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Bioenergy production from sugarcane bagasse with carbon capture and storage: Surrogate models for techno-economic decisions

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ABSTRACT

The use of biomass in cogeneration is a sustainable alternative of energy production, allowing replacing fossil fuels and reduction of greenhouse gas emissions. This work discloses an integrated process analyzer framework comprising surrogate models for estimation of fixed capital investment, revenues, costs of manufacturing as well as several performance responses of cogeneration units of sugarcane-biorefineries burning bagasse, with/without post-combustion carbon capture and storage. A restricted number of inputs are required, namely bagasse availability and heat requirements of the sugarcane-biorefinery. To develop the investment models, a 3^3 factorial computational-experimental design was performed, where AspenOne Portfolio was used in each run to simulate the process allowing estimating the fixed capital investment. Surrogate models were adjusted to fit capital estimates, resulting in 1.9% and 1.3% mean errors for the cogeneration and the post-combustion capture steps, respectively. Capture costs were estimated by analytical equations using the investment values and other estimates from the process analyzer framework, reaching 262 USD/t, but can be as low as 17.2 USD/t if limitations from the agricultural sector are disregarded; namely seasonality, operating time and capacity. The developed framework can assist in sugarcane-biorefinery investment decision making regarding bioenergy with carbon capture and storage or to develop carbon mitigation policies.

1. Introduction

Recently environment concerns have become a major issue on political agendas, directing efforts towards using renewable sources to supply the increasing energy demand. Forecasts indicates that by 2050 renewable fuels should displace petroleum as primary energy source [1], indicating the relevance of biomass-based Combined Heat and Power (CHP) systems. Biomass-based CHP, or cogeneration, produces steam and electricity with optimum efficiency [2] from the same source of bioenergy. Biomass-based CHP can fulfill two objectives: supply energy efficiently in many forms with economic benefits while mitigating greenhouse gas emission [3]. It also entails the benefit of energy security; i.e., the uninterrupted electricity availability during shortages of the main tributary of the energy matrix, e.g., hydroelectric plants in drought periods [4]. That is, biomass-based CHP becomes a sustainable and reliable alternative for energy production, allowing replacing fossil fuels.

Interest in biomass-based CHP has increased due to increasing fossil fuel costs and environmental concerns. Not only biomass, but a variety of renewable fuels can be used in CHP, including biogas, landfill-gas, solar energy, fuel-cells and waste-heat [5]. Natural gas can also be used in CHP, but it has environmental issues and gives inferior economic response than biogas in some configurations [5]. In the present study, bioethanol production from sugarcane takes advantage of the bagasse availability to feed CHP.

Brazil is one of the biggest bioethanol world producers, accounting for ≈26% of world production, second only to the USA which responds for ≈58% [6]. In the past, low-pressure boilers were used in Brazilian sugarcane-biorefineries because they were cheaper and sufficient for

Abbreviations: APEA, ASPEN Process Economics Analyzer; BECCS, Bioenergy with Carbon Capture and Storage; BRL, BR Real; BST, Biomass Steam-Turbine; CCS, Carbon Capture and Storage; CW, Cooling-Water; CHP, Combined Heat-and-Power; EOR, Enhanced Oil Recovery; GTCC, Gas-Turbine Combined-Cycle; LPS, Low-Pressure Steam; MEA, Monoethanolamine; MMUSD, Million US Dollar; MPS, Medium-Pressure Steam; NGCC, Natural Gas Combined-Cycle; SBAF, Sugarcane-Biorefinery Analyzer Framework; SHPS, Super High-Pressure Steam.

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Nomenclature

bb	: Boiler blowdown (%)
C_i	: Component *i* cost (MMUSD/t)
CC	: Capture cost (USD/tCO2)
COL, COM, CUT	: Labor, manufacturing and utility costs (MMUSD/a)
FCI	: Fixed capital investment (MMUSD)
dh	: Fraction of MPS and LPS for heating (%)
$\widehat{H}_i$	: Specific enthalpy of stream i (GJ/t)
$\Delta\widehat{H}_{S2-S1} = \widehat{H}_{S2} - \widehat{H}_{S1}$	: Difference of specific enthalpies S2 and S1 (GJ/t)
LF	: Dimensionless location factor
LHV	: Lower heating value (MJ/kg)
$\dot{m}$	: Flowrate (t/h)
MARR	: Minimal acceptable rate of return (%)
MSP	: Minimum steam selling-price (USD/t)
mu	: Make-up water (%)
NPV	: Net present value (MMUSD)
OT	: Operating time per annum (h/a)
P	: Pressure (bar)
T	: Temperature (°C)
W	: Power (MW)
X	: Bagasse flowrate (t/h)
Y	: Percentage of SHPS for MPS+LPS production (%)
Z	: MPS percentage in LPS plus MPS (%)

energy needs. However, the integrated sugarcane-biorefinery producing bioethanol, sugar and electricity surplus from bagasse burnt in high-pressure boilers was the most common configuration in the new 2007 Brazilian bioenergy projects [7]. This configuration became competitive with the 2004 reform of the Brazilian electricity sector that created conditions for commercialization of surplus electricity to the grid [8].

