Techno-Economic and Environmental Assessment of Biomass Gasification and Fischer–Tropsch Synthesis Integrated to Sugarcane Biorefineries

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1. Techno-economic parameters for simulation of integrated gasification and Fisher–Tropsch synthesis

The flow diagram of thermochemical route for Fisher-Tropsch (FT) biofuels production considered in this study is presented in Figure S1. The process design comprises the preprocessing, gasification, syngas cleaning and conditioning, fuel synthesis, hydrotreatment, steam and power generation, and utilities areas. It is mostly based on Dutta et al. [1], Nexant [2] and Swanson et al. [3] processes configurations. Figure S2 shows the flow diagram of steam and power generation and utilities areas, including water system, which is mostly based on Dutta et al. [1]. The following subsections contain a detailed technical process overview, including the assumptions for material and energy balances, and for installed equipment cost estimations in each process area. Energy estimates for cooling and heating are based on polynomial heat capacity at constant pressure - parameters from Perry [4] and Sandler [5]. Parameters for economic and risk assessments are also presented.

1.1. Preprocessing Area

The preprocessing area includes the unit operations required for preparing the lignocellulosic material (LCM) for feeding into the gasifier. The initial moisture content and composition are shown in Table S1. The higher heating values of biomasses on a dry basis (HHV_d), presented in Table S1, were calculated using carbon (C), hydrogen (H), oxygen (O), nitrogen (N), sulphur (S), and ash (A) contents (mass percentages on dry basis) and the Channiwala and Parikh’s unified correlation (equation 1) [6].

\[ HHV_d (\text{MJ/kg}) = 0.3491 \times C + 1.1783 \times H + 0.1005 \times S - 0.1034 \times O - 0.0151 \times N - 0.0211 \times A \]  

(1)

The low heating value in a dry and dry-and-ash-free basis (respectively LHV_d and LHV_daf), KJ/kg, may be related to the higher heating value (HHV) (in kJ/kg), its hydrogen (H) and moisture...
mass fractions in a dry-and-ash-free basis, and ash content \( A \) in dry basis, as presented in equations 2, 3 and 4 [7]:

\[
LHV_d \left( \frac{kJ}{kg} \right) = LHV_{daf} * (1 - A_d) \tag{2}
\]

\[
LHV_{daf} \left( \frac{kJ}{kg} \right) = HHV_{daf} - H_{daf} * 20300 - M_{daf} * 2260 \tag{3}
\]

where,

\[
HHV_{daf} \left( \frac{kJ}{kg} \right) = \frac{HHV_d}{(1 - A_d)} \tag{4}
\]

Table S1. Lingocellulosic biomass characterization.

| Composition (weight %, dry basis) | Conventional Sugarcane Bagasse ¹ | Sugarcane Straw ² | Energy Cane Bagasse ³ | Eucalyptus |
|----------------------------------|----------------------------------|-----------------|----------------------|------------|
| Carbon                           | 46.0⁴                          | 45.6⁴          | 46.0⁶               | 46.2⁷      |
| Oxygen                           | 44.5⁴                          | 43.9⁴          | 44.5⁶               | 47.2²      |
| Hydrogen                         | 5.8⁴                           | 5.7⁴           | 5.8⁶                | 5.8⁴       |
| Nitrogen                         | 0.6⁵                           | 0.5⁵           | 0.6⁶                | 0.2⁷       |
| Sulphur                          | 0.1⁵                           | 0.1⁵           | 0.1⁵                | 0.02⁷      |
| Chlorine                         | 0.02⁴                          | 0.1⁵           | 0.02⁵               | 0.0012⁷    |
| Ash                              | 2.9⁴                           | 4.1⁴           | 2.9⁶                | 0.5⁷       |
| Moisture                         | 50⁴                            | 15⁴            | 50⁴                 | 45⁸        |
| HHV, MJ/kg, dry basis            | 18.3                           | 18.0           | 18.2                | 18.1       |
| LHV, MJ/kg, dry basis            | 16.8                           | 16.7           | 16.8                | 16.7       |
| LHV, MJ/kg, dry and ash free basis | 17.4                         | 17.3           | 17.3                | 16.7       |

¹ Including mineral and vegetal impurities. ² Composition for baled straw. ³ Considering integral composition (stalk and straw). ⁴ Carbon, oxygen, hydrogen, ash and moisture content of sugarcane LCM were calculated from the VSB database [8]. ⁵ Data from Tufaile Neto [9]. ⁶ Assumed equal to conventional sugarcane. ⁷ Data from Telmo [10]. ⁸ Data from Baliban [11] and Brito [12].

Baled straw (from CS), is unbaled, chopped and sieved whereas eucalyptus LCM passes through a chipper. After size reduction, the feedstock is dried to 10% wt. moisture in a direct-contact steam rotatory dryer [3,13]. Rotatory dryers are less sensitive to the material’s particle size and are the most common type for biomass [14]. Advantages of using steam for drying instead of hot flue gas from downstream process include the possibility of energy recovery [15] and the better heat transfer, resulting in shorter residence time [3]. Hot flue gases from the char combustor are used to heat steam (2.5 bar) to 210°C. The wet feedstock is dried with superheated steam, which end up still with superheating [3]. From the steam leaving the dryer, 90% is recirculated and 10%, corresponding to the amount of water evaporated from the biomass, is removed and used in other parts of the plant as 2.5-bar steam [14]. A 2% purge (of removed steam) is considered. The main parameters adopted in the process simulation are summarized in Table S2. Capital costs for equipment in this area are estimated considering three sub areas: bale (straw) reception, eucalyptus chipping and biomass drying, according to Table S3.
Figure S1. Integrated gasification and Fisher-Tropsch synthesis process flow diagram.
Figure S2. Steam and power generation and utilities flow diagram.
Table S2. Main parameters adopted in process simulation.

| Area          | Parameter                                                      | Value                              | Reference |
|---------------|----------------------------------------------------------------|------------------------------------|-----------|
| Preprocessing | Electricity requirements in chipper for eucalyptus size reduction | 2 kWh/m³                            | [16]      |
|               | Electricity demand in sieve and choppers                       | 3.73 kWh/t dry straw               | [17]      |
|               | Efficiency of dirty removal from bales                         | 65%                                | [18]      |
|               | Target biomass moisture content                                 | 10%                                | [13]      |
|               | Recirculated steam (% of steam living the dryer)               | 90%                                | [14]      |
|               | Water evaporated from the biomass (% of steam living the dryer) | 10%                                | [14]      |
|               | Heat requirements for drying                                   | 1.5 MJ/kg evaporated water         | Calculated|
| Gasification  | Steam/dry biomass ratio                                        | 0.4 t/t                            | [1]       |
|               | Excess air                                                     | 20 %                               | [1]       |
|               | Gasifier Temperature                                           | 869 °C                             | [1]       |
|               | Gasifier Pressure                                              | 2.3 bar                            | [1]       |
|               | Combustor temperature                                          | 987 °C                             | [1]       |
|               | Combustor pressure                                             | 2 bar                              | [1]       |
|               | S to char                                                      | 8.3 %                              | [1]       |
|               | N to char                                                      | 6.6 %                              | [1]       |
|               | Ash to char                                                    | 100 %                              | Assumed   |
|               | MgO / K ratio                                                 | 2 mol/mol                          | [1]       |
|               | Olivine/dry biomass ratio                                      | 27 kg/kg                           | [1]       |
|               | Primary gasifier cyclone efficiency                            | 99.9 %                             | [1]       |
|               | Secondary gasifier cyclone efficiency                          | 90 %                               | [1]       |
|               | Primary combustor cyclone efficiency                           | 99.9 %                             | [1]       |
| Reforming     | Reformer temperature                                           | 925 °C                             | [1]       |
|               | Reformer pressure                                              | 1.77                               | [1]       |
|               | Temperature of hot flue gas                                    | 969 °C                             | [1]       |
|               | Reformer catalyst makeup                                       | 0.03 kg/t syngas in                | [1]       |
|               | Supplemental combustions gases LHV lost                        | 2%                                 | Assumed   |
|               | Methane (CH₄) conversion                                       | 80%                                | [1]       |
|               | Ethane (C₂H₆) conversion                                       | 99%                                | [1]       |
|               | Ethylene (C₂H₄) conversion                                     | 90%                                | [1]       |
|               | Tars conversion                                                | 99%                                | [1]       |
|               | Ammonia conversion                                             | 90%                                | [1]       |
|               | Acetylene (C₂H₂) conversion                                    | 90%                                | [1]       |
|               | Excess air feed into the regenerator                           | 20%                                | Assumed   |
| Syngas        | Syngas temperature after cooling                               | 60 °C                              | [1]       |
| cooling,      | NH₃ removal on scrubbing system                                 | 99.9%                              | [19]      |
| quenching,    | Quench water circulation rate                                  | 1 L/m³ gas flow                    | [20]      |
| and scrubbing | Fresh quench water makeup                                      | 5 kg/t water circulating            | [1]       |
|               | Electricity requirement for quench water recirculation pump     | 0.1 kWh/t syngas                   | [2]       |
|               | Syngas pressure out                                            | 30 bar                             | [2]       |
| Section                  | Description                                                                 | Value                          | Source     |
|--------------------------|------------------------------------------------------------------------------|--------------------------------|------------|
| Compressors              | Syngas temperature out                                                      | 43 °C                          | [2]        |
|                          | Cooling water to intercoolers                                               | 32 t water/t inlet syngas      | [2]        |
|                          | Water mass fraction in syngas out                                           | 0.004                          | Assumed    |
|                          | Electricity requirements                                                    | 0.2 MWh/t inlet syngas         | [2]        |
| Acid gas removal         | Absorption capacity                                                         | 0.6 mol acid gas/mol amine      | [21]       |
|                          | DEA concentration                                                           | 30 %                           | [21]       |
|                          | H₂S removal                                                                 | 99.9 %                         | [2, 22]    |
|                          | CO₂ removal                                                                 | 92 %                           | [2]        |
|                          | DEA losses                                                                  | 0.01 %                         | [3]        |
|                          | 4 bar steam consumption on stripper reboiler                                | 3620 MJ/t acid gas sent hourly to sulfur recovery | [2]        |
|                          | Electricity for pumps                                                        | 1.06 kWh/mass circulating amine solvent | [2]        |
|                          | Cooling water                                                               | 32.2 t/t inlet syngas          | [2]        |
|                          | Sweet gas temperature                                                       | 46 °C                          | [2]        |
|                          | Temperature of reboiler condensate                                           | 130 °C                         | Assumed    |
|                          | Pressure of reboiler condensate                                             | 3 bar                          | Assumed    |
|                          | LO-CAT® chemicals                                                           | 0.08 t/t sulfur                 | [2]        |
|                          | LP steam (2.5 bar)                                                          | 0.28 kg/kmol acid gas          | [2]        |
|                          | Electricity consumption                                                     | 0.5 kWh/kmol acid gas          | [2]        |
|                          | Makeup water                                                                | 4.7% of water in acid gas      | [2]        |
|                          | ZnO bed absorbent                                                           | 48.5 m³ per kg H₂S in          | [2]        |
|                          | ZnO bed absorbent replacement period                                         | 1 year                         | [2]        |
|                          | ZnO bed operating temperature                                               | 349 °C                         | [2]        |
|                          | PSA efficiency of hydrogen removal                                          | 85 %                           | [3]        |
|                          | FT reactor inlet temperature                                                | 200 °C                         | [19]       |
|                          | FT reactor outlet temperature                                               | 240 °C                         | [19]       |
|                          | FT reactor operating pressure                                               | 25 bar                         | [3]        |
|                          | Recycle compressor’s electricity requirements                               | 0.05 MWh/t recycle             | [3]        |
|                          | FT reactor gas hourly space velocity                                        | 100 hr⁻¹                       | [3]        |
|                          | FT catalyst replacement period                                              | 3 years                        | [3]        |
| Product distribution ¹   | Hydrogen                                                                    | 0.8 wt %                       | [23]²      |
|                          | Fuel gas                                                                    | 8.1 wt %                       | [23]²      |
|                          | Liquid petroleum gas                                                        | 3.3 wt %                       | [23]²      |
|                          | Motor gasoline                                                              | 33.5 wt %                      | [23]²      |
|                          | Jet fuel                                                                    | 29.0 wt %                      | [23]²      |
|                          | Diesel fuel                                                                 | 25.3 wt %                      | [23]²      |
|                          | Electricity                                                                 | 0.125 MWh/t FT liquids processed | [3]        |
|                          | Fuel gas LHV lost                                                           | 2 %                            | Assumed    |
|                          | Air excess                                                                  | 20 %                           | Assumed    |
|                          | Energy (heat) requirements                                                  | 0.29 MWh/t FT liquids processed | Calculated³ |
|                          | Energy released (to cooling water)                                          | 0.43 MWh/t FT liquids processed | Calculated³ |
| Compressor electricity requirements | 112 kWh/t air in                      | Calculated³                   |            |
Steam and Power Generation

