEC) | A + B |
| direct costs as a percentage of BEC | |
| total installation cost (TIC)[18] | 80% of BEC |
| total direct plant cost (TDPC) | BEC + TIC |
| indirect cost (IC)[18] | 14% of TDPC |
| engineering, procurement, and construction (EPC) | TDPC + IC |
| contingency and owner's cost (C&OC) | |
| contingency[18] | 10% of EPC |
| owner's cost[18] | 5% of EPC |
| total C&OC | 15% of EPC |
| total plant cost (TPC) | EPC + C&OC |

The total plant costs (TPCs) include the cost of each piece of equipment in the plant (only representative equipment, "a" and "b", shown in Table 3 for brevity), installation costs and engineering, procurement, and construction (EPC) costs, along with the owner's costs and contingencies. The bare erected cost (BEC) includes costs of each process equipment of the plant. The total installation costs (TICs) comprise costs associated with on-site facilities and infrastructure (such as piping and instrumentation) and direct and indirect labor required for installation and construction. EPC costs include the total direct plant costs (TDPCs) and indirect costs (ICs). Process contingency represents any uncertainties associated with the process equipment. The TPC is then estimated, adding contingency and owner's costs to the EPC. The cost of individual equipment ($C$) is estimated using eq 3, where the sizing factor ($S/S_O$) is dimensionless and is dependent upon factors such as mass or energy flows that reflect the costs well. Reference equipment costs ($C_O$) were taken from the literature, and an appropriate cost scaling exponent ($n$) was used to scale the reference equipment ($S_O$) to the capacities ($S$) taken from the CLG–FT simulations. All reference equipment costs (Euros) are escalated to correspond to a pre-pandemic base year of 2018, used in this work, using the Chemical Engineering Plant Cost Index (CEPCI)[19] to account for inflation. A pre-pandemic base year is chosen here to avoid underestimation of the total plant costs and other economic uncertainties associated with post-pandemic recovery. Here, the number of required equipment ($m$) is taken as 1, as the plant is sized to 100 $MW_{th}$ of biomass input; thus, all of the front-end and synthesis processes are expected to occur in a single train. The reference equipment costs, capacities, and their corresponding cost scaling parameters used in the TPC estimation in this work are listed in Table 4.

$$C = mC_O\left(\frac{S}{mS_O}\right)^n \frac{\mathrm{CEPCI}_{2018}}{\mathrm{CEPCI}_{\mathrm{year},x}} \tag{3}$$

The costs associated with the CLG reactor system are based on the economic data from the GoBiGas plant,[20,26] which is a first-of-its-kind biomethane production plant that uses a bubbling fluidized bed (BFB) reactor on the gasification side and a circulating fluidized bed (CFB) reactor on the heat generation side of the dual-fluidized bed gasification system. The CLG plant is expected to scale similar to the GoBiGas plant as a result of its configuration on the front-end side of the overall CLG–FT process; thus, the costs associated with the reactor systems, fuel handling, and flue gas system of the GoBiGas plant[20] are considered here. Other downstream equipment in the product gas cleaning and fuel synthesis side are predominantly mature technologies and are expected to vary from the GoBiGas plant, which has a downstream methanation plant. Thus, costs are estimated for each equipment separately, with data reported in the literature. The annual fixed operation and maintenance costs ($O\&M_{fix}$) that include personnel and maintenance costs are calculated as per reported data for the GoBiGas plant.[20] Fixed operating and maintenance costs and costs associated with consumables ($c_{O\&M,consumables}$), such as cooling water, oxygen carrier, and other catalysts used in the process, are listed in Table 5.

## 5. RESULTS AND DISCUSSION

**5.1. Heat and Power Production Potential.** This section presents the heat recovery and power production potential of the two gas cleanup trains, as shown in Figure 2, with different process configurations, described in Table 1, considered in the overall CLG–FT process. Detailed process


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**Table 4. List of Reference Equipment Costs, Capacities, Cost-Scaling Parameters, and Scaling Exponents Used for Cost Estimation of Individual Equipment in Different Process Areas of the CLG−FT Plant**

| equipment unit | reference erected cost $C_O$ (M€)[a] | cost year | reference capacity | cost-scaling parameter | scale factor $n$ | reference |
|---|---|---|---|---|---|---|
| gasification island[b] | | | | | | |
| fuel handling system | 9.33 | 2014 | 32 | thermal feed input $MW_{LHV}$ | 0.8 | 20 |
| gasifier and combustion | | | | | | |
| flue gas system | | | | | | |
| product gas cleaning | | | | | | |
| fabric filter | 0.06 | 2002 | 15.6 | $Nm^3/s$ | | 23 |
| ceramic filter | 5.90 | 2010 | 1.47 | kmol/s | 0.67 | 10 |
| scrubber | 2.78 | 2002 | 12.3 | $Nm^3/s$ | 0.7 | 24 |
| syngas compression | 5 | 2010 | 10 | $MW_{el}$ | 0.67 | 10 |
| tar reforming | 2.99 | 2010 | 12 | $Nm^3/s$ | 0.6 | 25 |
| guard bed | 0.02 | 2002 | 8 | $Nm^3/s$ | 1 | 24 |
| Rectisol[c] | 26.67 | 2007 | 200000 | $Nm^3/h$ | 0.63 | 11 |
| FTS island | | | | | | |
| FT reactor[d] | 12.59 | 2007 | 71360 | $Nm^3/h$ | 0.75 | 11 |
| recycle compressor | 3.80 | 2010 | 10 | $MW_{el}$ | 0.67 | 10 |
| autothermal reformer | 13.55 | 2002 | 100 | $Nm^3/s$ | 0.6 | 24 |
| $CO_2$ compression | 5.84 | 2007 | 10 | $MW_{el}$ | 0.67 | 11 |
| air separation units | 45.5 | 2011 | 31.45 | kg/s | 0.67 | 21 |
| HRSC[e] | 61.76 | 2007 | 275 | $MW_{el}$ | 1 | 11 |
| heat recovery units | 5.2 | 2010 | 43.6 | $MW_{el}$ | 0.8 | 10 |
| boiler[f] | 48.5 | 2008 | 355 | $MW_{th}$ | 1 | 11 |

[a]Costs were converted to Euros where the original costs reported were either in U.S. dollars (1.08$/€) or Swedish krona (10.59 SEK/€). [b]Includes only the fuel handling system (external fuel feeding system and internal fuel feeding system, including lock hoppers, 50.4 $MSEK_{2014}$), gasifier and combustor (reactors and refractory, condensate treatment, and steam generation, 29.49 $MSEK_{2014}$), and flue gas system (flue gas cooler, flue gas filter, flue gas fan, and ash handling system, 18.93 $MSEK_{2014}$).[20] [c]The chosen Rectisol unit captures $H_2S$ and $CO_2$ as separate streams. [d]The cost includes heat exchangers and initial catalyst fill. [e]Includes costs of steam turbine, condenser, piping, and auxiliary equipment.[11] [f]Considered for cases with off-gas combustion for an additional heat supply to the HRSC.

flow diagrams of the two gas cleanup trains are presented in part 1[5] of this study (10.1021/acs.energyfuels.2c00819).

