Comparative techno-economic and exergetic analysis of gasification technologies

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ABSTRACT

Nowadays, the use of lignocellulosic materials to produce biofuels/electricity is a field of ever growing importance. Two options of the interesting residual biomass are sugarcane bagasse and straw. A technological pathway for the conversion of biomass is the thermochemical route. Hence, biomass gasification represents a promising technology within this route for power generation and large scale production of a variety of biofuels (*e.g.* Methanol, Dimethyl ether, Fischer-Tropsch fuels, Hydrogen, and Synthetic Natural Gas). In this work, the techno-economic and exergy analysis of the thermochemical routes integrated into a Combined Cycle are presented. The lignocellulosic gasification process uses fluidized-bed (circulating and dual bed) systems, with subsequent conversion of the synthesis gas (syngas). The process simulations were carried out using Aspen Plus® software, including biomass drying and gasification, syngas cleaning, conditioning, separation of synthesis products, and heat recovery. From the techno-economic assessment, the DFB-LCM model has the lowest total cost of investment, which represents a 40% less concerning the CFB-SCB system. In addition, the DFB-LCM model shows the lowest exergy destruction rate; as a consequence, the higher the exergy efficiency of the overall systems. Hence, the DFB-LCM model could represent a competitive application for the agro-industrial sector to explore the contribution/potential of lignocellulosic materials on carbon-neutral systems by converting syngas into electricity.

KEYWORDS

Lignocellulosic gasification, Techno-economic analysis, Exergy analysis, Process integration.

INTRODUCTION

Based on the large availability of processed lignocellulosic biomass in the sugarcane industry, as well as in other agro-industrial activities, it represents an effective feedstock to be harnessed. In this sense and looking for new perspectives for bioenergy systems, there is a growing interest in the simultaneous use of lignocellulosic biomass for liquid biofuel production and electricity generation. Currently, thermochemical (gasification) process is one of the most studied to break the complex lignocellulosic biomass compounds in utilizable molecules to produce liquid fuels and power at low and high temperatures [1]. However, although the lignocellulosic biomass is accessible at a reduced cost, these technological options for their conversion are still under development.

Thermodynamic and economic evaluation of biomass gasification

Recently, several comprehensive reviews have been published on biomass gasification involving the description of processes, trends, and technological issues. Asadullah [2], reports on the barriers in each of the steps from the collection of biomass for electricity generation. Furthermore, the author discusses the effects of operational parameters in supply chain management, pretreatment, conversion, cleaning, and utilization steps for power generation using syngas. Damartzis and Zabaniotou [3], present the thermochemical conversion of biomass to second-generation biofuels as well as indicate the emerging challenges and opportunities of process integration. Moreover, Gómez-Barea and Leckner [4], review the modeling of biomass gasification in bubbling and circulating fluidized-bed gasifiers. Besides, Safarian et al. [5] collect and analyze statistics on the growing number of gasification modelling studies and approaches used.

An extensive literature review of model modifications has been carried out to provide a better understanding of gasification modeling for future research. For instance, Silva et al. [6] present an updated review of the stoichiometric thermodynamic equilibrium model for biomass gasification applications. Also, Mehrpooya et al. [7] investigate the biomass gasification process through the modeling and simulation of 23 different kinds of biomass sources. The process operating performance was analyzed thermodynamically based on the Gibbs free energy minimization and the restricted equilibrium method. Rupesh et al. [8] analyses the performance of several biomasses during gasification through energy and exergy analysis. Thus, a quasi-equilibrium model was developed to simulate and compare the feasibility of different biomass materials as gasifier feedstock. Hence, it is noted that the multi-reaction equilibrium approach is a common method for gasifier modelling of biomass and coal sources as reported in the literature [9,10]. For the particular case of sugarcane bagasse gasification, recent analysis has shown the global reaction mechanism of syngas evolution, considering a semi-batch reactor with a fixed amount of sugarcane bagasse sample placed in a steady flow of high-temperature steam at atmospheric pressure [11].

On the other hand, studies with a focus on the economic assessment of biomass gasification have been recently published [12,13]. For instance, Rahimi et al. [14] develop a comprehensive software program to simulate biomass gasification, which utilized an experimental setup to calibrate the simulation results with appropriate modeling coefficients. It was concluded that multiplying equilibrium constants by 0.7 yields offers a better agreement between the simulation results and experimental values. Also, the sensitivity analysis shows that increasing the biomass moisture content will decrease CO relative composition and will increase H_2 and CH_4 relative compositions in the producer gas. Thus, the authors proposed a system that could save roughly 4 million Nm^3 of natural gas per year, and the period of return of the project investment report was 6 years.

Shahabuddin et al. [15] summaries the recent techno-economic analyses for advances configurations of the thermochemical production of hydrogen from biomass and residual wastes. This review finds that the thermal efficiency of hydrogen from gasification is near to 50 %. Also, the authors found that the levelized cost of hydrogen (LCOH) from biomass varies between 2.3 and 5.2 USD/kg at feedstock processing scales of 10 MW_{th} to 2.8–3.4 USD/kg at scales above 250 MW_{th}. Hence, preliminary estimates of the LCOH from residual wastes are in the range of 1.4–4.8 USD/kg, depending upon the waste gate fee and project scale.