- Gas turbine cycle overall efficiency: 40% [7,24,25]
- Gas turbine output flue gases temperature: 600 °C [25]
- Steam turbines isentropic efficiency: 85% [26]
- Generators efficiency: 98% [26]
- Steam temperature after Turbine 1: 344 °C Calculated 3
- 2.5 bar steam extraction temperature: 132 °C Calculated 3
- 4 bar steam extraction temperature: 172 °C Calculated 3
- Steam pressure in condensing turbine (turbine 2) output: 0.11 bar [27]
- Steam temperature after condensing turbine (turbine 2): 47.68 °C Calculated
- Blowdown to cooling tower temperature: 45 °C [1]
- Condensate temperature: 90 °C Assumed
- Condensate pressure: 2 bar Assumed
- Cooling water supply (CWS) temperature (Ts): 27 °C [1]
- Cooling water return (CWR) temperature (Tr): 38 °C [1]
- Losses from evaporation ($F_{evap}$): $0.00085 \times (Tr-Ts) \times \text{CWS rate (vol)}$ [1]
- Losses from drift: 0.2% of CWS rate [1]
- Cycles of concentration (C): 5 [1]
- Required blowdown: $F_{evap}/(C-1)$ [1]
- Flue gas washing water: 0.03 t water/ t flue gas [1]
- SO2 removal in flue gas scrubber: 70% [1]
- Electricity requirements: 0.3 kW/t CWS rate [1]

1 Excludes H2, CO, CO2, and H2O from LTFT synthesis.
2 Disregarding unrecovered organics and water from Klerk’s data [23].
3 Based on AspenPlus® simulation.

Table S3. Estimates for capital costs in each thermochemical process area.

| Area          | Section               | Base capacity          | Original equipment cost | Base Year | Scaling exponent | Cost Indexes Ratio | Equipment installation factor | References |
|---------------|-----------------------|------------------------|-------------------------|-----------|------------------|----------------------|-----------------------------|------------|
| Preprocessing | Bale reception        | 90,000 t/season        | R$ 14,000,000$0.2       | 2017      | 0.6              | -                    | -                          | Based on industry information |
|               | Eucalyptus chipping   | 60 m³/h                | R$ 256,500$0.23         | Oct 2018  | 0.94             | 2                    | 2.47                        | [16]        |
|               | Drying                | 74 t evaporated water/ h | US$ 7,627,455           | 2002      | 1                | 1.45                 | 2.47                        | [28]        |
| Gasification and Syngas Cleaning |          | 20.8 dry biomass t/h | US$ 9,700,000$0.6       | 2010      | 0.6              | 1.04                 | 2.31                        | [1]         |
| Component                        | Description                                                                 | Cost (US$)     | Year | 
|----------------------------------|-----------------------------------------------------------------------------|----------------|------| 
| Syngas Cleaning and Conditioning | Compression 28.93 MWe consumed                                               | 16,805,000     | 2005 | 0.62 | 1.22 | 2.47 | [2] |
|                                  | 66.4 t acid gas to Amine System sulfur recovery/h                            | 6,789,129      | 2007 | 0.75 | 1.09 | 3    | [3] |
|                                  | LO-CAT® 42.2 kg sulfur/h                                                     | 2,917,500      | 2009 | 0.75 | 1.1  | 1.35 | [1] |
| Fuel Synthesis                   | FT fixed bed reactor 100 MW - FT Liquids HHV                                | 6,700,000[^10] | 2000 | 1    | 1.26 | -    | [19]|
|                                  | Are 400 – besides the reactor 113.17 t syngas in/h                          | 4,512,300      | 2007 | 1    | 1.09 | -    | [3] |
| Hydroprocessing                  | Hydrocracking/Hydrotreating Units 0.159 m³ fuel/day                         | 4,000          | 2007 | 1    | 1.09 | 2.31 | [29]|
|                                  | Storage tanks 1000 m³                                                        | 314,700         | 2014 | 0.51 | 0.99 | 2.31 | [30]|
| Steam and Power Generation       | Water treatment system 216 t water/h                                         | 2,465,000      | 2007 | 0.7  | 1.09 | 1.15 | [1] |
|                                  | Boiler feed water, steam, and condensate treatment 216 t water/h            | 664,000/290,000 | 2007 | 0.3  | 0.65 | 1.09 | 2.42 | [1] |
|                                  | Back-pressure steam turbines/generators 44 MWe                             | 13,150,000      | 2011 | 0.7  | 1.39 | 1.8  | Based on industry information |
|                                  | Condensing extraction turbines/generators 44 MWe                           | 25,500,000      | 2011 | 0.7  | 1.39 | 1.8  | Based on industry information |
|                                  | Gas turbine/HRSG/compressor 1 kWe                                           | 258            | 2014 | 1    | 0.99 | -    | Based on industry information |
1.2. Gasification Area

In gasification process, a carbonaceous material is partially oxidized with air, steam or oxygen at high temperatures, producing a gas mixture [32]. Drying, devolatilization, combustion, and reduction stages take place during gasification [3]. Initially, as the feedstock is heated, the residual moisture is removed. Then, the volatile matter is released, and the feedstock is converted to char, the residual solids consisting of organic and inorganic materials. Depending on the feedstock, the volatiles may include H₂O, H₂, N₂, O₂, CO₂, CO, CH₄, H₂S, NH₃, C₂H₆ and very low levels of
unsaturated hydrocarbons such as acetylenes, olefins and aromatics and tars [24]. In combustion step, carbon dioxide and carbon monoxide are formed from carbon and oxygen, releasing the thermal energy for the reduction reactions to take place. The reduction reactions include water gas reaction, Boudouard reaction, water-gas shift reaction, and methanation reaction [3], represented by equations 5, 6, 7 and 8, respectively:

$$C + H_2O \rightleftharpoons CO + H_2 \quad \text{(5)}$$

$$C + CO_2 \rightleftharpoons 2CO \quad \text{(6)}$$

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \quad \text{(7)}$$

$$CO + 3H_2 \rightleftharpoons CH_4 + H_2O \quad \text{(8)}$$

In this study, the dried LCM (10 wt.% moisture) is partially oxidized with steam in a low-pressure indirectly heated circulating fluidized bed gasifier. According to Zhu et al. [13], the directly heated gasifier requires considerably more capital than the indirectly heated. The former’s total purchased equipment cost is nearly 41% higher than that associated to the latter, mostly due to air separation unit and to the reactor itself. Fluidized beds have the advantages of high carbon conversion efficiencies, good heat and material transfer between the gas and solid phases, resulting in the best temperature distribution, they tolerate wide variations in fuel quality and a broad particle size distribution, and can scale up easily [33]. The maximum gasifier capacity of 2,000 tonnes of dry biomass per day (83.3 t/h) is assumed. If the total capacity of thermochemical plant exceeds this capacity, parallel units are considered. The feasibility of this maximum capacity in one single train is an assumption in view of the lack of information about large scale commercial implementations of this technology. A similar processing capacity is assumed in other recent publications in this area [34,35].