*5.1.1. Case CG.* This section compares the two process configurations within the cold-gas cleanup train, i.e., with off-gas combustion (case CG2) and without off-gas combustion (case CG1), for increasing the product yield in the overall process, as described in Section 2.2. Figure 3 illustrates the FGBG analysis of the CLG−FT process with the cold-gas cleanup process train. It can be observed that the high-temperature excess heat increases in case CG2 as a result of the additional heat supply through off-gas combustion, and thus, the feedwater flow to the HRSC marginally increases by 7.7%, i.e., 13.0 kg/s in case CG1 to 14 kg/s in case CG2. The off-gas combustion raises the steam temperature marginally to 539.5 °C in case CG2 compared to case CG1, where the steam temperature is limited to 450 °C as a result of limitations in raw syngas cooling with steam. The steam extraction pressure levels in the HRSC (foreground process) are chosen iteratively (40/8/1 bar) to fully use the heat pockets of the background process, i.e., the CLG−FT process. In both cases, there is additional intermediate pressure (IP) steam production at the FT reactor that supplies a part of the steam demand in the process at 6 bar pressure levels. A 61% increase in electricity production and a 32.4% increase in DH output are estimated in case CG2 compared to case CG1. This is, however, at the cost of reduced $CO_2$ capture and FT crude output from the process, owing to the off-gas combustion and, thereby, reduced reforming of recycled hydrocarbon stream in case CG2.

*5.1.2. Case HG.* Figure 4 illustrates the FGBG analysis of the CLG−FT process with the hot-gas cleanup process train with its two different process configurations, i.e., cases HG1 and HG2, that incorporate the fate of off-gases produced in this CLG−FT process. Unlike case CG, which inherently requires two AGR units, in case HG, AGR is handled by just one unit, i.e., the Rectisol unit, prior to the FT reactor. This difference can be clearly seen when comparing the GCCs shown in Figures 3 and 4. The low-pressure (LP) steam demand in the CLG−FT process with the hot-gas cleanup train is significantly lower compared to the cold-gas cleanup train, which mainly stems from the steam demand in the gasifier upstream of the process. Thus, resulting in higher amounts of recoverable excess heat that is used in the integrated HRSC. The HRSC in this CLG−FT process train requires only one extraction level at a pressure level of 31 bar to satisfy the steam demand in the autothermal reformer and other steam demands at lower pressure levels. The surplus steam is then expanded in a back-pressure turbine at a 1 bar pressure level for supplying heat to a DH condenser. Off-gas combustion in case HG aids in increasing the steam temperature to 525 °C, and the feedwater supply to the HRSC increases from 15.8 kg/s in case HG1 to 19 kg/s in case HG2. An increase in electricity production by 26.6% and an increase in DH output by 22.4% are estimated for case HG2 compared to case HG1. The various competing factors in the CG and HG cases, with respect to process economics, negative $CO_2$ emissions, and FT crude output, are quantified and listed in Tables 6−8.

**5.2. Capital Costs and LCOF Estimation.** The electricity and steam balance in the four cases investigated are presented in Table S1 of the Supporting Information. The capital cost estimation of all cases is listed in Table 6, where costs associated with individual equipment are presented. The LCOF estimation, as per the cost estimation methodology

Table 5. List of O&M Costs Considered That Include Fixed Costs, Consumables, and Feedstock Costs

| fixed O&M costs | value | unit | reference |
|---|---|---|---|
| labor costs[a] | 181 | SEK/MWh | 20 |
| maintenance costs | 89 | SEK/MWh | 20 |
| O&M consumable costs | | | |
| process water makeup | 2 | €/m$^3$ | 18 |
| oxygen carrier | | | |
| OC costs[b] | 30 | €/tonne | 27 |
| assumed makeup requirement[c] | 6 | kg/MW$_{th}$ | 27 |
| Rectisol unit | | | |
| methanol (MeOH) cost | 1.48 | \$/gallon | 28 |
| specific MeOH consumption | 0.83−1.21 | kmol/kmol$_{syngas_feed}$ | 29 |
| specific MeOH loss | 10 | kg/h | 29 |
| ZnO guard bed | | | |
| cost of catalyst | 355 | \$/ft$^3$ | 30 |
| space velocity | 40 | h$^{-1}$ | 31 |
| catalyst attrition rate assumed | 10 | %/year | |
| void fraction assumed | 30 | % | |
| catalyst lifetime | 2.5 | years | 32 |
| FT reactor | | | |
| space velocity | 100 | h$^{-1}$ | 33 |
| cost of catalyst[d] | 6.31 | €/kg | 33 |
| catalyst lifetime assumed | 3 | years | |
| attrition rates assumed | 10 | % | |
| tar scrubber | | | |
| RME consumption | 31.7 | SEK/MWh | 20 |
| autothermal reformer | | | |
| oxygen cost | 51.7 | €/tonne | 23 |
| biomass feedstock cost[e] | 170 | SEK/MWh | 34 |

[a]On the basis of the 20 MW GoBiGas plant with FLH = 8000 h and scale factor $n$ = 0.1. [b]Sale price of slag-based products of 5−30 €/tonne in Sweden. LD slag is expected to be priced similarly, excluding transport and treatment costs.[27] [c]Corresponds to the replacement rate of silica sand in waste incineration operations. [d]Typically, in the range of 8.5−16 \$/lb.[33] [e]Average cost of refined wood fuel (286 SEK/MWh), wood chips (199 SEK/MWh), byproducts (170 SEK/MWh), and recycled wood fuel (96 SEK/MWh) between 2010 and 2019. Byproduct costs are considered.

listed in Table 3, is presented in Table 7. The total O&M costs, excluding feedstock-related costs, are estimated to be roughly 6.5% of TPC, which is slightly on the higher side compared to similar biomass-to-liquid fuel production plants (4−7%).[20,35,36] In the TPC estimation in Table 6, a process contingency of 10% is applied to all equipment listed, except for the gasification island in HG cases, which assumes a process contingency of 30%, which is typical for less mature technologies. This results in the gasification island incurring a higher share of the total BEC in HG cases (31−35%) compared to CG cases (25−27%). The share of BEC associated with acid gas removal is expectedly higher in CG cases (19−23%) compared to HG cases (10−13%), owing to the two AGR units, Rectisol, and the amine scrubber, in the cold-gas cleanup train compared to the just one Rectisol unit in the hot-gas cleanup train. The estimated bare erected cost of the four cases is estimated to be in the range of 62−71 M€$_{2018}$. Additional boiler costs that include steam generator, stack, and ductwork are applied to cases with off-gas combustion, namely, cases CG2 and HG2. In contrast to this, both cases with off-gas recirculation incur higher TPC, case CG1 (167 M€$_{2018}$) and case HG1 (163 M€$_{2018}$), when compared to their counterpart process configuration with off-gas combustion. Overall, for the four cases investigated, the total plant costs range from 146 to 162 M€$_{2018}$. More information on the fixed and variable O&M costs and the breakdown of total BECs can be found in Table S2 and Figures S1 and S2 of the Supporting Information. Thus, the reduced equipment costs expected in the hot-gas cleanup train cases are offset by the higher costs for filters and higher process contingency assumed for the gasification island.