Hannula [16] explore the potential to increase biofuels output from a gasification-based biorefinery using external hydrogen supply. The author found that the biofuel output could be increased by a factor of 2.6 to 3.1 for gasoline or methane production over reference plant configurations, respectively. The economic assessment shows that the average cost of low-carbon hydrogen below 2.2-2.8 €/kg becomes economically attractive over non-enhanced designs, depending on the process configuration. This study analyzed the use of all available wastes and residues in the European Union-EU (197 Mt/year, 2016) and its conversion to biofuels. Thus, the author claims that the total supply of hydrogen enhanced biofuels could displace up to 41-63 % of the EU's road transport fuel demand in 2030, depending on the process design selected.

Lastly, AlNouss et al. [17] propose a poly-generation process that utilizes multiple sources of biomass feedstock to produce high-quality urea, methanol, Fischer-Tropsch liquids and power to perform the economic, energy and environmental analyses. The results demonstrate that the methanol production is the most attractive process pathway with a net profit of approximately 0.03 USD/kg of biomass input; when considering the production capacity, liquid fuels production achieves net profits of approximately 0.27 USD/kg of product.

In this context, it must be underlined that biomass gasification is significantly flexible in terms of the feedstock/waste uses. These materials can be processed to either produce biofuels or cogenerate electricity and heat on demand. Thus, the gasification flexibility related to the feedstock and also the energy generation or fuel production options drive research and development opportunities for biomass gasification. For this work, a biomass gasification model was developed that can be used to assess its feasibility for power production. Hence, a performance comparison in terms of the economic indicators (*e.g.* CAPEX, OPEX, and NPC), exergy efficiency, and destroyed exergy rate of the thermochemical route integrating fluidized-bed gasification and the combined cycle is determined. Thus, a technical performance comparison of thermochemical systems using the exergy concept as an indicator was carried out, which could be relevant and support the decision-making process regarding further research on low-carbon technologies.

MATERIALS AND METHODS

The first step of the conversion process was the selection of a suitable feedstock capable of generating gaseous fuels (feedstock section). Next, the conversion technologies item described the fluidized bed gasifier which achieves promising economy-of-scale for fuel production and power co-generation. Later, the techno-economic analysis approach and exergy assessment are presented.

Feedstock

The modelled thermochemical systems include agroforestry residues feedstocks (Table 1) produced in Brazil. Thus, the simulation of the biomass gasification process is investigated and analysed considering different kinds of biomass sources.

Table 1. Feedstocks composition

Model	CFB-SCB	DFB-SCB	DFB-LCM
Type of gasifier	*Circulating*	*Fluidized bed*	*Fluidized bed*
Biomass Type	*Sugarcane bagasse*	*Sugarcane bagasse*	*Lignocellulosic material*
Ref.	[18]	[19]	[20]
PROXANAL			
Moisture	7.8	50	15
FC	10.81	18	17.17
VM	83.97	79.06	80.1
Ash	5.22	2.94	2.73
ULTANAL			
Carbon	44.52	46.96	46.92
Hydrogen	5.9	5.72	5.73
Nitrogen	0.32	0.27	0.33
Chlorine	0.29	0.02	0.04
Sulfur	0.1	0.04	0.05
Oxygen	43.65	44.05	44.2
Ash	5.22	2.94	2.73

Conversion technologies

Promising thermochemical pathways were considered in the technological assessment. Fluidized-bed gasifiers are noted for their excellent mixing and temperature uniformity. A fluidized bed is made of granular solids called bed materials, which are kept in a semi-suspended condition (fluidized state) by the passage of the gasifying medium through solids. Excellent gas-solid mixing and large thermal inertia of fluidized bed make this type of gasifier relatively insensitive to the fuel's quality [21]. The temperature uniformity in the fluidized bed dramatically reduces the risk of fuel agglomeration. The fluidized-bed design has proved to be particularly advantageous for the gasification of biomass. Its tar production lies between that for updraft (~50 g/Nm^3). In this work, two fluidized-bed types were evaluated, circulating and dual bed systems.

Fluidized bed gasifier (DFB-LCM model)

Fig. 1 shows the integrated gasification and combined cycle system. The gasifier was modelled with three reactor models (*e.g.*, RCT-1, RCT-2, and RCT-3). The first one is an *RYield block* used to decompose the biomass (lignocellulosic materials) into its constituent elements, converting the BIOMASS stream, which consists of the non-conventional component (defined by the ultimate and proximate analyses), into a stream of conventional ($C_{volatile}$, H_2, N_2, Cl_2, S), solid (C_{solid}) and non-conventional components (ash), as defined in Tab. 1.

A fraction of the solid carbon (char) is addressed to the combustion process in the *RStoic reactor* RCT-3 to provide the heat duty necessary to maintain the temperature (880 °C) in the gasifier. Thus, the combustion zone is represented in the dual gasifier. The inlet stream of air is composed of 21 % O_2 and 79 % N_2, molar basis, and its flow rate is calculated with 15 % excess concerning the stoichiometric flow necessary for total combustion of the inlet char.

A RCT-2 reactor (*RGibbs block*) is used to modelled the gasification zone, by minimizing the Gibbs free energy to obtain the product's composition under conditions of thermodynamic equilibrium. A stream of superheated steam at 480 °C is fed into this reactor. Its mass flow rate is computed using steam to biomass (STB) ratio of 0.75. This term represents the mass flow rate of biomass moisture plus the injected steam, divided by the dry biomass. Besides, the cleaning process was represented as a cyclone (*SSplit block*) that separates the ash, and a flash drum operating at 25 °C to remove most of the water.