Superheated steam at 310 °C and 2.5 bar is injected into the gasifier, acting as a fluidizing medium and reactant at the high gasifier temperatures. An advantage of using steam instead of air in the gasifier is that there is no dilution of syngas by N₂. To solve this problem, it is also possible to use pure oxygen, however, oxygen production is expensive, evidencing a great advantage of steam gasifiers [19]. On the other hand, most of them operate at low pressure, requiring compression of the product gas for downstream processes [1].

Pilot gasifiers were developed for biomass gasification, including the Battelle Columbus Laboratory’s (BCL) low pressure indirectly heated gasifier. Original data from this test facility were reported by Feldmann et al. [36]. The gasifier output data were modeled as a function of its temperature, and the data and original correlations are presented by Bain [37]. The BCL’s data and derived correlation models were used by several authors in techno-economic analyses conducted in the US National Renewable Energy Laboratory (NREL) [1,31,38,39]. These correlations were fitted from wood biomass data, however in this work it was assumed that the results for sugarcane LCM has similar behavior and the same correlations are applied. Then, the total amount of syngas produced and the quantities of its components CO₂, CH₄, C₂H₄, C₂H₆, H₂, C₂H₂ are determined, as well as the amount of tar and the water conversion. Although there is a correlation for the CO content in dry gas, it is in fact calculated to ensure that the sum of syngas components would be 100%. However, this result must always be compared with that one predicted by the correlation.

The water input flow considers both biomass moisture and injected steam. Thus, through water conversion data (obtained from BCL’s correlations), it is possible to estimate the amount of water in the gasifier output. The total amount of carbon in syngas and tar (approximated to C₆H₈) can be determined. Considering the total flow of carbon atoms present in biomass feedstock, the residual carbon is attached to the char. The material balance of atomic species is also conducted for oxygen, sulfur, nitrogen, chlorine, and hydrogen. The difference between the oxygen present in biomass and in the syngas and water is accounted for in the char. It is considered that 8.3% of sulfur and 6.6% of nitrogen are attached to the char, and all remaining sulfur and nitrogen are set as H₂S and NH₃ in the
syngas, respectively [1]. All the chlorine is assumed to be set as HCl in syngas. Finally, the hydrogen balance determines the amount of this species in the char. All ash is parsed to the char.

The solid carbon-rich char produced in the gasifier is reacted with air in a separated combustor to provide the energy for the endothermic steam gasification reactions. In order to produce enough char to provide all the heat necessary for gasification, indirect gasifiers are normally operated at relatively low temperatures, 704 to 871 °C [37]. Assuming that no supplemental fuel is delivered to the combustor, the thermal equilibrium between the gasifier and combustor is reached based on the amount of char formed, which is an inverse function of temperature. Lower temperatures imply more char formed and, thus, more heat generated in the char combustion. It increases the gasifier temperature and less char is produced. In the equilibrium temperature, the produced char is enough to satisfy the energy requirements of the operation. The equilibrium temperature applied in this assessment is the result obtained by Dutta et al. [1]: 869 °C. The heat generated in the combustor is transferred to the gasifier by the recirculation of a hot synthetic sand (olivine), containing enstatite (MgSiO3), forsterite (Mg2SiO3), and hematite (FeO) [1].

Two stages of cyclone separators at the exit of the gasifier are assumed. The olivine and char presented in the syngas are separated and fed to the char combustor. A similar cyclone system is also present at the combustor exit. The olivine from combustor is separated from the combustion flue gas in the primary cyclone and sent back to the gasifier. The char combustor flue gas also passes by a secondary cyclone that separates the residual fines (ash and olivine fines), which are cooled, wet and sent to the field [1,28]. Fresh olivine is added to compensate the cyclones’ losses. A small amount of magnesium oxide (MgO) is added to the fresh olivine to interact with the silicate compounds and sequester the potassium, which avoids glass formation, and the bed medium to agglomerate and defluidize [1]. After passing through the cyclone system, hot flue gas from the char combustor is utilized for heat recovery in feedstock drying (preprocessing) and to superheating medium pressure (MP) steam. Then, it is directed to the flue gas scrubber for residual particulates and SO2 removal.

The gasifier’s power output, $Q$ (MWth), is related to the amount of fuel fed into the gasifier and the amount of gasifying medium. The relation between the volume flow rate of the product gas, $V_g$ (Nm$^3$/s), its lower heating value, LHV$_g$ (MJ/Nm$^3$), and power output is determined by equation 9 [32]:

$$V_g = \frac{Q}{LHV_g} \quad (9)$$

The biomass feed rate, $M_{bm}$, the required power output (Q), the biomass lower heating value (LHV$_{bm}$) and the gasifier efficiency, $\eta_{gef}$, are related as shown in equation 10 [32]:

$$M_{bm} = \frac{Q}{LHV_{bm} \eta_{gef}} \quad (10)$$

The syngas lower heating value, LHV$_g$, can be estimated from equations 4 and 5. The volume of syngas, in Nm$^3$, i.e. 0 °C and 1 atm, is estimated using Ideal gas law from its molar flow. It is also possible to calculate syngas volume from BCL’s derived correlation models [37]. Table S3 summarizes the principal parameters assumed in the process simulation of gasification area. The reference quotation for capital costs [1] comprises biomass feeding system, gasification process reactors (including gasification and combustion reactors, as well as thermal tar reformer for gas conditioning, that will be described in the next section), cyclones and interconnecting ducting, combustion air blower, start-up equipment, syngas polishing scrubber for cleanup and particulate removal (that will be discussed in the next section), ash discharge system, steam generation system, emergency flare system, process instrumentation, olivine and magnesium oxide makeup system and inert gas system. Since gas cleanup equipment are also included, and following the methodology adopted by Dutta et al. [1], two-thirds of the total purchased equipment costs (calculated from Table S2 information) is allocated to gasification area. One-third is allocated to the tar reformer, syngas cooling, quench and scrubbing systems (part of syngas cleaning and conditioning area).
1.3. Syngas Cleaning and Conditioning Area

Syngas from gasifier contains contaminants that must be removed, such as particulates, tars, sulfur and nitrogen compounds. Cleanup requirements depend on the following conversion technologies and other downstream equipment. In FT synthesis, these contaminants can poison the catalyst and a severe clean is essential. The tolerance for sulfur compounds are around 1ppmv; 10 ppmm for NH₃; 0.2 ppmv for NOₓ and 10 ppb for halides [2]. In the configuration assumed in this study, syngas cleaning and conditioning includes the reformer of tars, methane and other hydrocarbons to CO and H₂; particulates and other contaminants removing by a quenching and scrubbing process; syngas compression; acid gas removal via an amine system; and LO-CAT® hydrogen sulfide oxidation, isolating the carbon dioxide and recovering solid sulfur.

The tar reformer is an entrained-flow, fluidized catalytic reactor. Tar, methane and other light hydrocarbons are converted to CO and H₂, while NH₃ becomes N₂ and H₂ [1,2,32]. These reactions are represented in equations 11 to 16:

\[
\begin{align*}
\text{CH}_4 + H_2O & \rightarrow 3H_2 + CO + 206.1 \text{ kJ/mol} \\
C_2H_4 + 2H_2O & \rightarrow 4H_2 + 2CO + 210.3 \text{ kJ/mol} \\
\text{Tar (C}_{10}\text{H}_8) + 10H_2O & \rightarrow 14H_2 + 10CO + 1162 \text{ kJ/mol} \\
C_2H_6 + 2H_2O & \rightarrow 5H_2 + 2CO + 347.3 \text{ kJ/mol} \\
C_2H_2 + 2H_2O & \rightarrow 3H_2 + 2CO + 35.9 \text{ kJ/mol} \\
NH_3 & \rightarrow \frac{3}{2}H_2 + \frac{1}{2}N_2 + 46.1 \text{ kJ/mol},
\end{align*}
\]

where each heat of reaction is estimated from the heat of formation of products and reagents [4,32]. The syngas must be reformed previously to its cooling, so tar condensation and deposition on downstream equipment are prevented. The material balance of tar reformer is conducted using NREL’s target conversions [1] presented in Table S2. The reforming catalyst (Ni/Mg/K supported on alumina) has significant water-gas-shift activity, consequently the syngas exits the reformer at water-gas-shift equilibrium [1]. The optimum H₂/CO molar ratio in the syngas for FT synthesis is 2.1 [3]. Due to a less than the optimum H₂/CO ratio from the gasifier, steam is injected into the tar reformer to shift the equilibrium and adjust this ratio. Then, shift reaction (equation 7) converts carbon monoxide into hydrogen and it is slightly exothermic (\(\Delta H = -41.2 \text{ kJ/mol} \))

The equilibrium constant \(K_{\text{equilibrium}}\), at 1 atm, can be determined using the the Gibbs free energy \(\Delta G^0\), the temperature T (in K) and the gas constant R [32], according to equation 17. It can also be associated to the molar flow ratio of the species in equilibrium:

\[
K_{\text{equilibrium}} = \exp\left(-\frac{\Delta G^0}{RT}\right) = \frac{nCO_{\text{equil}}}{nCO_{\text{react}}} \frac{nH_2_{\text{equil}}}{nH_2_{\text{react}}}
\]