In Table 7, it can be observed that, while the total O&M costs remain relatively the same, the net heat and electricity production from each process significantly affects the total annual operating costs. While all four cases are electricity-deficient, the deficiency is much lower for HG cases as a result of the higher electricity demand in the multi-stage intercooled compression of clean syngas in the cold-gas cleanup train compared to HG cases, resulting from the fact that the CLG section operates at a moderately higher pressure (∼10 bar). The on-site consumption and production of electricity are listed in Table S1 of the Supporting Information, which includes the additional electricity requirement in the CLG section of HG cases, for air compression, which is marginally lower than what is required in the multi-stage intercooled syngas compressor. Along with this, higher DH output in the HG cases is estimated in comparison to CG cases, and this is reflected in net annual DH/EL costs in Table 7. It is observed that the LCOF of FT crude produced ranges from 115 to 146 €/MWh for the four cases. Clearly, the LCOF estimated for cases with off-gas combustion (CG2 and HG2) results in

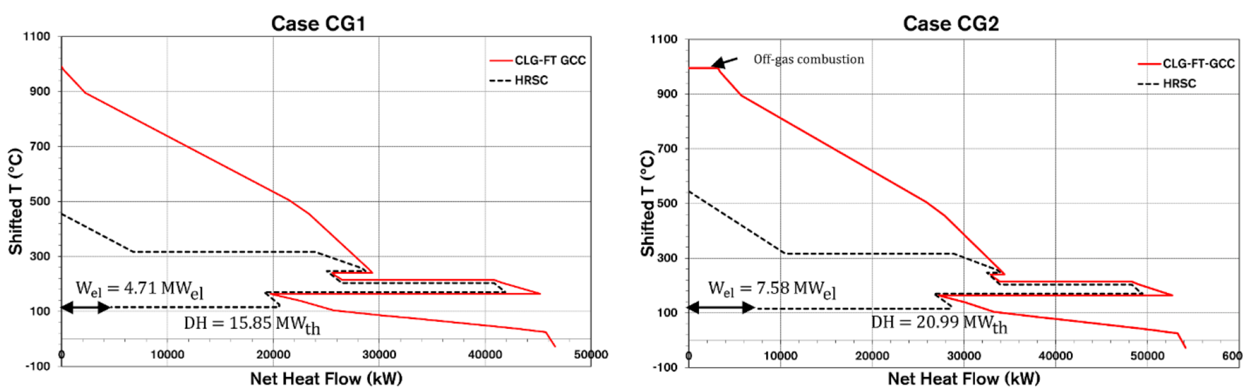


Figure 3. FGBG analysis of CLG−FT plants (cases CG1 and CG2) with the integration of a heat recovery steam cycle.


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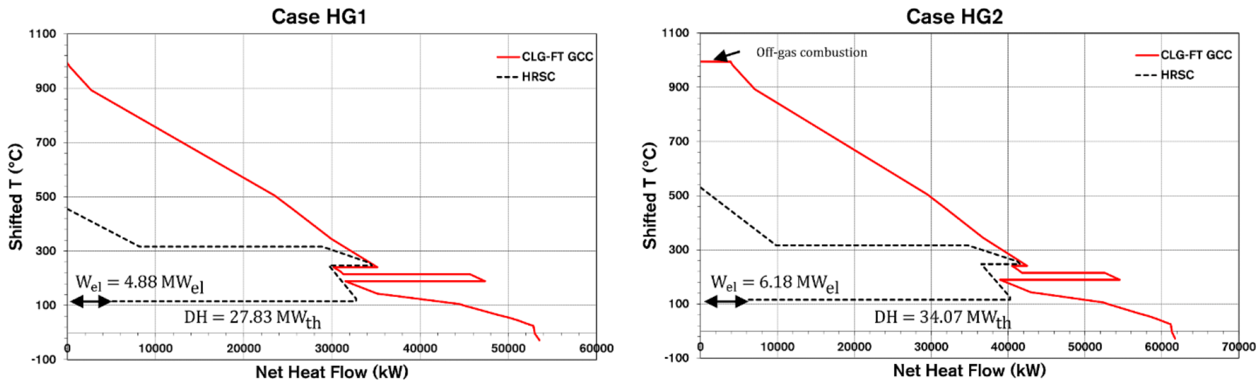