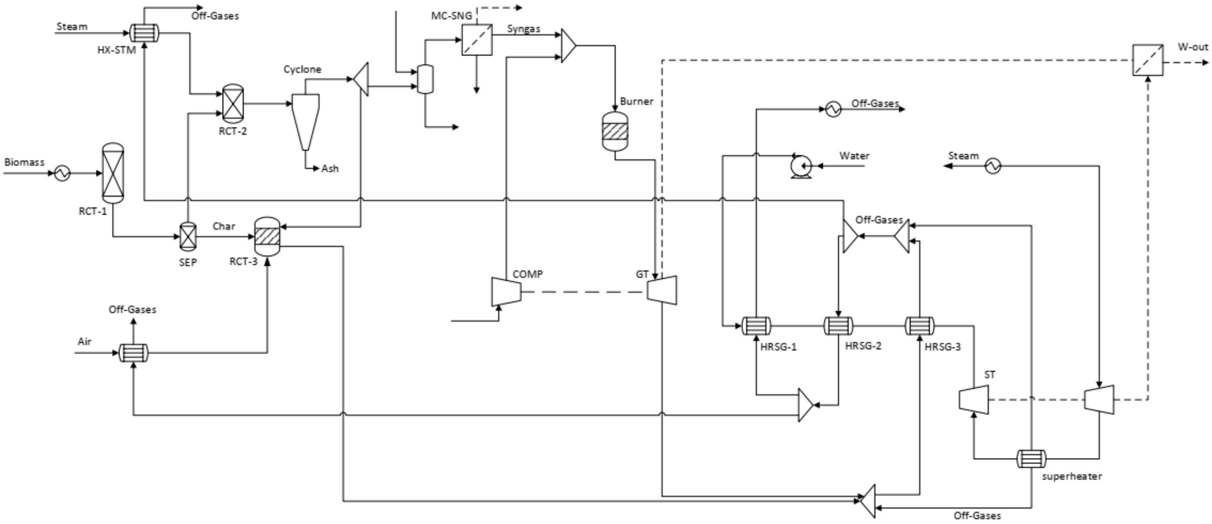

Figure 1. Fluidized bed gasification of lignocellulosic biomass

<u>*Circular fluidized bed (CFB-SCB model)*</u>

Fig. 2 shows the atmospheric circulating fluidized bed system integrated with the combined cycle. The first step is the biomass decomposition. In this step, the biomass is converted into its constituting components such as *carbon, hydrogen, oxygen, nitrogen, sulphur, chlorine, and ash* through a yield reactor. Later, the obtained stream (Ryield block, decomposition) goes through a separation column unit (Pyrolysis); at this step, the pyrolysis process occurs. A separation column model was used before the Gibbs reactor (Dryer) to separate the volatile materials and solids to perform the volatile reactions.

In the following step, char gasification, the gasifying medium (AIR), is provided to guarantee optimal gasification operational conditions. The char particles resulting from the devolatilization process consist of the reaming carbon fraction and ash. Furthermore, the cyclone model (Cyclone unit) represents the gas/solid separation step at the riser outlet. The top outlet stream (*Syngas*) is composed of all the gases from the Gibbs reactor. Lastly, the bottom outlet solid stream (*AS*H) represents the output from the ashes.

The following assumptions were adopted during the development of the gasification process: *i).* Steady-state and isothermal conditions; *ii).* Since the time required for volatile combustion is very short, the devolatilization process is considered to be instantaneous and to take place at the bottom of the bed; *iii).* Char and volatiles are formed in the pyrolysis process. The volatiles includes non-condensable (*e.g.*, H_2, CO, CO_2, CH_4, C_2H_2), condensable volatiles (tar), and water [4]; *iv).* Char only contains carbon and ash [22].

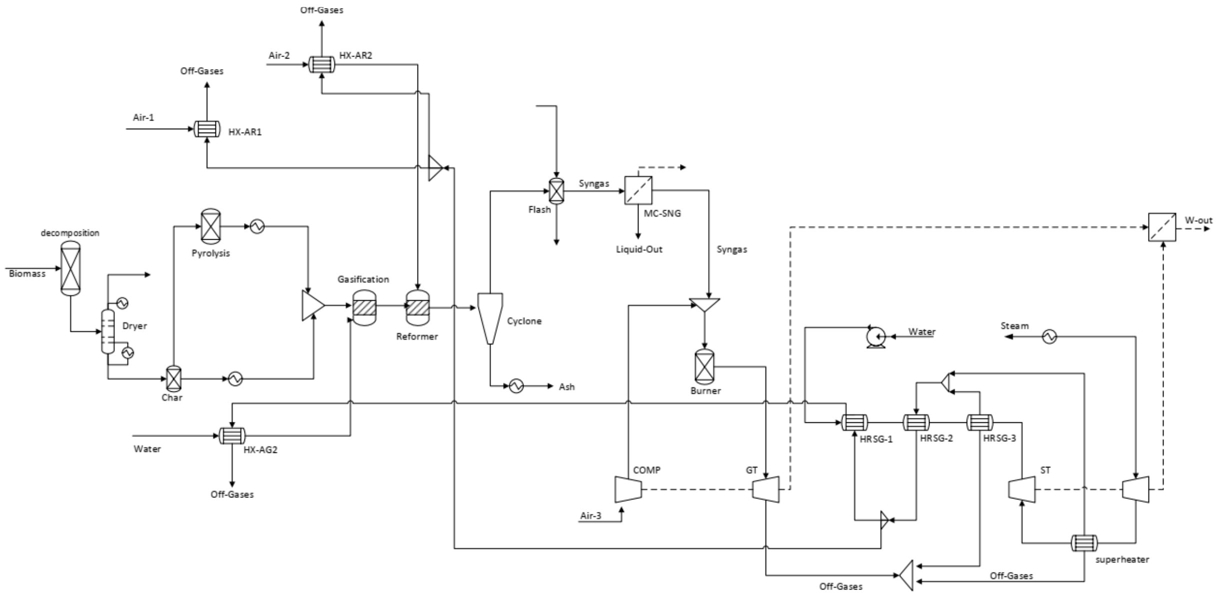

Figure 2. Circular fluidized bed model

Dual fluidized bed (DFB-SCB model)