Gibbs free energy can be calculated from equation 18 [32]:

\[
\Delta G^0(\frac{\text{kJ}}{\text{mol}}) = -32.197 + 0.031T - \left(\frac{1774.7}{T}\right)
\]

Although the pressure considered in the reformer is a little superior than 1 atm, the shift reaction is relatively insensitive to changes in pressure because there is no volume change, and the same equations are assumed. Then, the necessary amount of injected steam to achieve the required H₂/CO ratio in the equilibrium can be estimated by associating equation 18 and the molar balance equations written as functions of the extent of reaction.
The energy necessary for the reforming endothermic reactions (eq. 11 to 16) is provided by the circulating catalyst between tar reformer reactor and the catalyst regenerator. Air is fed into the catalyst regenerator reactor to provide the oxygen for combustion of carbon (coke) deposits on the catalyst particles, regenerating their activity and suppling heat for the reforming reactions. Since the amount of coke is not enough to sustain those reactions, supplemental unconverted syngas from FT synthesis area are also supplied to the regenerator [1]. The catalyst and combustion flue gases from the regenerator are separated in a two-stage cyclone. The catalyst flows back to the reformer and hot flue gases (969 °C) are used in heat integration: for superheating the 2.5 bar steam up to 310 °C to be injected into the reformer; for heating syngas up to the temperature of ZnO bed operation (fuel synthesis area); and for superheating steam from steam turbine 1 to steam turbine 2 (steam and power generation area).

Hot syngas is cooled to 124 °C in a high pressure (HP) heat recovery steam generator (HRSG), where 90 bar steam is generated, then superheated to 520 °C and sent to steam turbine 1 for power generation. Syngas is finally cooled to 60 °C using cooling water supply (CWS) and sent to the scrubbing system to remove particulates, ammonia, and residual tar. The scrubbing system is composed of a venturi scrubber, a cyclone separator, and a quench water recirculation system [1]. Sulfuric acid (H₂SO₄) is added to react with NH₃, improving its removal and reaching a cleaning efficiency of 99.9% [19]. The amount of H₂SO₄ feed into the system is stoichiometric to NH₃ initial content. Quench water circulation rate is 1 liter per cubic meter of gas flow [20]. The purge stream, which also includes process condensate, is sent to a clarifier to separate the solids from the water. The former is sent to disposal in field, and the latter is treated.

Due to the low-pressure gasifier, the use of a process gas compressor is required to raise the syngas pressure to the level appropriate for downstream treatment and FT synthesis. The syngas is compressed to around 30 bars using a 4-stage centrifugal compressor with interstage cooling [2]. Any residual condensate in the syngas exiting the scrubber is removed in the compressor drum until water mass fraction of 0.004. Syngas leaves this section at 43 °C [2].

The following step for syngas cleanup is the removal of carbon dioxide and hydrogen sulfide by an amine-based absorption system. A typical amine system consists of an absorber, a stripper column, a flash separator and heat exchangers [2]. The acid gases (CO₂ and H₂S) are removed from syngas stream in the absorber by chemical reactions with the amine-based solvent. In this study it is considered the use of a 30% concentrated diethanolamine (DEA) solution as the absorbent, that can absorb 0.6 mol of acid gas per mol of amine [21]. Among commercially available amine solution, DEA can be used at higher concentrations than monoethanolamine (MEA) and has the advantages of absorbing more acid gas per solution circulated, resulting in reduced the circulation rate e energy savings, and, therefore, lowering capital and operational costs. Moreover, it is also generally less corrosive than MEA and it is very resistant to degradation from RSH and COS [21]. DEA is used in medium-to-high-pressure systems [2]. It was considered the H₂S removal of 99.9% and 92% for CO₂ [2]. The sweet syngas exiting at the top of the absorber is sent to fuel synthesis area. The rich amine is regenerated through the flash separator, that removes absorbed hydrocarbons, and by a stripper column, that desorbs the H₂S and CO₂. Then, amine solution is cooled and returns to the absorber. The stripped acid gas stream is directed to the sulfur recovery unit [2].

The LO-CAT® sulfur recovery system is designed to isolate H₂S from CO₂ and convert it to solid sulfur. The process consists of an absorber section, an oxidizer for catalyst regeneration, and a sulfur management unit [2]. In the absorber, H₂S is oxidized to elemental sulfur using chelated iron as the catalyst. The overall process reaction is represented by equation:

\[
H_2S + \frac{1}{2} O_2 \rightarrow H_2O + S_{(Fe)}^0
\]

The LO-CAT® unit is designed to remove over 99.9% of the H₂S [22]. The spent catalyst is regenerated in the oxidizer by reacting with oxygen from air. Elemental sulfur is concentrated into a sulfur slurry, which is washed to recover any dragged catalyst. However, it still contains residual
catalyst and is classified as a low-value sulfur and undesirable as a chemical feedstock [2]. Since it is appropriate for agricultural purposes, the sulfur is sent to field and the revenues from its sale are not considered in this study.

Main parameters assumed for syngas cleaning and conditioning area modeling are summarized in Table S2. Capital costs are estimated based on data presented in Table S3. It is assumed an extra 10% of the one-third value allocated for reforming section and scrubbing in order to include heat integration.

### 1.4. Fuel Synthesis Area

Operations in fuel syntheses area involve zinc oxide (ZnO) gas polishing, hydrogen separation via pressure swing adsorption (PSA), FT synthesis, FT products separation, and unconverted syngas distribution. ZnO is often used as a catalytic absorbent for sulfur removal in a gas polishing step. By reaction with ZnO absorbent, H₂S is converted to zinc sulfide (eq. 20) and HCl forms a stable chloride [2].

\[
\text{ZnO} + \text{H}_2\text{S} \rightarrow \text{ZnS} + \text{H}_2\text{O} \tag{20}
\]

Sweet syngas coming from the amine-based acid gas removal is heated to 349 °C [2] by heat integration with hot flue gas from the syngas cleaning and conditioning area regenerator. A pressure drop of 1.37 bar is assumed [3]. ZnO absorbent cannot be regenerated, so it must be frequently replaced and disposed. A sulfur removal below 50 ppb is realistic with ZnO [2]. Just for simulation purposes, residual H₂S and HCl are completely removed. After passing throughout the ZnO beds, the syngas is cooled to 200 °C, adjusting the temperature for FT synthesis. Part of the cooling energy of the syngas is delivered to 2.5 bar-steam to superheat it before inserting it into the gasifier. The last part of the energy is delivered to cooling water, that arises the temperature from 27 to 38 °C.

The PSA unit isolates a stream of hydrogen to be used in hydrotreating facilities. A small percentage of syngas is departed to PSA unit since only a small amount of hydrogen is required. It is considered an efficiency of hydrogen removal around 85%, producing a pure H₂ stream [3]. The H₂/CO ratio leaving the reformer must be set a little above 2.1 to reach excess hydrogen for this separation. After the PSA, the syngas deviated is recompressed, rejoins the main syngas stream, and enters the FT reactor.

The FT catalytic reaction produces hydrocarbons of variable chain length and it is main mechanism is represented by equation 21 [19]:

\[
\text{CO} + 2.1 \text{H}_2 \rightarrow -\text{(CH}_2\text{)}_n - +\text{H}_2\text{O} - 165 \text{KJ/mol} \tag{21}
\]

The product carbon number distribution from a FT process can be estimated using the Anderson-Schulz-Flory (ASF) distribution. The yield of a hydrocarbon with length \(n\) is determined by the chain growth probability (\(\alpha\)) model, that is, the chance that a hydrocarbon chain grows with another -CH₂- group, instead of terminating, as described in the equation 22 [19]:

\[
x_n = \alpha^{n-1} (1 - \alpha) \tag{22}
\]

As the temperature decreases, longer hydrocarbon chains are formed. There are two FT operating modes depending on the desired product distribution. The high-temperature (300–350 °C) Fischer-Tropsch (HTFT) process with iron catalysts is used to produce gasoline-range and low molecular mass products predominantly linear and rich in olefins and oxygenates. For diesel-range and high molecular mass linear waxes products, the low-temperature (LTFT) (200–240 °C) process with either iron or cobalt catalysts is used [3,19,23,40]. Long-carbon-chain wax products, favored at low temperatures, require further hydrocracking to the diesel range. On the other hand, at high temperatures, selectivity favors methane and light gases, that is a disadvantage if the focus is liquid fuel production [3], as in the case of this study. Besides, the great amount of oxygenates in HTFT derived syncrude makes product stabilization a necessary step [41].
Since the FT reactions are highly exothermic (see eq. 21), the heat of reaction must be removed, or the catalyst can be deactivated due to sintering and fouling as well as resulting in high undesirable methane production [40]. Consequently, heat removal has been the focus of reactor design development [3]. The tubular fixed bed reactor (TFBR) and the slurry-phase reactor (SPR) are the most promising types [40]. Catalyst is kept on the tube side in fixed-bed reactors and heat removal is accomplished by steam generation on the shell side [42]. There is a high pressure drop across the reactor. In top-fed multi-tubular reactors the wax produced comes down and leaves the catalyst bed [40]. Thus, the TFBR is well suited for wax production due to simple liquid/wax removal and it is simple to operate. In the slurry-phase reactor (SPR) the syngas is bubbled through the reactor from the bottom and the catalyst is suspended in a liquid. Its operation is more complex and wax removal is difficult [3]. However, the cost is only 25% of that of multi-tubular system [40].