**Figure 4.** FGBG analysis of CLG–FT plants (cases HG1 and HG2) with the integration of a heat recovery steam cycle.

**Table 6. Cost Comparison and Estimation of the Levelized Cost of FT Crude Produced (LCOF)**

| capital cost estimation | case CG1 | | case CG2 | | case HG1 | | case HG2 | |
|---|---|---|---|---|---|---|---|---|
| bare erected cost (BEC) | $M€_{2018}$ | % of BEC | $M€_{2018}$ | % of BEC | $M€_{2018}$ | % of BEC | $M€_{2018}$ | % of BEC |
| gasification island | 18.03 | 25.5 | 18.03 | 26.7 | 21.64 | 31.4 | 21.64 | 34.9 |
| syngas compression | 3.54 | 5.0 | 3.54 | 5.2 | 1.41 | 2.0 | 1.37 | 2.2 |
| tar reformer | 2.93 | 4.1 | 2.93 | 4.3 | 2.68 | 3.9 | 2.68 | 4.3 |
| filter | 0.16 | 0.2 | 0.16 | 0.2 | 3.57 | 5.2 | 3.57 | 5.8 |
| water scrubber | 5.93 | 8.4 | 5.93 | 8.8 | 0.81 | 1.2 | 0.81 | 1.3 |
| guard bed | 0.03 | 0.0 | 0.03 | 0.0 | 0.00 | 0.0 | 0.00 | 0.0 |
| Rectisol | 12.86 | 18.2 | 9.56 | 14.2 | 13.44 | 19.5 | 10.90 | 17.6 |
| FT reactor with HXs | 9.24 | 13.0 | 9.24 | 13.7 | 9.96 | 14.4 | 7.82 | 12.6 |
| autothermal reformer | 3.02 | 4.3 | 3.02 | 4.5 | 3.43 | 5.0 | 2.77 | 4.5 |
| FT recycle compressor | 0.07 | 0.1 | 0.05 | 0.1 | 0.09 | 0.1 | 0.07 | 0.1 |
| $CO_2$ compression | 2.31 | 3.3 | 2.22 | 3.3 | 2.22 | 3.2 | 2.10 | 3.4 |
| oxygen production (ASU) | 5.16 | 7.3 | 3.62 | 5.4 | 6.16 | 8.9 | 4.36 | 7.0 |
| steam cycle | 1.21 | 1.7 | 2.03 | 3.0 | 1.26 | 1.8 | 1.59 | 2.6 |
| heat recovery units | 2.70 | 3.8 | 2.70 | 4.0 | 2.28 | 3.3 | 2.28 | 3.7 |
| WGS reactor | 0.02 | 0.0 | 0.02 | 0.0 | 0.00 | 0.0 | 0.00 | 0.0 |
| amine scrubber | 3.61 | 5.1 | 3.61 | 5.3 | 0.00 | 0.0 | 0.00 | 0.0 |
| boiler | 0.00 | 0.0 | 0.82 | 1.2 | 0.00 | 0.0 | 1.00 | 1.6 |
| total bare erected cost (BEC) | 70.81 | $M€_{2018}$ | 67.50 | $M€_{2018}$ | 68.95 | $M€_{2018}$ | 61.96 | $M€_{2018}$ |
| total plant cost (TPC) | | | | | | | | |
| TIC (80% of BEC) | 56.65 | $M€_{2018}$ | 54.00 | $M€_{2018}$ | 55.16 | $M€_{2018}$ | 49.56 | $M€_{2018}$ |
| TDPC | 127.46 | $M€_{2018}$ | 121.51 | $M€_{2018}$ | 124.10 | $M€_{2018}$ | 111.52 | $M€_{2018}$ |
| IC (14% of TDPC) | 17.85 | $M€_{2018}$ | 17.01 | $M€_{2018}$ | 17.37 | $M€_{2018}$ | 15.61 | $M€_{2018}$ |
| EPC | 145.31 | $M€_{2018}$ | 138.52 | $M€_{2018}$ | 141.48 | $M€_{2018}$ | 127.13 | $M€_{2018}$ |
| C&OC (15% of EPC)[a] | 21.80 | $M€_{2018}$ | 20.78 | $M€_{2018}$ | 21.22 | $M€_{2018}$ | 19.07 | $M€_{2018}$ |
| total plant cost (TPC) | 167.11 | $M€_{2018}$ | 159.29 | $M€_{2018}$ | 162.70 | $M€_{2018}$ | 146.20 | $M€_{2018}$ |

[a]10% EPC – contingency; 5% EPC – owner's cost.

**Table 7. LCOF Estimation for the Base Case Assuming No $CO_2$ Capture Tax Credit**

| LCOF estimation (base case) | units | case CG1 | case CG2 | case HG1 | case HG2 |
|---|---|---|---|---|---|
| annualized total plant cost (TPC)[a] | M€/year | 17.52 | 16.70 | 17.08 | 15.35 |
| annual fixed O&M costs ($c_{O\&M,fix}$) | M€/year | 9.642 | 9.642 | 9.640 | 9.640 |
| annual consumable costs ($c_{O\&M,consumables}$) | M€/year | 0.95 | 0.95 | 0.95 | 0.95 |
| annual feedstock costs[b] ($c_{O\&M,feedstock}$) | M€/year | 12.96 | 12.96 | 12.96 | 12.96 |
| annual electricity costs ($c_{net,EL}$) | M€/year | 4.3 | 2.7 | 2.3 | 1.4 |
| annual DH production ($c_{net,DH}$) | M€/year | −3.8 | −5.04 | −6.68 | −8.18 |
| total costs per year | M€/year | 41.58 | 37.93 | 36.38 | 32.23 |
| FT crude production | MWh/year | 321216 | 258273 | 304031 | 233471 |
| LCOF (eq 1) | €/MWh | 129.44 | 146.87 | 119.31 | 137.60 |
| LCOF (eq 1) | €/BOE[c] | 219.97 | 249.59 | 202.22 | 233.22 |

[a]Annualized TPC = TPC × CCF; CCF = 10.4%/year, calculated on the basis of ref 17. [b]Forest residues, 170 SEK/MWh.[20] [c]One barrel of oil equivalent = 1.69 MWh.

Table 8. LCOF Estimation Considering a $CO_2$ Tax Credit of 1190 SEK/tonne of $CO_2$[37] for the Net-Negative Emissions from the Process

| LCOF estimation, including $CO_2$ capture | units | case CG1 | case CG2 | case HG1 | case HG2 |
|---|---|---|---|---|---|
| net $CO_2$ captured on site | kilotonnes of $CO_2$/year | 178.56 | 158.61 | 167.01 | 143.64 |
| annual avoided costs | M€/year | 20.06 | 17.82 | 18.77 | 16.14 |
| total costs per year | M€/year | 21.51 | 20.11 | 17.61 | 16.09 |
| LCOF, including avoided costs | €/MWh | 66.97 | 77.86 | 57.58 | 68.47 |
| LCOF, including avoided costs | €/BOE | 113.51 | 131.97 | 98.19 | 116.81 |

roughly 13−14% higher LCOF compared to the cases without off-gas combustion (cases CG1 and HG1). It is seen that the costs avoided with higher on-site co-generation are offset significantly by the lower FT crude production in these cases. The FT crude output differs significantly for cases with off-gas combustion, owing to reduced recycling of unconverted hydrocarbons from the FT reactor as well as reduced syngas quality at the FT reactor inlet.

Considering the net $CO_2$ capture from each process and a $CO_2$ tax credit of 1190 SEK/tonne of $CO_2$ or roughly 112.4 €/tonne of $CO_2$,[37] the LCOF ranges from 53 to 78 €/MWh, corresponding to 91−132 €/BOE of FT crude, as shown in Table 8. An economic sensitivity analysis considering important factors and other uncertainties that affect LCOF estimation in the base-case calculations is described in Section 5.3.

Figure 5 illustrates the annual cost estimates and the LCOF estimated for each investigated case with and without a $CO_2$ tax credit. The largest contribution to the total annual costs is predominantly from capital costs (41%), followed by feedstock costs (32%) and O&M costs (26%).