The dual fluidized bed was modelled as indicated in Fig. 3. The model comprises biomass feed handling and drying, indirectly-heated gasification in dual circulating fluidized bed; and cyclone removal of particulates. This type of gasifier consists of two separate, interconnected, beds, through which hot bed material circulates and transfers heat between different zones. In the gasification zone (*gasification reactor, RGibbs model*) bed, steam is added as a sole gasifying agent. In contrast, in the combustion zone (*combustion reactor*, *RStoic model*) bed, the air is added as a combustion agent. Thus, high-quality syngas with a higher concentration of H_2 can be produced.

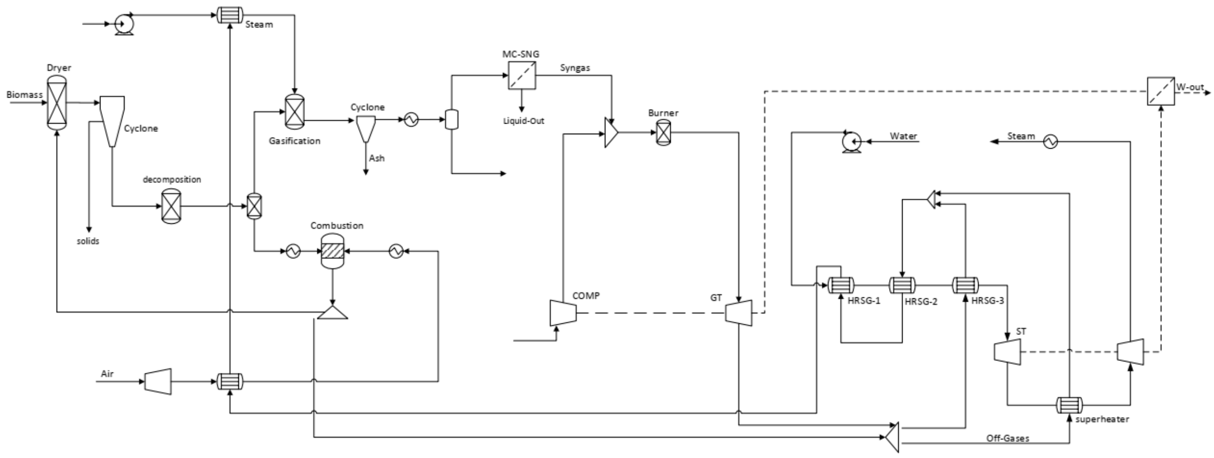

Figure 3. Dual fluidized bed model

In this model, wet sugarcane bagasse (BIOMASS stream) is used, which is then reduced to 10 wt °% in a rotary dryer using hot flue gases from the combustion of a fraction of the char that is formed during the pyrolysis reactions that take place in the gasifier. The dryer (*RYield block*) simulates drying of moisture with hot gases, while the *cyclone* separates gas and solid phases after drying. Since non-conventional components do not participate in phase and chemical equilibrium calculations, a decomposition block (*RYield reactor*) is firstly used to decompose biomass into its constituent elements (C, H_2, O_2, N_2, Cl, S, H_2O) to enable the subsequent calculations.

On the other hand, pre-heated air at 130°C is supplied into the combustion zone, where char is burned to provide heat for the endothermic gasification reactions. Then, the separator represents the separation of the char fraction that is used for combustion. The gasification zone is fed with biomass and saturated steam, which acts as gasifying agent at 2.5 bar (127.5 °C). The STB ratio (typical range of 0.2-2) adopted was 0.34, with a preference for lower values due to lower energy consumption [23]. The temperatures in the Gasification and the Combustion zones are assumed to be 950°C and 1000 °C, respectively, which is consistent with operating ranges reported for this type of gasifier [9,10]. Afterward, the gas passes through a cyclone to remove particulates. It is emphasized that the gasifier model adopts a multi-reaction equilibrium approach.

Combined cycle

The combined cycle comprises a gas turbine that simulates a Brayton Cycle and a steam turbine following a conventional Rankine Cycle. The simulation was carried out based on the following assumptions: *i). Steady-state operation; ii). isentropic compression process, iii). No heat losses in the combustion chamber.*

The inlet stream in the combined cycle is the cleaned syngas obtained in the gasification process. The mass flow rate of air is set to obtain a total outlet flow rate and compressed according to the specified pressure ratio of 20 bar. After compression, the air is led into the combustion chamber of the gas turbine (*RStoic block*). By combusting the syngas the flue gas together with the excess air stream is heated up to 1350 °C before it is entering the gas turbine. The GT represents the gas turbine and operates with an outlet pressure of 1.1 bar and an isentropic efficiency of 88 %. The outlet gases from the GT enter then the HRSG section in which it exchanges heat with a steam cycle. The steam cycle consists of one high-pressure steam turbine and one medium- pressure to a low steam turbine (ST). In the first steam turbine, the pressure is reduced from 200 to 50 bar.