FT process is generally conducted at pressures between 10 and 40 bar. In this study, it is assumed FT synthesis at 25 bar and using a cobalt catalyst [3]. The FT reactor inlet temperature is 200 °C and outlet temperature is 240 °C [19]. A multi-tubular fixed bed type reactor is selected since a large amount of wax is produced, considering the operating conditions. The assumed per-pass carbon monoxide conversion is 40% [3]. The product distribution follows the ASF distribution (eq. 22) where chain growth factor, $\alpha$, can be estimated from equation 23 for cobalt catalyst [43]:

$$\alpha = \left(0.2332 \frac{y_{CO}}{y_{CO} + y_{H_2}} + 0.6330\right) \times (1 - 0.0039(T - 533)),$$

(23)

where $y$ is the molar fraction of carbon monoxide or hydrogen and $T$ is the reactor temperature in kelvin. The methodology for estimating the material flow of each hydrocarbon chain (C1 until C25+) is based on the calculation of molar reaction extent for each corresponding reaction and detailed steps are described by Swanson et al. [3]. The exiting effluent is cooled from 240 °C to 35 °C and the liquid water and hydrocarbons are separated in a gas/liquid knock-out separator [3]. Part of energy released from cooling is recovered for condensate and makeup water pre-heating. Another part is used for heating FT recycle and for heating medium pressure (MP) boiler feed water (BFW) to saturation temperature. The remaining energy is delivered to cooling water.

Due to the great amount of waxes in the FT liquids, it is required hydrocracking operation in the hydroprocessing area before the final fuel blend is molded. Unconverted syngas is split into a recycle stream to the FT reactor, a supplemental combustion gases stream to the catalyst regenerator in syngas cleaning and conditioning area, and a stream to the gas turbine. The recycle is compressed and heated until 200 °C before being fed into the FT reactor. For recycle heating is used part of energy released from FT products cooling. The necessary amount of FT cobalt catalyst is estimated considering the total volume of gas flow into the reactor, a gas hourly space velocity of 100 hr$^{-1}$ and a replacement period of 3 years [3]. Table S2 summarizes the main parameters of fuel synthesis area. Capital costs are estimated based on data presented in Table S3.

### 1.5. Hydroprocessing Area

FT product contains a significant amount of high molecular weight waxes that is cracked using hydrogen from PSA. Klerk [23] describes different designs of refineries to optimize the production of on-specification motor-gasoline, jet fuel and diesel fuel from both HTFT and LTFT syncrude. A refinery design for LTFT syncrude, after initial FT liquid distillation, is selected from Klerk’s scenarios [23] for the present study modelling. This area is modeled as a “black box” considering the final product distribution in Table S2. The amount of carbon and hydrogen in the FT product stream is calculated. The carbon mass flow is the same as that of the final blend stream flow. Thus, based on carbon balance, the amount of each component is calculated from blend fractions in Table S2. The amount of hydrogen in the final blend is then calculated. The difference between hydrogen in FT product stream and hydrogen in the final blend is determined. FT products are hydrogen deficient compared to the final blend, so hydrogen is required in hydroprocessing. The amount of syngas deviated to the PSA unit is varied to achieve the required hydrogen stream to hydroprocessing.
Even though Klerk’s study considers European specifications for motor-gasoline and diesel fuel, these fuels remain on-specification comparing with equivalent Brazilian National Agency of Petroleum, Natural Gas and Biofuels specifications [44,45]. The ASTM standard specification for aviation turbine and fuel containing synthesized hydrocarbons is considered for Jet A-1[46]. Part of the fuel gases from hydroprocessing (that includes hydrogen, fuel gas and liquid petroleum gas fractions presented in Table S2) are burned in fired heaters to meet the area’s thermal demand. Flue gas from fired heaters moves to flue gas scrubber. The remaining part of hydroprocessing fuel gases feeds the gas turbine. The main parameters of this area are listed in Table S2 and capital costs are calculated based on data from Table S3.

1.6. Steam and Power Generation Area

A gas turbine and steam turbines association provide the power and heat required throughout the gasification plant in an arrangement known as integrated gasification combined cycle (IGCC) power generation. An IGCC plant can reach higher overall efficiency than a combustion plant with a steam turbine only [32]. Unconverted syngas from FT synthesis and fuel gas from hydroprocessing area and compressed air (20 bar) react in the combustor [3]. Hot pressurized flue gases expand in the gas turbine, producing shaft work. Exiting flue gases still have high temperatures, exchanging heat with high pressure (HP) BFW in a heat recovery steam generator (HRSG) to produce steam at 90 bar. HP steam is also produced through recovery of heat from syngas cooling after tar reformer [2]. Steam at 90 bar is superheated to 520 °C. MP steam (25 bar) is generated on the shell side of the FT reactor and superheated to 360 °C via heat recovery of flue gas from char combustor. The steam cycle comprises two stages of steam turbines with intermediate reheat [1]. In the first stage, a back-pressure turbine (steam turbine 1) drops the steam pressure from 90 to 25 bar, producing shaft work. MP steam from steam turbine 1, after reheated to 360 °C, and from shell side of FT reactor feed the condensing extraction turbine (steam turbine 2). The amount of process steam at 4 bar and 2.5 bar extracted is controlled to meet the process demand and the remaining steam is directed to the condensing stages of the turbine to produce power. The vacuum steam is directed to the condenser.

Gas turbine net electric output is calculated using equation 24 [7]:

\[\eta_{ overall} = \frac{\text{net electric output}}{\text{fuel output}},\]  

considering a cycle overall efficiency (\(\eta_{overall}\)) of 40% [7,24,25]. Compressor electric requirements are estimated from Aspen Plus® simulation. The steam turbines isentropic efficiencies are assumed to be 85% and generator efficiency is 98% [26]. Thus, the shaft work (W) can be estimated from equation 25:

\[W = 0.98 \times m_{steam} \times (H_{steam \text{ in}} - H_{steam \text{ out}}),\]  

where \(m_{steam}\) is the mass of steam extracted and \(H\) is the specific enthalpy of steam in the conditions of temperature and pressure at the entrance and exit of the turbine. Steam production must supply the thermal demand of the thermochemical plant and the integrated 1G ethanol distillery. The resulting power generated supplies electricity requirements of the plant and the excess is exported as a coproduct. In sugarcane off-season, the demand of steam at 2.5 bar is inferior, consequently more steam is directed to condensing stages in steam turbine 2 and more power is generated. This increase is also estimated in scenarios without integration between thermochemical plant and 1G ethanol distilleries. Thermochemical operations requiring steam at 2.5 bar include preprocessing, gasification, reforming, sulfur recovery and reheating the recycle of FT reactor. Acid gas removal unit demands 4 bar-steam.

When indirect heat exchange is applied, the process condensate is collected, as well as the steam turbine condensate. Then it is treated, pre-heated and mixed with treated and pre-heated makeup water, as represented in Figure S2. Makeup water treatment system includes reverse osmosis and electrodeionization [1]. The water is sent to the deaerator to remove any dissolved gases. Before steam generation, the water from the deaerator is pumped to 25 or 90 bar and pre-heated to its saturation
temperature. HP and MP steam are then generated by hot syngas and gas turbine flue gas heat recovery, and on the shell side of the FT reactor, as described. Saturated steam is collected in steam drums and a blowdown avoids solids accumulation [1]. Saturated steam is superheated before it goes to the steam turbines. The main considerations for material and energy balances of steam and power generation area are described in Table S2. Capital costs data are presented in Table S3.

1.7. Utilities Area

The cooling tower provides cooling water to the plant (Figure S2). Water losses from the cooling water system include evaporation, drift, and tower basin blowdown [1]. The parameters for the cooling water system’s design are shown in Table S2. A compressed air system, flue gas scrubbing system, water treatment and storage of chemicals are also included for capital costs estimates (Table S2). The amount of NaOH required in the flue gas scrubber is calculated related to SO2 removed.

1.8. Additional Economic Modeling Information

The main parameters considered in the economic analysis and the basis for variable operating costs (adjusted to Dec 2017 US$) are summarized in Tables S4 and S5. The parameters considered for risk assessment are detailed in Table S6.

| Table S4. Main parameters considered in the economic analysis. |
|---------------------------------------------------------------|
| **Parameter** | **Value** | **Reference** |
| Expected plant lifetime | 25 years | [8] |
| Discount rate | 12% | [8] |
| Reference data | December 2017 | Assumption |
| Exchange rate | 3.29 R$/US$ | Market data |
| Depreciation | 10 years, linear | [8] |
| Working capital | 10% of fixed capital investment | [8] |
| IRPJ + CSLL | 25% + 9% of taxable income | [8] |
| Maintenance | 3% of fixed capital investment | [1] |
| Insurance and other | 0.7% of fixed capital investment | [1] |
| Hydrous ethanol price1 | 0.48 US$/L | [47] |
| Gasoline price1 | 0.51 US$/L | [48] |
| Jet fuel price1 | 0.60 US$/L | [48] |
| Diesel price1 | 0.57 US$/L | [48] |
| Electricity price1 | 60.54 US$/MWh | [49] |
| Carbon price | 10 US$/t CO2-eq avoided | [50] |
| Salaries (including benefits and overhead) | US$3292.21 (2G) and US$880.41 (1G) | [51] |

1 Selling price for biorefinery products to the producer based on Brazilian historical market data, without taxes.

| Table S5. Basis for estimating industrial chemicals variable operating costs1. |
|---------------------------------------------------------------|
| **Variable** | **Cost information** | **Reference** |
| Olivine 2 | US$ 363.8/t | [1] |
| MgO 2 | US$ 740.4/t | [52] |
| Tar reformer catalyst 2 | US$ 93.1/kg | [1] |
| H2SO4 3 | US$ 0.31/kg | [53] |
| DEA make-up 3 | US$ 1.43/t | [54] |
| LO-CAT® chemicals 2 | US$ 250.4/t Sulfur | [3] |
Table S6. Ranges in parameters (triangular distributions) considered for uncertainty and risk assessment.