The two main configurations of biomass-fired CHP are: (i) biomass gasification integrated to Gas-Turbine-Combined-Cycle (GTCC); and (ii) Biomass-to-Steam-Turbine (BST). The former converts the biomass into fuel gas to the GTCC, while the latter, a mature technology, converts the biomass heating value into super high-pressure steam (SHPS) for steam-turbine expansion in Rankine Cycles. One can, for example, prescribe BST sugarcane-biorefineries producing first and second-generation bioethanol and still generating power surpluses when SHPS boilers are employed [9]. Dantas et al. [8] found that BST has greater viability after comparing three bagasse utilizations in sugarcane-biorefineries: (i) BST; (ii) biomass gasification with GTCC; and (iii) second-generation ethanol production.

The higher the boiler pressure, the higher the fixed capital investment (FCI, MMUSD) but the increase in electricity production is even higher. Therefore, high-pressure boilers are worthwhile [9], but it is not always possible to sell the electricity surplus if facilities are located far from demand sites [10]. This high FCI together with the feedstock supply assurance are the biggest barriers to new projects [11]. Moreover, the high number of uncertain variables seems to repel investors [12]. Papadimitriou et al. [13] analyzed eight Greek CHP operating plants showing that only half achieved the break-even point and five were oversized, i.e., operate far below design capacities.

Using fuzzy logic Ngan et al. [14] analyzed actions to mitigate risks in CHP palm-oil pyrolysis plants for bio-oil production in Malaysia, wherein risks were classified into regulatory, financial, technological, supply-chain, business and social-environmental. Financial incentives to reduce interest rate were found the most impactful actions for risk reduction, followed by revision of the feed-in tariff. The least important factor to risk management was the increase in bio-oil demand by replacing fossil fuel utilization. Moreover, supply-chain improvements were found more important than technology and process improvements.

Regarding carbon management of a CHP plant, the carbon dioxide (CO_2) content in dry flue-gas of bagasse-fired boilers is $\sim$ 12%mol [15], while in a natural gas combined-cycle power plants (NGCC) it is only ≈4.2%mol increasing to 6.6%mol if flue-gas recirculation is adopted [16]. The advantage of flue-gas recirculation is a higher %mol CO_2 in exhaust-gas, increasing by 0.5% the efficiency of Carbon Capture and Storage (CCS) [16]. Even so, the exergy efficiency of power generation decreases from 53.5% to 44.8% with CCS [17], indicating the economic impact of CCS in NGCC's. Thus, the high CO_2 content in bagasse-fired flue-gas is beneficial to CCS.

Massive implementation of bioenergy with CCS (BECCS) technologies is considered the best strategy to limit the increase in the average planetary temperature by a maximum of 1.5 °C by 2100, since BECCS technologies are the only carbon-negative processes available [18]. Gibon et al. [19] performed Life-Cycle Assessment of various electricity production technologies, using renewable sources or not, with/without BECCS/CCS. BECCS solutions achieved negative net emissions in all scenarios, but had a high increase in resource utilization as a trade-off. The BECCS potential of sugarcane-biorefineries at 1000 t/h of sugarcane corresponds to 659.6 tCO_2/h in the CHP and only 39.7 t/h in the bioethanol fermentation step [20], where the CO_2 emission rate is estimated as 700 $kgCO_2$/MWh for the bagasse-fired cogeneration, while for a NGCC it reaches 400 $kgCO_2$/MWh. However, the Life-Cycle Assessment estimated a Global Warming Potential of only 8.6–10 $kgCO_2$/MWh for district heat and 32–38 $kgCO_2$/MWh for CHP electricity production [21], vis-à-vis 460 $kgCO_2$/MWh when the CHP is a natural gas GTCC [22].

There are five major operational bioethanol BECCS plants in the world, located in the USA and Canada [23]. The biggest one stores geologically 1 Mtpa of CO_2 from corn-to-ethanol fermentation, while the other projects inject CO_2 for enhanced oil recovery (EOR) purposes. CO_2 from bioethanol fermentation is the only CO_2 source in all projects. A UK project in early development breaks this pattern: the Drax BECCS Project plans to capture 4.3 Mtpa of CO_2 from the flue-gas of a biomass-fired boiler, with geological storage as destination.