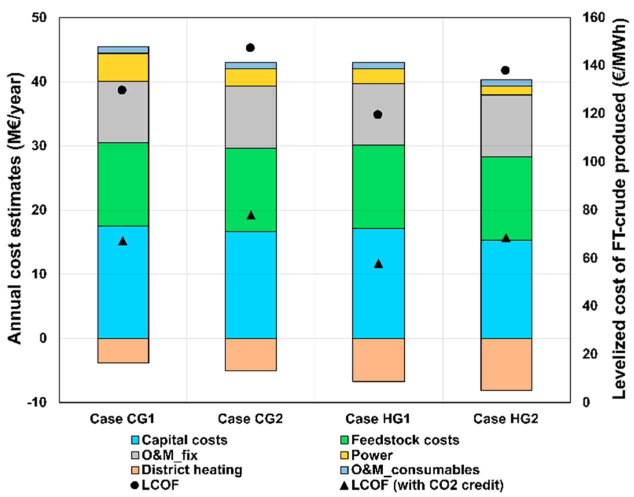


Figure 5. Annual cost estimates (€/year) and levelized production costs (€/MWh) of FT crude for a CLG plant of 100 $MW_{th}$ biomass scale, with and without a $CO_2$ credit (taken here as 1190 SEK/tonne of $CO_2$ or ∼112.4 €/tonne of $CO_2$[4]).

**5.3. Economic Sensitivity Analysis.** The cost estimation methodology applied in this work typically has an accuracy level of ±30%.[21] An economic sensitivity analysis is performed with parameters that are expected to influence the levelized cost of fuel (LCOF), reported in this section as $€_{2018}$/MWh of fuel (FT crude) produced. Here, six economic parameters are considered, namely, capital charge factor (CCF), feedstock costs, $CO_2$ tax credit, district heating and electricity costs, and estimated total plant cost (TPC). Capital cost parameters, such as the TPC and CCF, are given a variation of ±30%, and the operating cost parameters are given a variation of ±50%, as per the EBTF report,[21] to the base case reference values assumed in Tables 6−8. All sensitivity analyses in this section are performed with no $CO_2$ tax credit for the captured $CO_2$ or negative $CO_2$ emission from the CLG−FT process, except for Figure 8b, where a varying $CO_2$ tax credit ranging from 50 to 150 €/tonne of $CO_2$ is considered to illustrate the influence of the $CO_2$ tax credit on the specific production costs for FT crude (€/MWh).

The influence of the cost of district heating and electricity on the specific fuel production costs from the different CLG−FT cases is shown in Figure 6. Here, the cost of electricity and district heating is varied (±50%) from their base case reference values of 50 and 30 €/MWh, between 30 and 70 €/MWh and between 15 and 45 €/MWh, respectively. The LCOF produced is illustrated using a color map in Figure 6 that ranges between 90 and 180 €/MWh, representing the maximum and minimum estimated LCOF in these four cases. The general trend observed in all four cases, unsurprisingly, is that, with increasing electricity costs and decreasing district heating costs, the levelized cost of fuel produced increases significantly. However, it is important to note here that cases HG2 and CG2, which include off-gas combustion, both have higher district heat delivery (∼22 and 32%) and electricity production (∼26 and 61%) compared to their counterpart configurations with a cold-gas cleanup train yet result in higher levelized fuel production costs. First, this can be explained by the diminished FT crude efficiency with off-gas combustion in these cases, thus resulting in significantly lower annual production of FT crude (cf. Table 7). Second, it also highlights the role of optimally using the recovered heat to minimize the levelized fuel production costs. The CLG−FT plants modeled in part 1[5] of this work (10.1021/acs.energyfuels.2c00819) were designed to maximize heat recovery for internal use and deliver heat to the district heating network, operating in a co-generation mode. Thus, the integrated CLG−FT plants are net producers and net consumers of heat and electricity, respectively. From Figure 6, it can be seen that the cost of electricity has a greater influence on the cases with higher electrical deficiency, i.e., cases with the cold-gas cleanup train. On average, the LCOF increases by roughly 7−10% for CG cases and 4−6% for HG cases, with an increase in the cost of electricity from their reference electricity cost of 50 €/MWh. The optimal use of recovered heat and process configuration is significantly dependent upon the region where such integrated biomass-to-liquid plants operate. In a recent study by Maier et al.,[38] this phenomenon is explored in detail, and it was concluded that the regional heat and electricity market determines the optimal process design of such plants, while regions with low feedstock costs are favorable for lower specific fuel production costs. Overall, Figure 7 shows that case HG1, which includes a hot-


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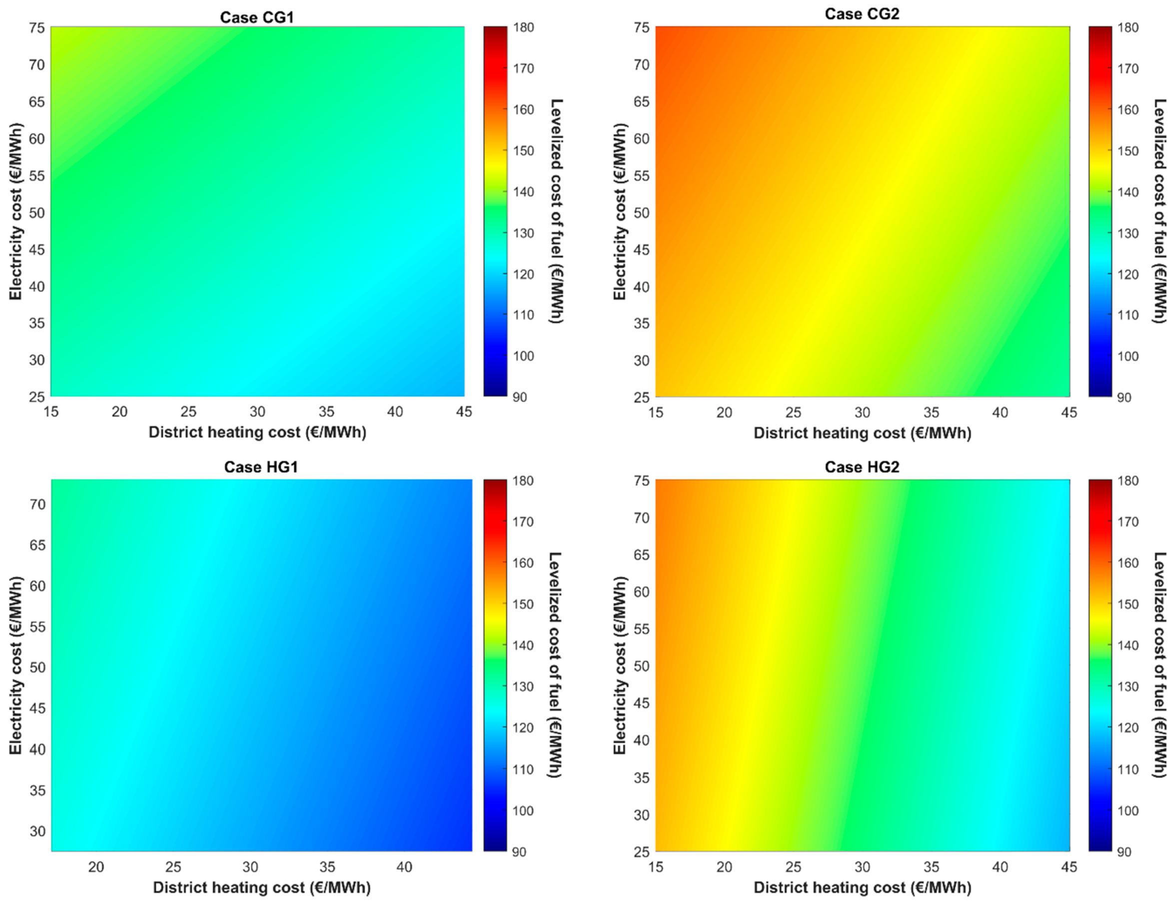