After an intermediate superheating step (to 550 °C), the steam is led into the second steam turbine and expanding until a pressure of 50 bar and a condensate fraction of 5 %. A heat exchanger is used to condense the water that will be pumped back and begin another cycle. More details of the technical performance of gas turbines using low heating value fuel can be found elsewhere [24,25].

Techno-economic evaluation

The TEPET (*Techno-Economic Process Evaluation Tool*) methodology was used for estimating Capital investment costs (CAPEX), Operational expenditures (OPEX), and Net production costs (NPC). The cost estimation is expected to have an accuracy of ±30% for well-known chemical processes [26], according to class three and four of the classification system of Association for the Advancement of Cost Engineering [27].

The CAPEX includes the Fixed capital investment (FCI) costs, which consist of equipment costs (EC) and further capital requirements in the construction phase. EC for all installed units is calculated in TEPET based mainly on [28]. A database consisting of cost functions for main chemical process equipment as well as for fuel synthesis equipment was implemented from the TEPET. On the other hand, the OPEX could be broken down in costs for raw materials and utilities and other indirect operational costs (*e.g.*, maintenance, labour, insurances, administration, and taxes). Since exact costs are difficult to predict, typical estimates are used based on historical data form the chemical process industry [28]. More details of the TEPET methodology can be found elsewhere [26].

Exergy Assessment

The value and usefulness of resources is related to their capacity to do useful work. The second law of thermodynamics accounts for the fact that, even under ideal conditions, heat cannot be entirely converted into work. Thus, the concept of exergy represents the maximum ability of a system to do work concerning a reference state [29]. The technological scenarios are based on the calculation of the steady-state mass, energy, and exergy balances, according to Equations (1-3), respectively for each one of the control volume.

$$\sum_{inlet} \dot{m}_i = \sum_{outlet} \dot{m}_e \tag{1}$$

$$\sum_{inlet} \dot{m}_i h_i + \dot{Q}_{CV} = \sum_{outlet} \dot{m}_e h_e + \dot{W}_{CV} \tag{2}$$

$$\sum_{inlet} \dot{m}_i b_i + \dot{Q}_{CV_i}\left(1 - \frac{To}{T}\right) = \sum_{outlet} \dot{m}_e b_e + \dot{W}_{CV_e} + \dot{I} \tag{3}$$

where $\sum_{inlet} \dot{m}_i b_i$ represents the exergy of the process inputs ($\dot{B}_{inputs}$), $\sum_{outlet} \dot{m}_e b_e$ the exergy of the process output ($\dot{B}_{products}$), and ($\dot{I}$) the Irreversibility (exergy losses). In this work, the chemical (b_{CH}) and physical (b_{PH}) exergies are measured due to the physic-chemical processes involved. Thus, b_{PH} was determinate according to equation (4).

$$B_{PH} = H - H_0 - T_0(S - S_0) \tag{4}$$

Where *H* (in kW) is the enthalpy flow rate at P, T, *S* (in kW/K) represents the entropy rate/flow rate at P, T, T_o (in K) is the Temperature at the reference state, H_o (in kW) denotes the Enthalpy flow rate at P_o,T_o and S_o (in kW/K) is the Entropy rate/flow rate at P_o,T_o.

In particular, natural resources including lignocellulosic resources other chemical-based raw materials, the b_{CH} term is the most significant contribution to its exergy value [30]. Conceptually, it quantifies the value of a chemical substance/compound, as measured against a selected reference environment [31]. Equation (5) defines the chemical exergy:

$$B_{CH} = \mathrm{n}_{mix}\left[\sum_i x_i b_i^{ch} + R_u T_0 \sum_i x_i ln \Upsilon_i x_i\right] \tag{5}$$

Where n_{mix} is the total amount of moles of all constituents in a mixture, and x_i is the mole fraction of component *i*. The influence of Υ was evaluated for each compound permitting to observe that it offers values close to one. Consequently, Υ was adopted equal to one, an ideal solution, in mixtures for b_{CH} calculations [32]. The b_i^{ch} term denoted the standard chemical exergy. The chemical exergies for various compounds are found in the [33] and [34].

Performance analysis of the systems

Exergy efficiency: It is determinate by the ratio between the exergy of the products and the exergy of the resources, as given in Eq. (6).

$$\eta_B = \frac{\sum \dot{\mathrm{B}}_{products}}{\sum \dot{\mathrm{B}}_{resources}} \tag{6}$$

Irreversibility rate: The irreversibility reported was obtain by applying the exergy balance expression introduced in Eq. (3).

RESULTS AND DISCUSSION

After the energy/exergy analyses implemented, the minimum energy required, and the processes integration of the systems were evaluated to focus on maximize the electricity generation through the combined cycle system. Figures 4 to 6 show the total production costs estimated for each configuration. These graphical representations indicated the cost contribution of the electricity generation in all scenarios by each process step. For instance, the annuity represented the highest participation in the production costs. This term was calculated using the relation between the annuity factor, the fixed capital investment (FCI), the total capital investment (TCI) and the interest rate, as indicated in Albrecht et al. [26]. In this study, an operating time of the plant of 20 years, an interest rate of 10 %, and an annual full load of 8000 hours were considered. Another aspect of being highlighted in the techno-economic analysis is related to the raw material logistic (availability, transportation, cost production, and storage). In this context, Brazil is characterized by seasonal availability and low-cost biomass. Hence, a representative market price for the season 2018-2019, it was fixed in 25 €/t of biomass [35].