Fuss et al. [24] performed a literature review on BECCS costs for bioethanol production considering different CO_2 sources, finding that it is cheaper for the fermentation CO_2 (40–120 USD/tCO_2) than for both fermentation CO_2 and CHP CO_2 (180–200 USD/tCO_2). The reason is because the fermentation CO_2 is almost pure and trivially captured [25], though it is just released into the atmosphere in practically all current plants. When BECCS is performed only in the flue-gas from a biomass-fired power plant, capture costs of 88–288 USD/tCO_2 were reported, but various capture technologies and power plant configurations were considered in this range. Assuming geological storage as destination, Laude et al. [26] evaluated the BECCS potential for the fermentation CO_2 only and coupled to the CHP CO_2 finding CCS costs of 86 USD/tCO_2 and 143 USD/tCO_2 respectively, considering a 2050 carbon price of 200 EUR/tCO_2 for the base-case. The authors claim that the 60,000 m^3/a capacity of the bioethanol biorefinery is the reason behind this poor performance.

The work of de Souza et al. [27] evaluated the integration of CHP with a sugarcane-biorefinery, finding that the electricity cost is competitive compared to small systems. Carminati et al. [6] estimated the Net Present Value (NPV) of a sugarcane-biorefinery with/without BECCS. The economic performance of the BECCS biorefinery is even better than the non-BECCS counterpart, considering some stringent carbon-market scenarios reacting to a high climate-change severity and oil dependence. Guandalini et al. [28] studied the CCS of a biomass-based power plant disclosing a cost estimation model that can be used to foresee if a new technology performs better than the benchmark post-combustion capture via chemical-absorption with aqueous-monoethanolamine (aqueous-MEA). It also estimates the breakeven FCI and Cost of Manufacturing (COM) of a potential new technology attaining same capture cost of the benchmark aqueous-MEA absorption. Authors claim that capture technologies requiring mainly heat should be considered first when heat is available, as in the CHP case, due to a lower COM.

1.1. The present work

Following our previous studies [6,20,25] on onuses and bonuses related to BECCS implementation in sugarcane-biorefineries, this work develops and demonstrates quantitative tools to estimate FCI and COM changes, as well as the carbon capture cost, related to BECCS implementation in sugarcane-biorefineries. Since, for comparison purposes, the non-BECCS biorefinery has to be modeled beforehand, this work discloses surrogate models created by adjusting response surfaces over computer-experiments to estimate FCI and COM for installation of a new CHP plant coupled to sugarcane-biorefinery. The CHP uses BST technology with/without BECCS. The proposed generic-location model is applied to the South-East Brazilian scenario as one should realize that FCI and COM from other regions may substantially differ. The only required inputs are the energy requirements of the sugarcane-biorefinery – i.e., mass flowrates of low-pressure steam (LPS) and medium-pressure steam (MPS) – and bagasse availability. The sugarcane-biorefinery can be, for example, a sugar mill or a bioethanol fermentation-distillation plant. Considering typical steam consumption for sugarcane-biorefinery with distillery, the FCI curve against produced power is presented and compared to the literature and to power auctions in Brazil. The developed models are used to evaluate the Minimum Steam Selling-Price (MSP), BECCS costs and carbon capture cost under different scenarios of capacity and market prices. The scope of this work and the inputs/outputs of the developed models are highlighted in Fig. 1. It is adopted a distinct owner scenario, wherein bagasse is bought from a sugarcane-biorefinery and steam/electricity are sold to it. The price of exported electricity, sold to the sugarcane-biorefinery or to the grid as surplus, is obtained from national auctions. Thus, BECCS costs are allocated to the CHP plant, allowing comparisons with/without BECCS. Moreover, this is the first sugarcane-biorefinery work analyzing FCI against the capacity of the bagasse-fired CHP plant with and without BECCS. It is also reported a robust mass/energy balance model for estimating material/power streams of the bagasse-fired CHP, besides bleed-steam and net/total power given bagasse mass flowrate and energy requirements.

The main advantages of the FCI surrogate model are its accuracy, compactness and simplicity, besides the possibility to participate in plant optimizations, where the optimum capacity of each area is sought. The sugarcane-biorefinery feasibility for sugar/bioethanol productions [29] can be surrogated based on sugarcane intake and on the desired sugar/ethanol allocation. Conceivable novel technological extensions of the sugarcane-biorefinery operating in BECCS mode – such as production of ethylene/H_2 from bioethanol [30] or CO_2-to-methanol [17] to substitute the geological storage – can also be surrogated and included in the predictive framework developed here to optimize such extended sugarcane-biorefineries. Analogously, other complex bioenergy processes that have BECCS potential – e.g., anaerobic digestion, 2nd generation ethanol fermentation, syngas-to-ethanol anaerobic fermentation, biomass pyrolysis – can be efficiently modeled and optimized with such strategy of surrogate modeling.