Figure 6. Influence of varied electricity (€/MWh) and district heating (€/MWh) costs on the levelized cost of fuel (€/MWh) for the four cases investigated. The color bar shows the LCOF (€/MWh) ranging between 90 and 180 €/MWh, assuming no $CO_2$ tax credit for the negative $CO_2$ emissions.

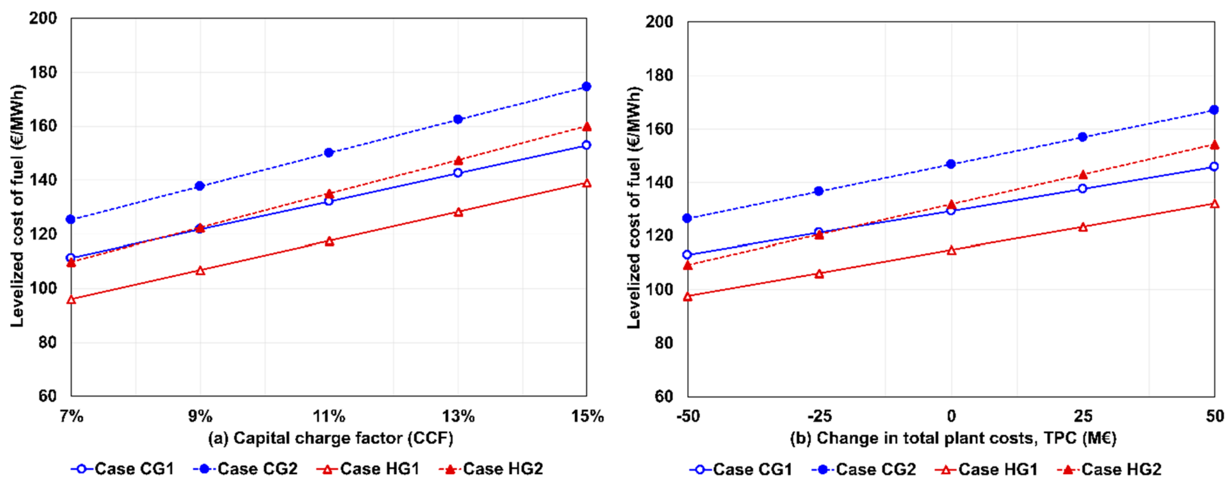


Figure 7. Sensitivity of levelized cost of fuel production (€/MWh) toward capital cost parameters, such as (a) capital charge factor and (b) total plant costs, with a variation of ±30% to the base case LCOF estimation, with no $CO_2$ tax credit, listed in Table 7.

gas cleanup train that typically has higher amounts of recoverable heat and maximizes FT crude production with off-gas recycling, incurs the lowest LCOF, which ranges from 97 to 145 €/MWh, and this is followed by cases CG1 (109–153 €/MWh), HG2 (109–173 €/MWh), and CG2 (124–174 €/MWh).

Figure 7 shows the influence of capital cost parameters, such as the capital charge factor and the total plant costs on the levelized cost of fuel produced, with a variation of ±30% to the estimated base case reference values for the CCF (~10.4%) and the varied TPC estimated for the four cases, listed in Table 6. The influence of CCF is rather marginal and maintains the overall trend of higher LCOF for case CG2, followed by HG2, CG1, and HG1, until a CCF of 9%, as shown in Figure 7a. A CCF lower than 9% yields lower LCOF for case CG1 than case HG2. Similar trends are observed with a change in total plant


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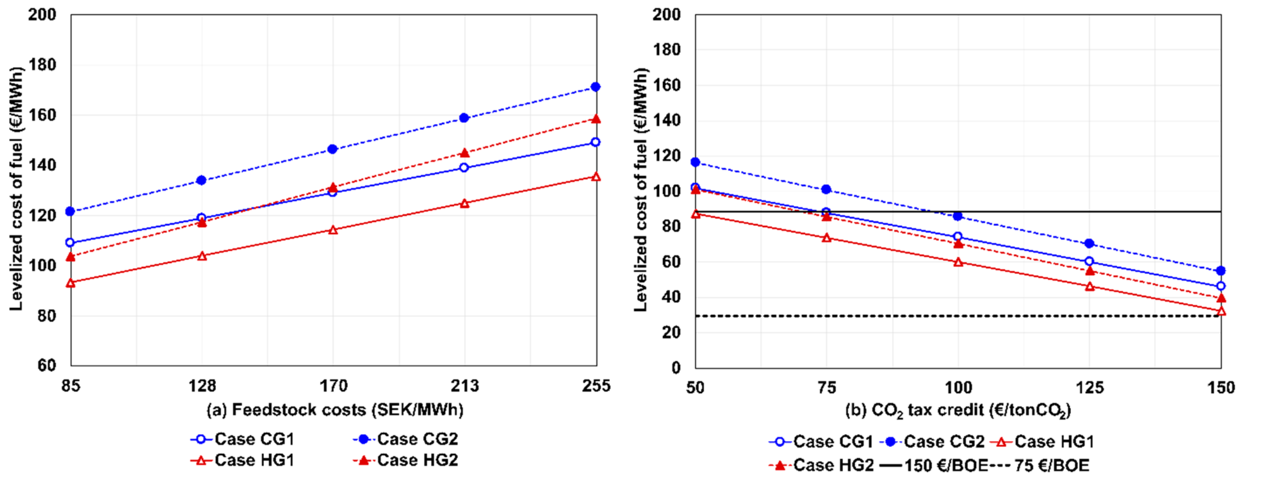


**Figure 8.** Sensitivity of levelized cost of fuel production (€/MWh) toward operating cost parameters, such as (a) feedstock costs with a variation of ±50% to the base case LCOF estimation, listed in Table 7, with no $CO_2$ tax credit and (b) $CO_2$ tax credit with a variation of ±50% to the base case LCOF estimation, listed in Table 8, with a $CO_2$ tax credit of 1190 SEK/tonne of $CO_2$ captured. Solid and dashed black lines indicate the LCOF levels of 150 and 75 €/BOE, respectively. Note the different scales of the *y* axis in panels a and b.

costs, as shown in Figure 7b, with the lowest overall fuel production cost (€/MWh) for case HG1.