Fig. 4 presents the results concerning the fluidized bed gasifier that used lignocellulosic materials as feedstock integrated into the combined cycle. Furthermore, the results related to the integration between the fluidized bed system operating with sugarcane bagasse and the combined cycle is given in Fig. 5. Lastly, Fig. 6 shows the outcomes related to the circulating fluidized-bed system coupled into the combined cycle.

In addition, Table 2 displays the economic analysis results for each system obtained through the TEPET evaluation. Thus, a break-down of the investment and operational costs are given in order to determine the global behaviour of the system in light of the techno-economic feasibility assessment. It must be underlined that the TCI term represents the CAPEX (Capital investment costs) of the systems. Besides, OPEX and NPC are given in this table.

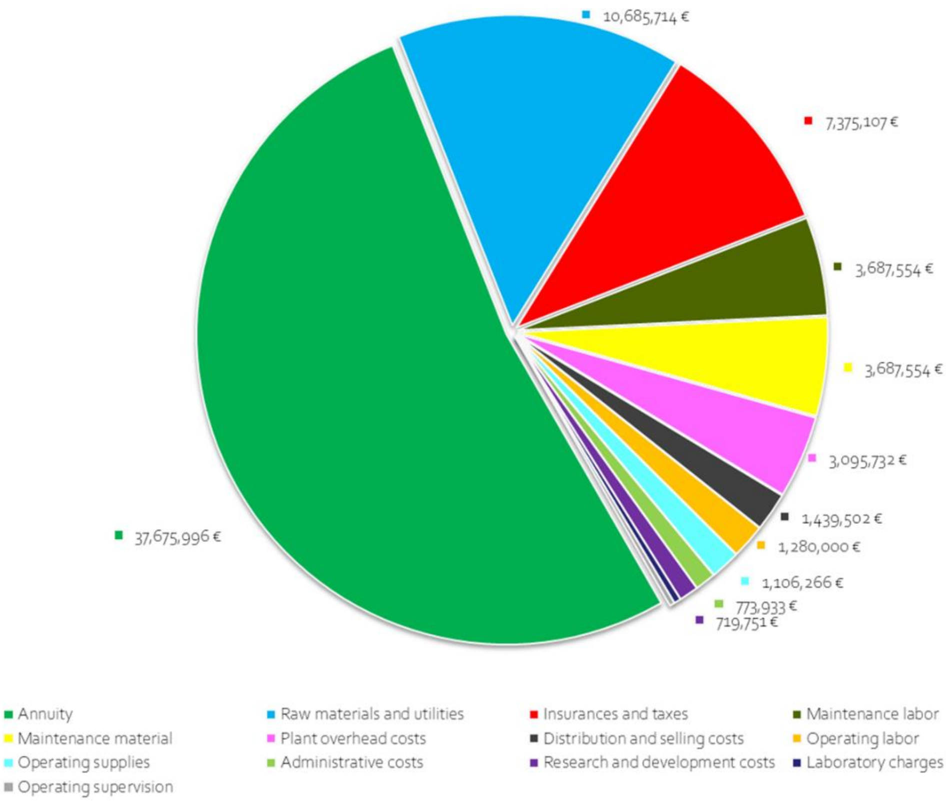

Figure 4. Total production costs of the DFB-LCM model

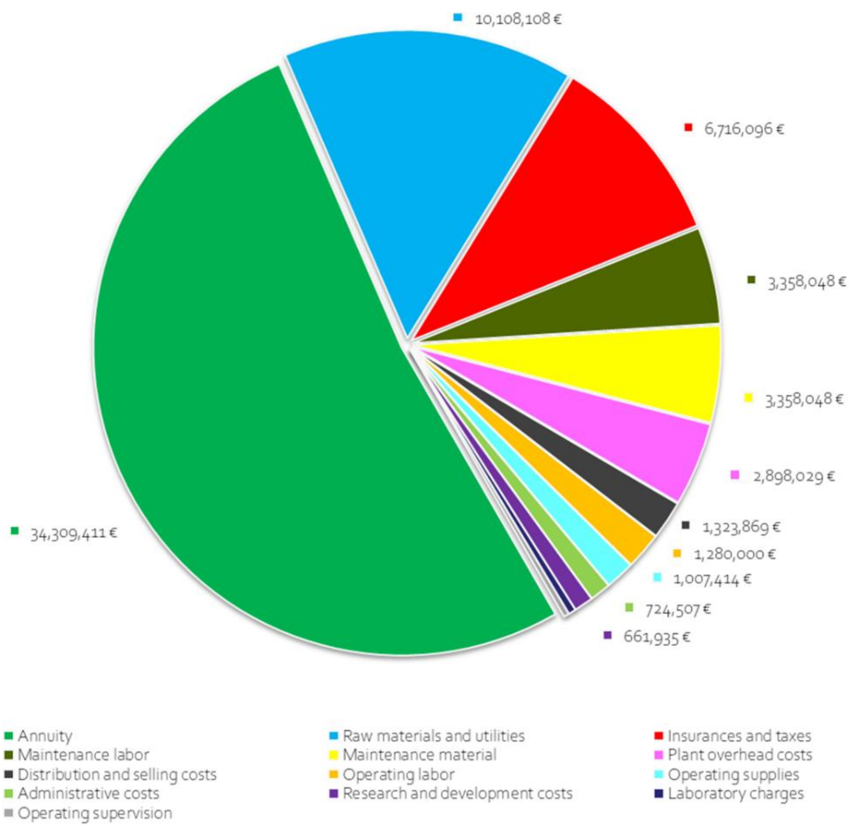

Figure 5. Total production costs of the DFB-SCB model

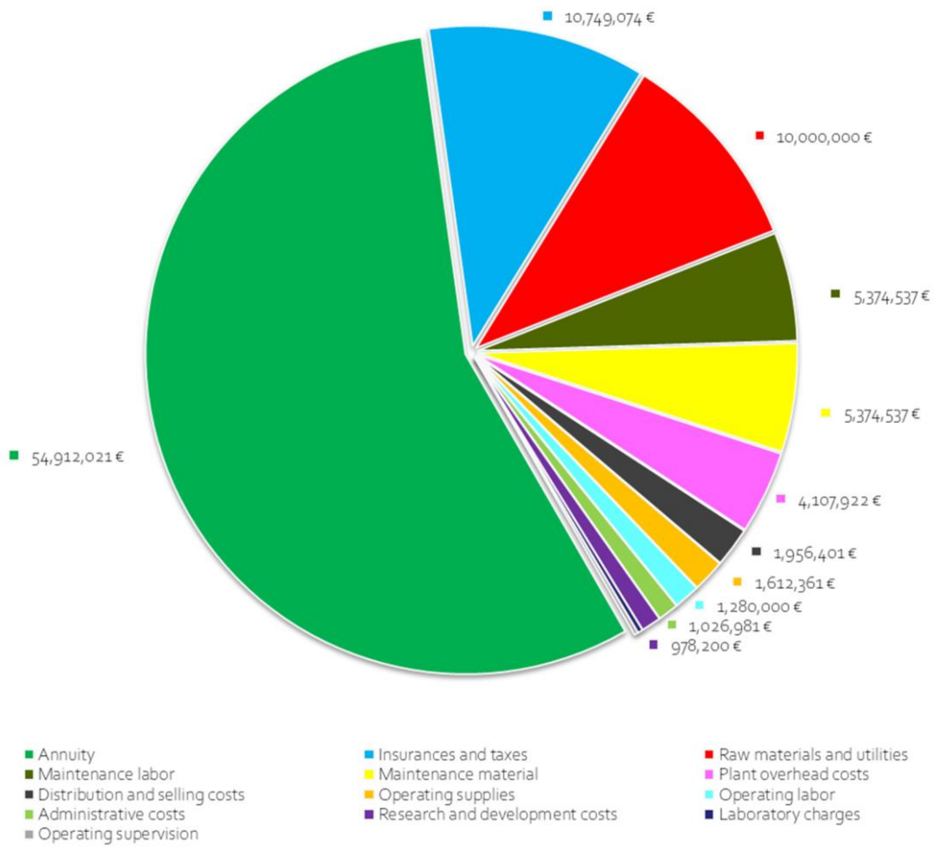

Figure 6. Total production costs of the CFB-SCB model

Table 2. Results of TEPET evaluation

Model		CFB-SCB	DFB-SCB	DFB-LCM
Type of gasifier		*Circulating*	*Fluidized bed*	*Fluidized bed*
Biomass Type	*Units*	*Sugarcane bagasse*	*Sugarcane bagasse*	*Lignocellulosic material*
Equipment costs	k€	116,190	77,386	84,583
Direct capital costs	k€	388,849	241,929	267,259
Indirect capital costs	k€	82,784	53,707	57,839
FCI (fixed capital investment)	k€	537,454	335,805	368,755
TCI (total capital investment)	k€	597,171	373,116	409,728
Annuity (annualized CAPEX)	k€/year	54,912	34,309	37,676
Expenses (feedstock & utilities)	k€/year	10,000	10,108	10,686