2. Methods

The developed method consists of four steps: (i) process flowsheet development; (ii) flowsheet simulation in Aspentech portfolio; (iii) factorial 3^3 design of computational-experiments, leading to 27 flowsheet simulations for the CHP and three for the CCS configuration, allowing estimating FCI of flowsheets (base-year 2018) with Aspen Process Economic Analyzer (APEA) V11; and (iv) generation of surrogate models for FCI and COM based on the sugarcane-biorefinery energy requirements and bagasse availability. A surface response for FCI prediction is proposed, while COM is evaluated based on Turton et al. [31], for the estimated FCI. The FCI model is used in all conducted analyses, including sensitivity analyses and FCI curves of the CHP plant for typical Brazilian sugarcane-biorefineries.

2.1. Process flowsheet

The CHP-CCS flowsheet comprehends three areas: biomass-fired boiler (A10), power-house (A20) and CCS plant (A30, if present). The boiler burns bagasse from the sugarcane milling producing SHPS. The power-house exports heating utilities (LPS/MPS) and electricity from SHPS. The CCS plant captures and processes CO_2 abating biorefinery emissions.

In the boiler area in Fig. 2, it is considered the Bubbling Fluidized Bed type which offers an efficient, fuel-flexible and cost-effective burner for low-grade biomass with modest heating value and high humidity/ash contents [32]. The lignocellulosic bagasse stream (1) feeds the boiler furnaces (H-11). The boiler is also fed with demineralized-water (12), which is pre-heated with flue-gas, vaporized and superheated as SHPS in the radiation zone. The boiler exports SHPS (13) to the power-house (A20) steam-turbines; flue-gas (8) to the CCS plant (or to atmosphere) and ashes to the plantation fields (7). The boiler also receives primary

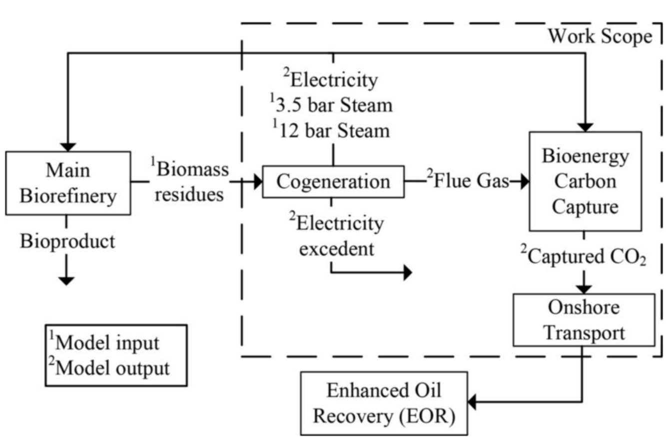

Fig. 1. Work scope diagram.

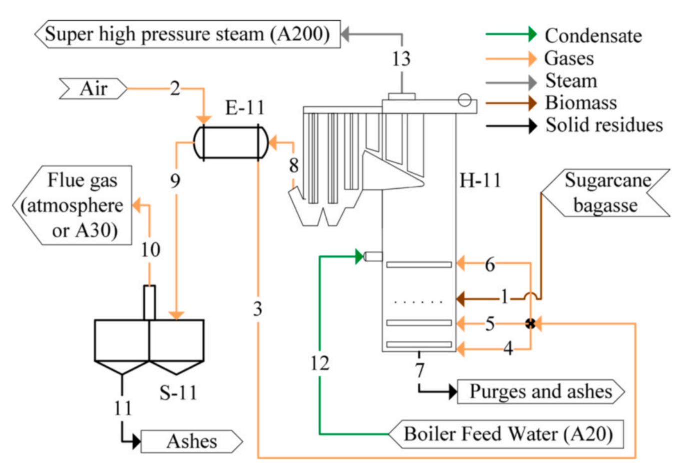

Fig. 2. Boiler flowsheet (A10).

(4), secondary (5) and tertiary (6) pre-heated air streams after pre-heating atmospheric air (2) via heat-recovery (E−11) from hot flue-gas (8). Cooled flue-gas is sent to the electrostatic precipitator (S-11), where particulates (soot) are removed as ashes (11), normally used as fertilizer. The final flue-gas with low particulate content (10) is liberated in the atmosphere or sent to the CCS unit (A30).