The techno-economic evaluation, described in Section 4, is performed with conditions relevant to Sweden, which also has other favorable regional aspects, such as a relatively high $CO_2$ tax,[37] higher availability of forest residues,[39] and a peculiar district heating and electricity market.[40] Thus, the two former operating cost drivers, i.e., the feedstock costs and the $CO_2$ tax credit, influencing the LCOF are chosen for sensitivity analysis, with a variation of ±50% to their base case reference values of 170 SEK/MWh and 1190 SEK/tonne of $CO_2$ (∼112.4 €/tonne of $CO_2$), shown in panels a and b of Figure 8, respectively. In Figure 8b, horizontal lines are used to show the specific fuel production cost levels as Euros per barrel of oil equivalent (€/BOE).

Feedstock costs contribute nearly a third of the total annual costs of the plant, and expectedly, the cost of the feedstock has a significant effect on the LCOF estimates. A percentage change of roughly ±20% in all four cases is seen for the specific fuel production costs that correspond with a range of ±50% from the reference feedstock cost of 170 SEK/MWh. It is important to note that no feedstock pretreatment is considered in this analysis; thus, a moderately high feedstock cost of 170 €/MWh was considered in the LCOF estimation. Typically, on-site feedstock pretreatment would reduce the low-temperature heat sold to the DH network. An interplay of untreated feedstock costs and DH costs is expected to reduce the specific costs of FT crude produced. In comparison to the feedstock costs, the $CO_2$ tax credit tends to have an even greater influence on the LCOF estimated in each case, as shown in Figure 8. The four cases yield a LCOF range of 87−116 and 32−55 €/MWh, for a $CO_2$ tax credit ranging from 50 and 150 €/tonne of $CO_2$ captured, respectively. It is seen that the LCOFs for all cases remain above the 75 €/BOE level, even with a $CO_2$ tax credit as high as 150 €/tonne of $CO_2$. The highest LCOF is estimated for case CG2, followed by CG1, HG1, and HG2.

The LCOF range estimated in this work, without a $CO_2$ tax credit, that ranges from 120 to 147 €/MWh is comparable to the reported LCOF values reported by the Subgroup on Advanced Biofuels (SGAB) report[41] and the IEA Bioenergy Task 41 report[42] that estimate a total production cost of 86−129 and 75−144 €/MWh, respectively, for biomass-derived FT crude.[42] Similarly, Hannula[35] reported a LCOF of 82.8 $€_{2018}$/MWh (∼23 $€_{2010}$/GJ) for a pressurized oxy-blown fluidized bed gasifier used for gasifying forest residues for the production of gasoline via FTS. Although the fuel production costs estimated in this work are comparable to these studies, it is rather difficult to make a direct comparison as a result of the differences in gasification technology used for syngas production, other downstream process equipment, and various underlying economic assumptions that could differ from this study. Nonetheless, considering only the biomass-to-X routes presented in the review by Poluzzi et al.,[43] it can be seen that the lowest biofuel production costs are incurred by the production of synthetic natural gas (8.5−20.4 $€_{2019}$/GJ), followed by methanol (11−20.6 $€_{2019}$/GJ), gasoline (22.6−31.3 $€_{2019}$/GJ), and FT crude (61.7−73.1 $€_{2019}$/GJ) from biomass. In comparison to this, the LCOF estimated in this work, when converted to the same unit and cost year, is roughly 34−42 $€_{2019}$/GJ. This difference in the LCOF for the reported biomass-to-FT pathway[43,44] is most likely resulting from the different cost estimation methodology, higher electricity costs, and additional process units, such as biomass pretreatment and pure $CO_2$ gas conditioning, considered in their work. However, in comparison of the total plant costs (M $€_{2018}$) normalized to the capacity of the gasification plant ($MW_{LHV}$) or specific investment costs (reported as either M €/$MW_{th}$ or €/$kW_{th}$), the 100 $MW_{th}$ CLG−FT plant studied in this work yields roughly 1.46−1.67 M$€_{2018}$/$MW_{LHV}$, which is lower than the values reported by Holmgren et al.[12] In this study, several biomass-to-FTS plants, with mainly oxy-blown pressurized gasification technologies, were scaled to a plant capacity of 480 $MW_{th}$ and compared, where the specific investment costs ranged from 1.75 to 2.89 M$€_{2018}$/ $MW_{LHV}$. It is imperative to note here that the latter could likely have significant advantages of economies of scale on the levelized fuel production costs, yet the 100 $MW_{th}$ CLG−FT plant yields a relatively lower specific investment cost.


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## 6. CONCLUSION

CLG for FT crude production with net-negative $CO_2$ emissions has been analyzed and compared to different process configurations with respect to the gas cleanup train and the off-gas valorization routes. In addition to this, heat and power production potential are also estimated. The total plant costs estimated for the CLG−FT plant, sized to 100 $MW_{th}$ input of biomass, are estimated to be around 159−167 and 146−162 M€ for the cases with cold-gas cleanup and hot-gas cleanup process trains, respectively. Even with high process contingency, assumed in the gasification side of the HG cases, the TPC is in a relatively narrow range of 146−167 M€, owing to avoided equipment costs in the HG process train. In general, a CLG−FT plant with a cold-gas cleanup train is estimated to have, on average, 5% higher total plant costs compared to the hot-gas cleanup plant configurations. Additionally, the cases with off-gas combustion have lower TPC, by roughly 5−10%, when compared to their counterpart configurations with off-gas recirculation to increase the FT crude production.