OPEX	k€/year	42,908	31,884	34,299
Product output	kWh/year	496,961	457,498	662,239
NPC (net production costs)	k€/year	97,820	66,193	71,975
Specific net production costs	*€/kWh*	*0.250*	*0.203*	*0.151*

According to Table 2, the CFB-SCB system represents the higher total cost of investment when compared to the fluidized bed systems, which is associated with the equipment costs and fixed capital investment. It is emphasized that the specific net production cost of the CFB-SCB system was the highest. In contrast, the DFB-LCM model presents the most attractive product output per year among the integrated gasification combined cycle.

In light of these results, the Lang factor was considered individually for each system to reflect the other costs contribution to plant cost. For all the systems the Lang Factor was calculated (4.6, 4.3, 4.4, respectively), which is a common value for mixed fluids-solids processing plants [28]. For instance, the Lang Factor for the production costs of syngas from lignocellulosic biomass in Brazil was determined as 3.8 through a biomass supply configuration focus on bulk chemicals [36].

Moreover, the payback period, which refers to the amount of time it takes to recover the cost of an investment, was determined for each system (18, 12, and 9 years, respectively). The results show that the DFB-LCM model offers a shorter payback period. Thus, this configuration could represent the more attractive investments under this indicator and also could represent a lower risk of the projects. In fact, the specific net production cost of the electricity via DFB-LCM model (0.15 €/kWh) could be competitive within the Brazilian market, when compare coal thermal plants (0.16 €/kWh) or nuclear systems (0.25 €/kWh) [37].