The considered power-house (A20) in Fig. 3 is the topping cycle, which consists in electricity generation prior to LPS/MPS generation to fulfill biorefinery demand [23]. SHPS (13) feeds a counter-pressure turbine (M − 21) with MPS extraction (14). A fraction of the expanded steam from the backpressure-turbine (17) becomes the LPS (18) for process heating purposes, while the remainder goes to the condensing turbine (M − 22). The low-pressure turbine outlet stream (20) is condensed (E−21) and pumped (P-21) to the pressurized deaerator (V-21), which also receives process condensates (23) and aerated water make-up (24). V-21 is injected with part of the MPS extracted from the counter-pressure turbine (16) – so-called bleed--stream – to adjust V-21 temperature providing the separation of purged air (25). The boiler feed water (26) is pumped by P-22.

2.2. Simulation of process flowsheets

The flowsheets of CHP and CCS plants are respectively installed in two process simulators, namely, ASPEN-Plus® v11 and ASPEN-HYSYS® v11. A10 and A20 are simulated using the ASPEN NRTL-RK thermodynamic package, whereas the CCS plant uses the HYSYS Acid-Gas Amine Package. Bagasse is modeled as cellulose during combustion, where 99.9% undergoes complete combustion Eqs. (1a) and (0.1)% follows incomplete combustion in Eq. (1b). Enthalpy data of cellulose is adjusted to match its lower heating value (*LHV*) of 7.5 MJ/kg (50%w/w moisture). Table 1 shows the general simulation assumptions and Table 2 lists Base-Case assumptions.

$$C_6H_{10}O_5(\text{bagasse}) + 6O_2 \rightarrow 6CO_2 + 5\,H_2O \tag{1a}$$

$$C_6H_{10}O_5(\text{bagasse}) + 3O_2 \rightarrow 6\,CO + 5\,H_2O \tag{1b}$$

2.3. Design of computational-experiments

A factorial 3^3 computational-experimental design is created with the following factors: bagasse capacity as X (t/h); LPS plus MPS productions as percentage of SHPS production Y(%); MPS as percentage of LPS plus MPS production Z(%). Eq. (2) defines the factors, where $\dot{m}_i$ is the mass flowrate of stream i (t/h).

$$X(t/h) = \dot{m}_1;\ \ Y(\%) = 100\left(\dot{m}_{15} + \dot{m}_{18}\right)\Big/\dot{m}_{13};\ Z(\%) = 100\dot{m}_{15}\Big/\left(\dot{m}_{15} + \dot{m}_{18}\right) \tag{2}$$

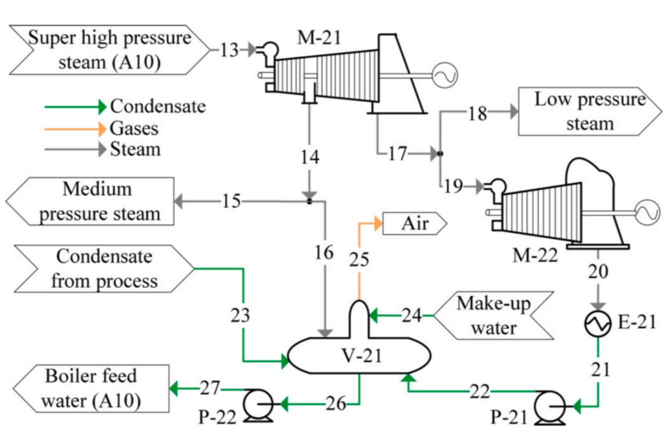

Fig. 3. Power-house flowsheet (A20).

Table 1
Process simulation assumptions.

Item	Description	Area	Assumptions
1	Bagasse	A10	Dry-bagasse = 47.8%w/w; Moisture = 50%w/w; Ash = 2.2%w/w; LHV = 7.5 MJ/kg.
2	Combustion Air	A10	Excess air: 50%.
E-11	Combustion Air Heater	A10	T_{out} = 120 °C; ΔP = 0 bar.
H-11	Biomass-Fired Boiler	A10	P = 90 bar; T_{SHPS} = 520 °C; Purged-Ashes = 90%; T_{Bed} = 1000 °C; P_{Bed} = 1 atm.
S-11	Electrostatic Precipitator	A10	Isobaric-Adiabatic Flash.
12	Boiler Feed-Water	A10	P_{inlet} = 109 bar; Steam:Bagasse Mass-Ratio = 2.093; Flowrate(kg/h) = Bagasse (kg/h)*2.093/0.99.
M-21	Extraction Backpressure-Turbine	A20	Type Isentropic; Adiabatic Efficiency: η = 85%; Mechanical Efficiency:η^{M}_{mech} = 96%; $P_{Extraction}$ = 12 bar; $P_{Discharge}$ = 3.5 bar;
M-22	Condensing-Turbine	A20	Type Isentropic; Adiabatic Efficiency η = 85%; Mechanical Efficiency: = 96%; $P_{Discharge}$ = 0.1 bar;
E-22	M-22 Condenser	A20	ΔP = 0 bar; Vapor-Fraction = 0.
P-21	Condenser Pump	A20	P_{outlet} = 3.55 bar.
23	Condensate	A20	P_{inlet} = 3.5 bar; Vapor-Fraction = 0.
24	Demineralized-Water	A20	Air = 0.1%w/w; Water = 99.9%w/w; P_{inlet} = 3.5 bar; T_{inlet} = 25 °C.
V-21	Deaerator V-21	A20	P = 3.5 bar; T = 100 °C; Recovery: Water = 98%, Air = 100%.
P-22	Boiler Pump	A20	P^{outlet} = 109 bar. Mechanical Efficiency η^{P}_{mech}=96%;
15 + 18	Total Steam	A20	50% Direct-Heating; 50% Shell-and-Tube Exchangers
All	CCS Plant	A30	Aqueous-MEA Chemical-Absorption [9]; Solvent: MEA = 30%w/w; Water = 70% w/w; CO_2 Capture Efficiency: 90%.