The fixed operating costs are relatively the same for all plant configurations. The total annual operating costs vary considerably for each plant configuration, mainly as a result of varying chemical efficiency or biomass-to-FT crude production efficiency, net-electricity consumption, and excess high-temperature heat available for recovery. The sensitivity analyses performed in this work indicate that the biomass-to-FT crude production efficiency of the process, unsurprisingly, has the biggest impact on the estimated specific production cost for FT crude produced. Thus, cases with off-gas recirculation incurred lower specific fuel production costs. A CLG−FT plant with a cold-gas cleanup train has higher net-electricity consumption and, expectedly, lower amounts of excess heat available at desired temperature levels for heat recovery when compared to a CLG−FT plant with a hot-gas cleanup train, the former, thus resulting in an overall system energy efficiency of 45% compared to 63% in HG cases. This is reflected in the final LCOF estimation, where the LCOFs of HG cases are approximately 10% lower than in CG cases. Sensitivity analysis is performed on the economic assessment with parameters such as capital charge factor, varying feedstock costs, $CO_2$ tax credit, and varying electricity and district heating costs to evaluate the factors that influence the most on the process economics and compare the two-gas cleaning configurations. The factors that had the most influence on the specific production costs for FT crude were the $CO_2$ tax credit for the negative $CO_2$ emissions, electricity and district heating costs, and feedstock costs, in that order. The sensitivity analysis with varying electricity highlights the need for optimally using the recovered excess heat available in the CLG−FT plants, which are typically net consumers of electricity. The lowest levelized fuel production costs of FT crude were estimated for case HG1 (120 €/MWh) with off-gas recirculation and higher amounts of recoverable heat, followed by cases CG1 (129 €/MWh), HG2 (138 €/MWh), and CG2 (147 €/MWh), assuming no $CO_2$ tax credit for net-negative emissions from these CLG−FT processes. Although, with off-gas recirculation, the biomass-to-FT crude production efficiency is similar for cases CG1 (40.3%) and HG1 (38%), the latter has considerably lower levelized fuel production costs, owing to the higher amount of recoverable heat that could be used for co-generation of heat and power. A $CO_2$ tax credit similar to that of Sweden's carbon tax of 1190 SEK/tonne of $CO_2$ or roughly 112.4 €/tonne of $CO_2$ captured in the CLG−FT process is estimated to reduce the estimated LCOF by roughly 50.3%, bringing down the LCOF to 58−78 €/MWh, corresponding to 98−133 €/BOE, bringing it within a competitive range of high crude oil prices. The estimated LCOFs of 120−147 €/MWh for the CLG−FT process are comparable to other biomass-to-FTS processes (∼86−144 €/MWh) reported in the literature. However, it is expected to have a marginally lower specific investment of roughly 1.46−1.67 $M€_{2018}/MW_{th,\ LHV}$ compared to the latter, at a similar FT crude production capacity.

CLG−FT plant employing a hot-gas cleanup train without off-gas combustion clearly shows the lowest LCOF estimates with high overall efficiency compared to other process configurations. Additionally, the hot-gas cleanup configurations are expected to have more operational flexibility, owing to higher amounts of recoverable heat. Further, off-gas valorization routes explored in this work provide more flexibility to the CLG−FT plant, which could operate to either maximize fuel production via off-gas recirculation or alternatively switch to off-gas combustion for maximizing on-site steam generation for heat and power production. The optimal operating mode of the heat recovery steam cycle would, however, depend upon external factors, such as regional market conditions. Thus, the ideal process configuration for a CLG−FT plant that co-produces heat, electricity, and FT crude and provides negative $CO_2$ emissions as a service would eventually depend upon several factors, such as the future development of processes, with respect to the current operational uncertainties, in the hot-gas cleanup train, regional market conditions, and incentives for negative $CO_2$ emissions. Nevertheless, a CLG−FT plant with a cold-gas cleanup train with relatively lower operation uncertainties would enable FT crude production with negative $CO_2$ emissions, with marginally higher (∼10%) levelized fuel production costs. A $CO_2$ tax credit for the net-negative emissions from the process is, however, critical to making CLG−FT a viable solution to replace fossil-based transportation fuels in critical sectors, such as aviation and maritime sectors.

## ■ ASSOCIATED CONTENT

### Supporting Information

The Supporting Information is available free of charge at https://pubs.acs.org/doi/10.1021/acs.energyfuels.2c01184.

Steam and electricity balances in CLG−FT for CG and HG cases (Table S1), annual O&M fixed cost estimates and annual consumable cost estimates (Table S2), and share of individual equipment in the total bare erected costs for cases CG1 and HG1 (Figures S1 and S2) (PDF)

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Tharun Roshan Kumar, conceptualization, methodology, writing, visualization, and original draft; Tobias Mattisson, conceptualization, supervision, and writing—review and editing; and Magnus Rydén, conceptualization, supervision, and writing—review and editing.

### Notes

The authors declare no competing financial interest.

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## ■ NOMENCLATURE

AGR = acid gas removal
AR = air reactor
ASU = air separation unit
BEC = bare erected cost (M€$_{2018}$)
BOE = barrel of oil equivalent
BUA = bottom-up approach
C&OC = contingency and owner's cost
CCF = capital charge factor (%/year)
CCS = carbon capture and storage
CEPCI = Chemical Engineering Plant Cost Index
CG1 = cold-gas cleanup train with off-gas recycling
CG2 = cold-gas cleanup train with off-gas combustion
CLG = chemical looping gasification
CLG-FT = overall process within the system boundary in Figure [2]
DH = district heating
EBTF = European Benchmarking Task Force
FGBG = foreground−background
FLH = full load hour (h/year)
FR = fuel reactor
FT = Fischer−Tropsch
FT crude = Fischer−Tropsch synthesis crude or syncrude
GCC = grand composite curve
HC = hydrocarbon
HG1 = hot-gas cleanup train with off-gas recycling
HG2 = hot-gas cleanup train with off-gas combustion
HP = high pressure
HRSC = heat recovery steam cycle
IC = indirect cost
LCOF = levelized cost of fuel
LD slag = Linz−Donawitz slag or steel-converted slag
LHV = lower heating value
LP = low pressure
LT = lifetime (year)
MC = moisture content (wt %)
MeOH = methanol
NETL = National Energy Technology Laboratory
O&M = operations and maintenance
OC = oxygen carrier
RME = rapeseed methyl ester
SNG = synthetic natural gas
SGAB = Subgroup on Advanced Biofuels
TDPC = total direct plant cost (M€$_{2018}$)
TIC = total installation cost (M€$_{2018}$)
TPC = total plant costs (M€$_{2018}$)

### Symbols

$c$ = specific annual operating costs (M€$_{2018}$/year)
$C$ = equipment cost with the capacity $S$
$C_o$ = reference cost of equipment with reference capacity $S_o$
$m$ = number of installed units
$n$ = cost scaling factor
$Q$ = product output (MW$_{th}$)
$S$ = capacity (scaling parameter in cost functions)
$S_o$ = reference equipment capacity
$W$ = work output
$x$ = year when reference equipment cost was estimated
$\eta$ = efficiency
$X$ = liquid biofuel

### Subscripts/Superscripts/Exponents

DH = district heating
EL = electricity
fix = fixed
o = reference
O&M = operations and maintenance
sys = system
th = thermal
var = variable

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