In order to synthesize the impact of the coupled systems (Gasification and Power generation) in terms of the performance indicators selected for the biomass residues feedstocks, the exergy efficiency and destroyed exergy rate are given for each model. Thus, Fig. 7 shows the overall system performance results of the technologies.

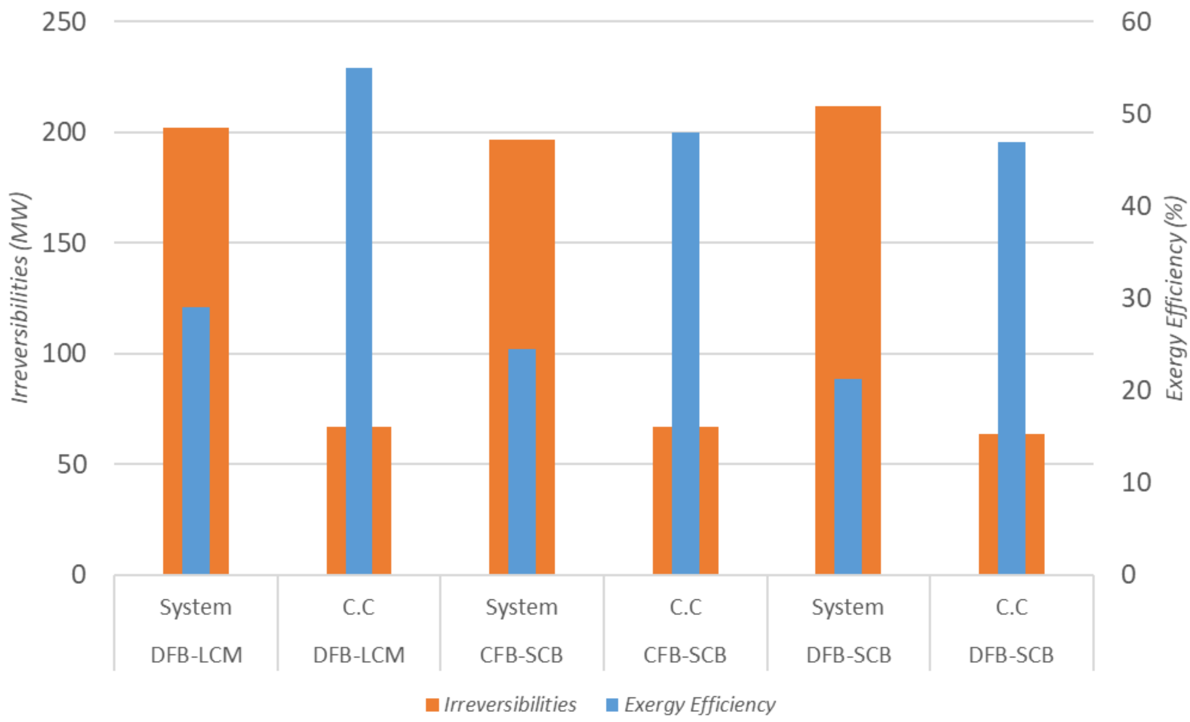

Figure 7. Systems Performance

Accordingly, the DFB-LCM model (fluidized-bed using lignocellulosic materials) shows higher exergetic efficiency as a consequence of the lower irreversibilities of the overall systems, when compared with the models integrated the gasification process and the combined cycle. However, DFB-LCM and DFB-SCB (dual fluidized-bed) systems present similar results when only is evaluated the combined cycle (Power generation section, indicated as *C.C* in Fig. 7). Hence, the DFB-LCM model represents the better-operating conditions to maximize the exergy efficiency of this system by maximizing heat recovery. In general, the key factors concerning the thermodynamic optimization of the systems are focus on setting the optimal performance conditions, such as, the excess air fraction that exists in the combustion, the air inlet temperature, the air-fuel ratio and preheating the combustion air.

CONCLUSION

The techno-economic assessment of the fluidized-bed gasification systems couple into a combined cycle show that the DFB-LCM model has the lowest total cost of investment, which represent a 40% less with respect to the CFB-SCB system. Hence, the DFB-LCM model presents a competitive advantage against the CFB-SCB system, since the former offers better product output per year. Another aspect of being highlighted is that the raw materials and utilities represented the highest participation in the production costs for all the evaluated systems.

From the exergetic analysis point of view, the DFB-LCM model shows the lowest exergy destruction rate; as a consequence, the higher exergy efficiency of the overall systems when contrast with the integrated gasification system and the power generation scenarios. Thus, the DFB-LCM configuration represents an application of value-added from biomass supply chain residues. This fact could contribute to determining the potential of lignocellulosic materials on carbon-neutral systems.

In this respect, it must be underlined that a sensitivity analysis focus on the valorization of the syngas and by-products could be covered in future research. For instance, the conceptual development of the biomass supply chains could be analyzed to explore other products and possibilities including feedstock production and collection, biomass gasification, syngas conditioning, and downstream processing. Hence, technological scenarios should be addressed to assess the syngas conversion into biofuel production, bulk-chemical/fuels or hybrid systems (chemical, fuels, and power generation).

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NOMENCLATURE

Capital investment costs (CAPEX)
Circular fluidized bed gasifier (CFB)
Dual fluidized bed gasifier (DFB)
Equipment costs (EC)
Fixed capital investment (FCI)
Lignocellulosic materials (LCM)
Net production costs (NPC)
Operational expenditures (OPEX)
Techno-Economic Process Evaluation Tool (TEPET)
Sugarcane bagasse (SCB)

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