Table 2
Base-case assumptions.

Variable	Value	Reference
Sugarcane (t/h)	500	[33]
Bagasse:Sugarcane Ratio ($t^{Bagasse}/t^{Sugarcane}$)	0.276	[29]
Biorefinery Steam Requirements (kg/$t^{Sugarcane}$ sugarcane)	LPS = 429; MPS = 36.9	[29]
Biorefinery Power Requirement (kWh/$t^{Sugarcane}$ sugarcane)	28.0	[29]
Power Surplus (kWh/$t^{Sugarcane}$)	83.4	[29]

Only X affects flue-gas (10) flowrate, and consequently, the CCS plant (A30). Therefore, A30 has only three simulation runs instead of 27. The X levels are 33.75 (−1), 67.50 (0) and 135.0 (+1); while Y and Z levels are 25% (−1), 50% (0) and 75% (+1). Table 3 shows the planning of computational runs.

The response variables of each run comprise the FCI^{CHP} and FCI^{CCS} (2018 Brazil conditions). FCI is estimated via APEA, after equipment sizing via mass/energy balance results from simulations. In APEA, the boiler H-11 is evaluated as an oil-fired erected boiler. APEA is configured with 32% contingency for FCI^{CCS} [34], and 15% contingency for FCI^{CHP} since CHP is a mature technology. The resulting FCI is multiplied by the process plant location factor LF = 1.18 from Intratec [35] to convert 2018 USA FCI to 2018 Brazil conditions. FCI results from the computational-experimental simulations are used via minimization of the sum of residue squares for estimation of parameters a, b, c, d, $a^{\#}$, $b^{\#}$ in Eq. (3) and Eq. (4) for predicting FCI^{CHP} and FCI^{CCS}, respectively. In the right hand side of Eq. (3), the first term is a cost-to-capacity function to evaluate FCI at different capacities(X); the second term reflects the fact that larger net SHPS flowrate (i.e., discounting LPS+MPS) entails larger FCI of steam-turbines resulting in the $(1-Y/100)$

Table 3
Three-level, three-factor experimental planning (X,Y,Z).

Run 1 −1,-1,-1	**Run 2** −1,-1,0	**Run 3** −1,-1,1	**Run 4** −1,0,-1	**Run 5** −1,0,0	**Run 6** −1,0,1	**Run 7** −1,1,-1	**Run 8** −1,1,0	**Run 9** −1,1,1
Run 10 0,-1,-1	**Run 11** 0,-1,0	**Run 12** 0,-1,1	**Run 13** 0,0,-1	**Run 14** 0,0,0	**Run 15** 0,0,1	**Run 16** 0,1,-1	**Run 17** 0,1,0	**Run 18** 0,1,1
Run 19 1,-1,-1	**Run 20** 1,-1,0	**Run 21** 1,-1,1	**Run 22** 1,0,-1	**Run 23** 1,0,0	**Run 24** 1,0,1	**Run 25** 1,1,-1	**Run 26** 1,1,0	**Run 27** 1,1,1

exponent; and the last term translates the fact that high X and Y increase Z relevance. As these models were formulated following typical FCI behaviors, they have a prediction capacity outside the analyzed range of factors.