| Parameter                        | Unit                  | Minimum | Value | Maximum | Reference       |
|----------------------------------|-----------------------|---------|-------|---------|-----------------|
| CAPEX                            | % of basis CAPEX      | 80%     | 100%  | 130%    | Assumption      |
| Industrial inputs cost           | % of basis cost       | 70%     | 100%  | 130%    | Assumption      |
| Labor cost                       | % of basis cost       | 70%     | 100%  | 130%    | Assumption      |
| CS yield                          | TC/ha/year            | 60      | 80    | 100     | [56]            |
| EC yield                          | TC/ha/year            | 100     | 140   | 200     | [57,58]         |
| Eucalyptus yield                 | m³/ha/year            | 23.0    | 46.14 | 58.0    | Assumption      |
| Diesel consumption               | % of basis consumption| 80%     | 100%  | 120%    | Assumption      |
| Fertilizer consumption           | % of basis requirement| 80%     | 100%  | 120%    | Assumption      |
| CS stalks cost                   | % of basis cost       | 80%     | 100%  | 120%    | Assumption      |
| Sugarcane straw cost             | % of basis cost       | 80%     | 100%  | 120%    | Assumption      |
| EC cost                          | % of basis cost       | 80%     | 100%  | 120%    | Assumption      |
| Eucalyptus cost                  | % of basis cost       | 80%     | 100%  | 120%    | Assumption      |
| Diesel price                     | US$/L                 | 0.32    | 0.57  | 0.71    | [48]            |
| Gasoline price                   | US$/L                 | 0.26    | 0.51  | 0.66    | [48]            |
| Jet fuel price                   | US$/L                 | 0.29    | 0.60  | 0.94    | [48]            |
| Electricity price                | US$/MWh               | 27.20   | 60.54 | 99.25   | [49]            |
| Hydrous ethanol price            | US$/L                 | 0.29    | 0.48  | 0.65    | [47]            |
| Carbon price                     | US$/tCO₂eq avoided    | 0       | 10    | 50      | [50]            |

1 Only in CE production system.

Table S7. Selected biomass production system technical results of evaluated scenarios.

| Agricultural information | 1G-CS | 1G2-CS | 1G2G-EC |
|--------------------------|-------|--------|---------|
| CS (and straw)           | CS (and straw) | Eucalyptus | EC | Eucalyptus |

2. Detailed Technical Results

The main technical results from simulations are presented in Tables S7, S8 and S9.
### Key Technical Results of the Thermochemical Processes

#### Table S8

| Area | Response | 1G2G-CS | 1G2G-EC |
|------|----------|---------|---------|
|      |          | Season  | Off-season | Season | Off-season |
| 1G   | Processed biomass | CS      | -        | EC     | -         |
|      | Required electric power (MW) | 26.6    | -        | 25.3   | -         |
|      | Fiber in mills (t/h) | 124.6   | -        | 177.6  | -         |
|      | Imbibition water in mills (t/h) | 267.9   | -        | 382.0  | -         |
|      | Sugar extraction in mills (%) | 95    | -        | 87     | -         |
|      | 2.5 bar steam consumed (t/h) | 260.2   | -        | 399.2  | -         |

**Preprocessing**

| Processed biomass | Eucalyptus | Eucalyptus |
|-------------------|------------|------------|
| CS bagasse, CS straw and Eucalyptus | Eucalyptus | Eucalyptus |

| Total LCM processed (dry t/h) | 203.6 | 224.0 |
| Total LCM processed (MW) | 948.2 | 942.1 | 1044.1 | 1036.2 |
| Steam from rejected moisture (t/h) | 140.0 | 141.1 | 189.0 | 155.2 |
| Required electric power (MW) | 0.3 | 1.1 | 0.2 | 1.2 |

**Gasification**

| H₂/CO (molar) | 0.57 | 0.57 |

---

1 Including ash and filter cake transport. 2 Including biomass transportation to the biorefinery.
| Energy Parameters                  | Value 1       | Value 2       | Value 3       | Value 4       |
|-----------------------------------|---------------|---------------|---------------|---------------|
| Parallel gasifiers (number)       | 3             | 3             |               |               |
| Qth (1 gasifier) (MWth)           | 222.4         | 221.2         | 244.7         | 243.2         |
| LHVgas (daf) (MJ/Nm³)             | 15            | 15            | 15            | 15            |
| Gasifier efficiency LHV basis (%) | 70.4          | 70.4          | 70.3          | 70.4          |
| Gasifier efficiency HHV basis (%) | 76.5          | 76.5          | 76.4          | 76.5          |
| Syngas composition (mole %) (dry) |               |               |               |               |
| CO                               | 41.87         | 42.16         | 41.86         | 42.16         |
| H₂                               | 23.67         | 23.83         | 23.66         | 23.83         |
| CH₄                              | 15.29         | 15.40         | 15.29         | 15.40         |
| CO₂                              | 12.75         | 12.83         | 12.74         | 12.83         |
| C₂H₄                             | 4.34          | 4.36          | 4.33          | 4.36          |
| C₂H₆                              | 0.29          | 0.30          | 0.29          | 0.30          |
| C₂H₂                              | 0.40          | 0.41          | 0.40          | 0.41          |
| NH₃                              | 0.97          | 0.38          | 1.02          | 0.38          |
| H₂S                              | 0.07          | 0.02          | 0.07          | 0.02          |
| HCl                              | 0.03          | 0.001         | 0.01          | 0.001         |
| TAR (C₁₀H₈)                      | 0.31          | 0.31          | 41.86         | 0.31          |
| Required electric power (MW)      | 1.7           | 1.7           | 1.9           | 1.9           |
| Syngas Cleaning and Conditioning  |               |               |               |               |
| Supplemental combustions gases LHV (regenerator) (MJ/kg) | 25.5 | 26.1 | 21.9 | 22.1 |
| Supplemental combustions gases flow (regenerator) (t/h) | 16.0 | 15.8 | 21.1 | 20.8 |
| Required electric power (MW)      | 53.4          | 53.3          | 58.3          | 58.2          |
| Fuel Synthesis                    |               |               |               |               |
| Syngas fraction deviated to PSA unit (%) | 4.6 | 3.1 |               |               |
| α (T, yCO, yH₂)                  | 0.87          | 0.87          |               |               |
| CO global conversion (%)          | 96.1          | 96.3          | 63.4          | 63.6          |
| Total reaction volume (m³)        | 772           |               | 314           |               |
| Recycle ratio (mole of recycle per mole of fresh feed) | 3.5 | 0.6 |               |               |
| Required electric power (MW)      | 11.0          | 10.9          | 1.3           | 1.3           |
| Hydroprocessing                   |               |               |               |               |
| FT transportation fuel yield (wt. %) | 87.8 |     | 87.8          |               |
| Required electric power (MW)      | 5.2           | 5.0           | 3.6           | 3.6           |
| Steam and Power Generation        |               |               |               |               |
| Total generated electric power (MW) | 160.9 | 177.3 | 330.0 | 367.6 |
| 2.5-bar steam extracted (ton/h)   | 284.7         | 24.7          | 383.2         | 19.4          |
| 4-bar steam extracted (ton/h)     | 159.6         | 160.6         | 171.2         | 172.4         |
| Steam to condensing stages of steam turbine (ton/h) | 2.9 | 259.6 | 0.4 | 361.1 |
| Required electric power (MW)      | 43.6          | 42.8          | 113.2         | 112.4         |
| Utilities                         |               |               |               |               |
| Make-up water (m³/h)³             | 1,745.0       | 491.7         | 1,697.7       | 478.2         |
| Required electric power (MW)⁴    | 9.1           | 9.1           | 8.3           | 8.1           |
| Global                            |               |               |               |               |
| 2.5 bar steam consumed (t/h)      | 164.3         | 165.8         | 173.0         | 174.6         |
| 4 bar steam consumed (t/h)        | 159.6         | 160.6         | 171.2         | 172.4         |
Required electric power (MW) | 115.2 | 114.9 | 178.7 | 178.8
---|---|---|---|---
Global Energetic efficiency (%) | 57.7 | 50.9 | 21 of 26
Total process required electric power (MW) | 141.7 | 114.9 | 204.0 | 178.8

1Excluding residual char, olivine and MgO. 2Around 93% of electric power required in the syngas cleaning and conditioning area is for syngas compression. 3To attend 1G and 2G processes demand. 4Electric consumption related to 1G makeup water pumping is discounted from this value, as it has already been counted in the electric consumption for 1G. 5Energy output (MJ per year) per energy content in biomass including 1G and 2G processes, calculated according to Furtado Júnior et al. [59].