$$FCI^{CHP} = 1.15LF\left(aX^{b} + cX^{1-Y/100} - dX(Y/100)(Z/100)\right) \tag{3}$$

$$FCI^{CCS} = 150 + 1.32LF\left(a^{\#}X^{b^{\#}}\right) \tag{4}$$

COM^{CHP} and COM^{CCS} are estimated via Eq. (5) according to Turton et al. [31], where COL(MMUSD/a) is the annual cost of labor; CRM (MMUSD/a) and CUT (MMUSD/a) represent annual costs of raw-materials and utilities. COL is independent of capacity, as the number of operators typically depends on the number of equipment items and not on their capacities. $(CRM+CUT)^{CHP}$ and $(CRM+CUT)^{CCS}$ are respectively calculated by Eq. (6) and Eq. (10), where C_{CW} (USD/t), C_{DW} (USD/t), C_X (USD/t), and OT (h/a) respectively represent cooling-water (CW) cost, demineralized-water cost, bagasse cost and annual operation time. Moreover, $\left(\frac{\dot{m}_{CW}}{\dot{m}_{20}}\right)^{*}$, $\left(\frac{\dot{m}_{13}}{X}\right)^{*}$ and $\left(\frac{\dot{m}_{CO2}}{X}\right)^{*}$ are constant ratios obtained from the Base-Case simulation. Eq. (7) estimates the CO_2 mass flowrate in the flue-gas. Only FCI has parameters to be estimated since COM^{CHP} and COM^{CCS} are available via Eq. (5) once FCI^{CHP} and FCI^{CCS} are estimated.

$$COM = 0.18\mathrm{FCI} + 3.969COL + 1.235(CRM + CUT) \tag{5}$$

$$\dot{m}^{CCS}_{LPS} = X\left(\frac{\dot{m}^{CCS}_{LPS}}{X}\right)^{*} \tag{8}$$

$$W^{CCS} = X\left(\frac{W^{CCS}}{X}\right)^{*} \tag{9}$$

$$(CRM + CUT)^{CCS} = XC_{CW}\left(\frac{\dot{m}^{CCS}_{CW}}{X}\right)^{*} OT*10^{-6} \tag{10}$$

LPS, MPS and electricity are CHP products leading to revenues. Moreover, the captured CO_2 is monetized considering EOR activity and cap-and-trade regulations (similar to the EU Emissions Trading System) [36]. The BECCS biorefinery considers CO_2 sold to oil operators as EOR fluid (e.g., as in the offshore Brazilian Pre-Salt basin). Indeed, in the Pre-Salt oil recovery can be as low as 24% (e.g., offshore Campos basin [37]), creating a market for the captured CO_2. The price of CO_2-to-EOR derives from oil price, since up to 4.35 bbl of oil can be recovered with 1 tCO_2.

2.4. Sugarcane-biorefinery analyzer framework

The sugarcane-biorefinery analyzer framework (SBAF) starts with Eqs. (3) and (4), and comprehends all forthcoming equations in Secs. 2.3/2.4/2.5. SBAF calculates all FCI and cost terms, revenues, power effects and all critical streams from inputs X, Y, Z. To do this, the first step is to obtain Y and Z from X and LPS/MPS biorefinery requirements. For example, with Eq. (2), $Y$is given via Eq. (11a); and consequently,

$$(CRM + CUT)^{CHP} = \left(X\left(\frac{\dot{m}_{13}}{X}\right)^{*}(1 - Y/100)\left(\frac{\dot{m}_{CW}}{\dot{m}_{20}}\right)^{*} C_{CW} + XC_{X} + \dot{m}_{24}C_{DW}\right) OT*10^{-6} \tag{6}$$

$$\dot{m}_{CO2} = X\left(\frac{\dot{m}_{CO2}}{X}\right)^{*} \tag{7}$$

When a CCS plant exists, the biorefinery energy requirements supplied by the CHP must be increased, since LPS is required in the aqueous-MEA reboiler and electricity (W) is required to drive CO_2 compressors for pipeline transportation. LPS consumption and W are estimated via Eqs. (8) and (9) respectively, where $\left(\frac{\dot{m}^{CCS}_{LPS}}{X}\right)^{*}$ and $\left(\frac{W^{CCS}}{X}\right)^{*}$ are constant ratios from the Base-Case simulation with CCS. FCI^{CCS} and COM^{CCS} are obtained via Eq. (4) and Eq. (5), where $(CRM+CUT)^{CCS}$ comprehends only CW consumption via Eq. (10), where $\left(\frac{\dot{m}^{CCS}_{CW}}{X}\right)^{*}$ is another constant ratio from the Base-Case simulation with CCS.