Table S9. Life Cycle Inventory of the biorefinery scenarios (reference flows are per year of plant operation).

| Flows                                      | Unit/year | 1G-CS | 1G2G-CS | 1G2G-EC |
|--------------------------------------------|-----------|-------|---------|---------|
| **Inputs**                                 |           |       |         |         |
| Natural resources                          |           |       |         |         |
| Air                                        | t         | 5,676,996 | 6,885,971 | 12,123,127 |
| Occupation, industrial area, built up (land) | m²a       | 0.006 | 0.016 | 0.017 |
| Transformation, from arable, non-irrigated (land) | m²       | 0.0001 | 0.0003 | 0.0003 |
| Transformation, to industrial area, built up (land) | m²       | 0.0001 | 0.0003 | 0.0003 |
| Materials/fuels                            |           |       |         |         |
| Sugarcane/Energy-cane (wet)                | t         | 4,000,000 | 4,000,000 | 4,000,000 |
| Straw (dry)                                | t         | 183,400 | 183,400 | - |
| Eucalyptus                                 | m³        | - | 2,052,199 | 2,214,819 |
| Quicklime                                  | t         | 2,550 | 2,550 | 2,429 |
| Sulfuric acid                              | t         | 1,679 | 2,783 | 2,370 |
| NaOH                                       | t         | - | 141 | 157 |
| Phosphoric acid (85% in H₂O)               | t         | - | 924 | 1342 |
| Antibiotic (1G)                            | t         | 11 | 11 | 10 |
| Flocculant polymer (1G)                    | t         | 5 | 5 | 3 |
| LO-CAT® chemicals                          | kg        | - | 73 | 81 |
| Tar reformer catalyst                      | t         | - | 87 | 93 |
| Olivine (make-up)                          | t         | - | 40,993 | 44,550 |
| FT catalyst replacement                    | t         | - | 264 | 107 |
| ZnO absorbent replacement                  | kg        | - | 19 | 21 |
| MgO                                        | t         | - | 6,959 | 8,190 |
| DEA (make-up)                              | t         | - | 321 | 345 |
| Lubricating oil                            | t         | 54 | 54 | 54 |
| Steel product manufacturing                | t         | 484 | 1,297 | 1,393 |
| Chromium steel product manufacturing,      | t         | 29 | 77 | 82 |
| Concrete (sole plate and foundation)       | m³        | 657 | 1,763 | 1,893 |
| Building, hall, steel construction         | m²        | 137 | 369 | 396 |
| Water treatment 1                          | t         | 6,000,000 | 9,848,564 | 9,556,473 |

Outputs
### Main products

| Product                  | t    | million L | GJ          |
|--------------------------|------|-----------|-------------|
| Hydrous ethanol          | 287,160 | 360 | 7,569,490  |
| Green diesel sold to market | -     | - | 59,479  |
| Renewable Jet Fuel       | 92,946 | 116 | 4,044,517  |
| Green Gasoline           | 107,346 | 145 | 4,671,133  |
| Electricity              | 751  | 286 | 2,703,195  |

### Residues sent to the field

| Residue                              | t    | million L | GJ          |
|--------------------------------------|------|-----------|-------------|
| Green diesel to agricultural phase   | 21,686 | 26 | 2,123,056  |
| Sulfur                               | 891  |  24 | 992         |
| Vinasse                              | 2,783,435 | 2,783,435 | 1,810,125  |
| Filter cake (wet)                    | 182,847 |  84 | 183,013    |
| Sand and ash (wet)                   | 64,998 | 88,800 | 95,114     |

### Emissions to air

| Emission                          | t    | million L | GJ          |
|-----------------------------------|------|-----------|-------------|
| Carbon dioxide, biogenic          | 270,452 | 2,123,056 | 2,548,088  |
| Carbon monoxide, biogenic         | 887  | 558     | 876         |
| Nitrogen oxides                   | 882  | 1071    | 2091        |
| Dinitrogen monoxide               | 49   |  33 | 54          |
| Methane, biogenic                 | 366  | 159     | 193         |

1 Related to necessary chemicals inputs. 2 Discounting diesel sent to agricultural phase.

### Supplementary References

1. Dutta, A.; Talmadge, M.; Hensley, J.; Worley, M.; Dudgeon, D.; Barton, D.; Groenendijk, P.; Ferrari, D.; Stears, B.; Searcy, E.M.; Wright, C.T.; Hess, J.R. *Process design and economics for conversion of lignocellulosic biomass to ethanol - Thermochemical pathway by indirect gasification and mixed alcohol synthesis; NREL/TP-5100-51400*. National Renewable Energy Laboratory: Golden, CO, 2011;

2. Nexant Inc *Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup, and Oxygen Separation Equipment - Task 2: Gas Cleanup Design and Cost Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas; NREL/SR-510-39945*. Work performed by Nexant Inc., San Francisco, CA. National Renewable Energy Laboratory: Golden, CO, 2006;

3. Swanson, R.M.; Satrio, J.A.; Brown, R.C.; Platon, A.; Hsu, D.D. *Techno-economic analysis of biofuels production based on gasification; NREL/TP-6A20-46587*. National Renewable Energy Laboratory: Golden, CO, 2010;

4. Perry, R.H.; Green, D. w.; Maloney, J.O. *Perry’s Chemical Engineers’ Handbook*; 7th ed.; The McGraw-Hill
Companies, Inc: New York, 1997;

5. Sandler, S.I. *Chemical, biochemical, and engineering thermodynamics*; 4th ed.; John Wiley & Sons, 2006;

6. Channiwala, S.A.; Parikh, P.P. A unified correlation for estimating HHV of solid, liquid and gaseous fuels. *Fuel* **2002**, *81*, 1051–1063, doi:10.1016/S0016-2361(01)00131-4.

7. Quaak, P.; Knoef, H.; Sfassen, H. *Energy from biomass - a review of combustion and gasification technologies*; Energy Series, World Bank Technical Paper n° 422, 1999;

8. Bonomi, A.; Cavalett, O.; Cunha, M.P.; Lima, M.A.P. *Virtual Biorefinery - An Optimization Strategy for Renewable Carbon Valorization*; Green Energy and Technology; 1st ed.; Basel: Springer International Publishing, 2016; ISBN 978-3-319-26043-3.

9. Tufaile Neto, M.A. Characterization of sugar cane trash and bagasse. In *Biomass power generation: sugar cane bagasse and trash*; Hassuani, S., Leal, M.R.L.V., Macedo, I.C., Eds.; Centro de Tecnologia Canavieira (CTC): Piracicaba, SP, Brazil, 2005; pp. 24–26.

10. Telmo, C.; Lousada, J.; Moreira, N. Proximate analysis, backwards stepwise regression between gross calorific value, ultimate and chemical analysis of wood. *Bioresour. Technol.* **2010**, *101*, 3808–3815, doi:10.1016/j.biortech.2010.01.021.

11. Baliban, R.C.; Elia, J.A.; Floudas, C.A.; Gurau, B.; Weingarten, M.B.; Klotz, S.D. Hardwood biomass to gasoline, diesel, and jet fuel: 1. Process synthesis and global optimization of a thermochemical refinery. *Energy and Fuels* **2013**, *27*, 4302–4324, doi:10.1021/ef302003f.

12. Brito, J.O.; Barrichelo, L.E.G.; Couto, H.T.Z. do; Fazzio, E.C.M.; Corradini, L.; Carrara, M.A.; Migliorini, A.J. Avaliação das características dos resíduos de exploração florestal do eucalipto para fins energéticos; Circular técnica n° 2. Piracicaba, SP: Instituto de Pesquisas e Estudos Florestais (IPF), 1979;

13. Zhu, Y.; Tjokro Rahardjo, S.; Valkenburg, C.; Snowden-Swan, L.; Jones, S.; Machinal, M. Techno-economic Analysis for the Thermochemical Conversion of Biomass to Liquid Fuels (U.S. DOE). *Doe* **2011**, 152, doi:PNNL-19009.

14. Amos, W.A. *Report on Biomass Drying Technology*; NREL/TP-570-25885. National Renewable Energy Laboratory: Golden, CO, 1998;

15. Fagernäs, L.; Brammer, J.; Wilén, C.; Lauer, M.; Verhoeoff, F. Drying of biomass for second generation synfuel production. *Biomass and Bioenergy* **2010**, *34*, 1267–1277, doi:10.1016/j.biombioe.2010.04.005.

16. LIPPEL - Soluções Integradas para Biomassa. Personal communication, 2018.

17. Delfini, P. Cana crua x Extração. In 13º Seminário Brasileiro Agroindustrial: a usina em números. *Rev. da Soc. dos técnicos açucareiros e alcooleiros do Bras.* **2012**.

18. Cardoso, T.F. Avaliação socioeconômica e ambiental de Sistemas de recolhimento e uso da palha de cana-de-açúcar, Ph.D. Thesis, University of Campinas - UNICAMP, Campinas, SP, Brazil, 2014.

19. Tijmensen, M.J.A.; Faaij, A.P.C.; Hamelinck, C.N.; Van Hardeveld, M.R.M. Exploration of the possibilities for production of Fischer Tropsch liquids and power via biomass gasification. *Biomass and Bioenergy* **2002**, *23*, 129–152, doi:10.1016/S0961-9534(02)00037-5.

20. Ravi, G.; Gupta, S.K.; Viswanathan, S.; Ray, M.B. Optimization of venturi scrubbers using genetic algorithm. *Ind. Eng. Chem. Res.* **2002**, *41*, 2988–3002, doi:10.1021/ie010531b.

21. Francis, S.M.; Richard, E.T. *Oilfield processing of petroleum: natural gas. Vol. 1*; PennWell Books: Tusla, Oklahoma, 1991; ISBN 978-0878143436.

22. Merichem Company. LO-CAT Sulfur Recovery Technology. Available online: https://www.merichem.com/technology/sulfur-recovery-with-lo-cat/.

23. Klerk, A. De Fischer-Tropsch fuels refinery design. *Energy Environ. Sci.* **2011**, *4*, 1177–1205,
24. Rezaiyan, J.; Cheremisinoff, N.P. *Gasification Technologies - A primer for engineers and scientists*; CRC Press: Boca Raton, 2005; ISBN 978-0824722470.

25. WÄRTSILÄ. Gas turbine for Power Generation: Introduction. Available online: https://www.wartsila.com/energy/learn-more/technical-comparisons/gas-turbine-for-power-generation-introduction (accessed on Mar 14, 2018).

26. Dias, M.O.S.; Junqueira, T.L.; Jesus, C.D.F.; Rossell, C.E.V.; Maciel Filho, R.; Bonomi, A. Improving second generation ethanol production through optimization of first generation production process from sugarcane. *Energy 2012*, 43, 246–252, doi:10.1016/j.energy.2012.04.034.

27. Seabra, J.E.A. Avaliação técnico-econômica de opções para o aproveitamento integral da biomassa de cana no Brasil, PhD Thesis, Universidade Estadual de Campinas – UNICAMP, Campinas, SP, Brazil.

28. Phillips, S.D.; Tarud, J.K.; Biddy, M.J.; Dutta, A. *Gasoline from Wood via Integrated Gasification, Synthesis, and Methanol-to- Gasoline Technologies*; NREL/TP-5100-47594. National Renewable Energy Laboratory: Golden, CO, 2011;

29. Robinson, P.R.; Dolbear, G.E. Commercial Hydrotreating and Hydrocracking. In *Hydroprocessing Of Heavy Oils and Residua*; Anceyta, J., Speight, J. G., Eds.; CRC Press: Boca Raton, FL, 2007; p. 376 ISBN 9780849374197.

30. Matches. Process Equipment Cost Estimates. Available online: https://www.matche.com/equipcost/Default.html (accessed on Sep 17, 2018).

31. Phillips, S.; Aden, A.; Jechura, J.; Dayton, D.; Eggeman, T. *Thermochemical ethanol via indirect gasification and mixed alcohol synthesis of lignocellulosic biomass*; NREL/TP-510-41168. National Renewable Energy Laboratory: Golden, CO, 2007;

32. Basu, P. *Biomass Gasification and Pyrolysis*; Academic Press, 2010; ISBN 9780123749888.

33. Warnecke, R. Gasification of biomass: Comparison of fixed bed and fluidized bed gasifier. *Biomass and Bioenergy 2000*, 18, 489–497, doi:10.1016/S0961-9534(00)00009-X.

34. Haarlemmer, G.; Boissonnet, G.; Imbach, J.; Setier, P.A.; Peduzzi, E. Second generation BtL type biofuels - A production cost analysis. *Energy Environ. Sci. 2012*, 5, 8445–8456, doi:10.1039/c2ee21750c.

35. Dimitriou, I.; Goldingay, H.; Bridgwater, A. V. Techno-economic and uncertainty analysis of Biomass to Liquid (BTL) systems for transport fuel production. *Renew. Sustain. Energy Rev. 2018*, 88, 160–175, doi:10.1016/j.rser.2018.02.023.

36. Feldmann, H.F.; Paisley, M.A.; Appelbaum, H.R.; Taylor, D.R. *Conversion of Forest Residues To a Methane-Rich Gas in a High-Throughput Gasifier*; Battelle: Columbus, OH, 1988;

37. Bain, R.L. *Material and Energy Balances for Methanol from Biomass Using Biomass Gasifiers*; NREL/TP-510-17098. National Renewable Energy Laboratory: Golden, CO, 1992;

38. Craig, K.R.; Mann, M.K. *Cost and Performance Analysis of Biomass-Based Integrated Gasification Combined-Cycle (BIGCC) Power Systems*; NREL/TP-430-21657. National Renewable Energy Laboratory: Golden, CO, 1996;

39. Spath, P.; Aden, A.; Eggeman, T.; Ringer, M.; Wallace, B.; Jechura, J. *Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier*; NREL/TP-510-37408. National Renewable Energy Laboratory: Golden, CO, 2005;

40. Dry, M.E. The Fischer–Tropsch process: 1950–2000. *Catal. Today 2002*, 71, 227–241, doi:10.1016/S0920-5861(01)00453-9.

41. Petersen, A.M.; Farzad, S.; Görgens, J.F. Techno-economic assessment of integrating methanol or
Fischer-Tropsch synthesis in a South African sugar mill. *Bioresour. Technol.* **2015**, *183*, 141–152, doi:10.1016/j.biortech.2015.02.007.

42. Espinoza, R.L.; Steynberg, A.P.; Jager, B.; Vosloo, A.C. Low temperature Fischer-Tropsch synthesis from a Sasol perspective. *Appl. Catal. A Gen.* **1999**, *186*, 13–26, doi:10.1016/S0926-860X(99)00161-1.

43. Song, H.; Ramkrishna, D.; Trinh, S.; Wright, H. Operating strategies for Fischer-Tropsch reactors: A model-directed study. *Korean J. Chem. Eng.* **2004**, *21*, 308–317, doi:10.1007/BF02705414.

44. Agência nacional do petróleo gás natural e biocombustíveis (ANP) Resolution ANP n° 40, from October 25, 2013, about the specifications of automotive gasoline. Available online: http://legislacao.anp.gov.br/?path=legislacao-anp/resolanp/2013/outubro&item=ranp-40--2013 (accessed on Dec 6, 2018).

45. Agência nacional do petróleo gás natural e biocombustíveis (ANP). Resolution ANP n° 50, from December 23, 2013, about diesel specifications. Available online: http://legislacao.anp.gov.br/?path=legislacao-anp/resol-anp/2013/dezembro&item=ranp-50--2013 (accessed on Dec 6, 2018).

46. ASTM Standard D1655-20e1. Standard specification for aviation turbine fuels 2020. Available online: https://www.astm.org/Standards/D1655.htm (accessed on Dec 7, 2020).

47. Centro de Estudos Avançados em Economia Aplicada (CEPEA). Indicador mensal etanol hidratado CEPEA/ESALQ Combustível – Estado de São Paulo. Available online: http://cepea.esalq.usp.br/etanol/ (accessed on Dec 9, 2018).

48. Agência nacional do petróleo gás natural e biocombustíveis (ANP). Preços de produtores e importadores de derivados de petróleo. Available online: http://www.anp.gov.br/precos-e-defesa-da-concorrencia/precos/precos-de-produtores (accessed on Dec 6, 2018).

49. Câmara de Comercialização de Energia Elétrica (CCEE). Historical prices of electricity in the regulated market (auctions). Available online: https://www.ccee.org.br/portal/faces/pages_publico/o-que-fazemos/como_ccee_atua/precos/historico_preco_semanal?_afrLoop=11034628523276496&_adf.ctrl-state=h9vdymyi82#%40%40%3F_afrLoop%3D11034628523276496%26_adf.ctrl-state%3Dh9vdymyi86 (accessed on Apr 12, 2018).

50. Ministério de Minas e Energia (MME). RenovaBio. Consulta Pública - Anexo da Nota Técnica n° 12/2018/DBIO/SPG. Proposta de Metas Compulsórias Anuais de Redução de Emissões na Comercialização de Combustíveis. Available online: http://www.mme.gov.br/c/document_library/get_file?uuid=40d335ad0-582d-82e3-1de0-61979c5905ae&groupId=36224 (accessed on Jul 13, 2020).

51. Programa de Disseminação das Estatísticas do Trabalho (PDET). Base de dados online. Available online: http://pdet.mte.gov.br/acesso-online-as-bases-de-dados (accessed on Oct 22, 2019).

52. Independent Commodity Intelligence Services (ICIS). Chemical prices. Available online: https://www.icis.com/explore/commodities/chemicals/ (accessed on Oct 15, 2018).

53. Química e Derivados. Mercado, São Paulo. Available online: https://www.quimica.com.br/category/revista/ (accessed on Oct 15, 2018).

54. Ministério da Economia (ME). Estatísticas de Comércio Exterior. Comex Stat, Exportação e Importação Geral. Available online: http://comexstat.mdic.gov.br/pt/geral (accessed on Jul 13, 2020).

55. Instituto de Pesquisas e Educação Continuada em Economia e Gestão de Empresas (PECEGE). *Custos de produção de cana-de-açúcar, açúcar, etanol e eletricidade na região Centro-Sul do Brasil: acompanhamento da safra 2013/2014*; Piracicaba, SP, Brazil, 2014;
56. Cardoso, T.F.; Watanabe, M.D.B.; Souza, A.; Chagas, M.F.; Cavalett, O.; Morais, E.R.; Nogueira, L.A.H.; Leal, M.R.L.V.; Braunbeck, O.A.; Cortez, L.A.B.; et al. Economic, environmental, and social impacts of different sugarcane production systems. *Biofuels, Bioprod. Biorefining* **2018**, *12*, 68–82, doi:10.1002/bbb.1829.

57. Milanez, A.Y.; Nyko, D.; Valente, M.S.; Souza, L.C.; Bonomi, A.; Jesus, C.D.F.; Watanabe, M.D.B.; Chagas, M.F.; Rezende, M.C.A.F.; Cavalett, O.; et al. De promessa a realidade: como o etanol celulósico pode revolucionar a indústria da cana-de-açúcar - uma avaliação do potencial competitivo e sugestões de política pública. *Biocombustíveis BNDES setorial* **2015**, *41*, 237–294.

58. Silva, J.P. de L.; de Aquino, G.S.; Navroski, D.; de Souza, Italo F.; Soriano, H.L.; Bressiani, J.A.; Ralisch, R. Biometrics, productivity and technological quality of 23 energy sugarcane hybrid clones with higher lignocellulosic biomass. *Aust. J. Crop Sci.* **2018**, *12*, 915–921, doi:10.21475/ajcs.18.12.06.PNE936.

59. Furtado Júnior, J.C.; Palacio, J.C.E.; Leme, R.C.; Lora, E.E.S.; da Costa, J.E.L.; Reyes, A.M.M.; del Olmo, O.A. Biorefineries productive alternatives optimization in the brazilian sugar and alcohol industry. *Appl. Energy* **2020**, *259*, 113092, doi:10.1016/j.apenergy.2019.04.088.

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