since its maximum value is 100%, the minimum feasible X is obtained via Eq. (11b). Performing an energy balance around steam-turbines M − 21 and M − 22 (Fig. 3), one can show that the gross power $W^M$of M − 21 and M − 22 is obtained via Eq. (12), where $\Delta\widehat{H}_{S2-S1} \equiv \widehat{H}_{S2} - \widehat{H}_{S1}$ represents the difference of specific enthalpies S2 and S1, and η^{M}_{mech} is the mechanical efficiency of steam-turbines (Table 1). Therefore, knowing bagasse availability (X) and the biorefinery MPS/LPS requirements, one can obtain Zvia Eq. (2) and Yvia Eq. (11a). Terms $\Delta\widehat{H}_{S2-S1}$ are also constants from the Base-Case simulation, because the involved temperatures, pressures and compositions are invariant. The powers of pumps P-22 and P-21 (Fig. 3) are obtained via Eq. (13a) and Eq. (13b) respectively. The net CHP power W^{CHP} is obtained subtracting P-22 and P-21 powers from W^M in Eq. (14).

$$Y = 100\left(\dot{m}_{LPS} + \dot{m}_{MPS}\right)\left(X\left(\frac{\dot{m}_{13}}{X}\right)^{*}\right)^{-1} \tag{11a}$$

$$X_{min} = \left(\dot{m}_{LPS} + \dot{m}_{MPS}\right)\left(\left(\frac{\dot{m}_{13}}{X}\right)^{*}\right)^{-1} \tag{11b}$$

$$\frac{W^{M}}{\eta^{M}_{mech}} = \dot{m}_{MPS}\left(\frac{\Delta\widehat{H}_{13-14} + \Delta\widehat{H}_{13-17} + \Delta\widehat{H}_{19-20}}{(Y/100)(Z/100)} - \Delta\widehat{H}_{13-17} - \frac{\Delta\widehat{H}_{19-20}}{Z/100}\right) - \dot{m}_{16}\left(\Delta\widehat{H}_{13-17} + \Delta\widehat{H}_{19-20}\right) \tag{12}$$

$$\frac{W^{P22}}{\eta^{P22}_{mech}} = X\left(\frac{\dot{m}_{13}}{X}\right)^{*}\Delta\widehat{H}_{27-26} \tag{13a}$$

$$\frac{W^{P21}}{\eta^{P21}_{mech}} = \left(X\left(\frac{\dot{m}_{13}}{X}\right)^{*} - \frac{\dot{m}_{MPS}}{Z/100} - \dot{m}_{16}\right)\Delta\widehat{H}_{22-21} \tag{13b}$$

$$W^{CHP} = W^{M} - W^{P22} - W^{P21} \tag{14}$$

Performing mass and energy balances for the deaerator V-21 and using previous definitions, several streams related to V-21 are calculated: $\dot{m}_{26}$ via Eq. (15); $\dot{m}_{23}$ via Eq. (16); $\dot{m}_{24}$ via Eq. (17); $\dot{m}_{25}$ via Eq. (18); and finally the bleed-steam$\dot{m}_{16}$via Eq. (19), where mu, bb and dh represent, respectively, make-up percentage (2%), boiler blowdown percentage (1%) and percentage of MPS+LPS used as direct heat (50%).

$$\dot{m}_{26} = \frac{X}{(1 - bb/100)}\left(\frac{\dot{m}_{13}}{X}\right)^{*} \tag{15}$$

$$\dot{m}_{23} = (dh\,/\,100)\,(\dot{m}_{MPS} + \dot{m}_{LPS}\,) \tag{16}$$

$$\dot{m}_{24} = -\frac{X\left(\frac{\dot{m}_{13}}{X}\right)^{*}\left(1 - \frac{1}{1-bb/100} - mu\Big/100\right) + (\dot{m}_{MPS} + \dot{m}_{LPS}\,)(-1 + dh/100)}{1 - bb/100} \tag{17}$$

$$\dot{m}_{25} = X\left(\frac{\dot{m}_{13}}{X}\right)^{*}(mu\,/\,100) + \dot{m}_{24}(bb\,/\,100) \tag{18}$$

$$\dot{m}_{16} = \frac{\left(X\left(\frac{\dot{m}_{13}}{X}\right)^{*} - \dot{m}_{MPS} - \dot{m}_{LPS}\right)\widehat{H}_{22} + \dot{m}_{24}\widehat{H}_{24} + \dot{m}_{23}\widehat{H}_{23} - \dot{m}_{25}\widehat{H}_{25} - \dot{m}_{26}\widehat{H}_{26}}{\widehat{H}_{22} - \widehat{H}_{16}} \tag{19}$$

2.5. Application of the sugarcane-biorefinery analyzer framework

SBAF starts with the calibrated FCI predictor model, Eq. (3), which estimates FCI^{CHP} for a typical sugarcane biorefinery exporting electricity surplus for a given X and LPS/MPS consumptions, which are converted in terms of Y and Z. SBAF – Eqs. (3)–(19) – then estimates W^{CHP}, COM^{CHP} and all critical streams.

The Minimum Steam Selling-Price (*MSP*) of the produced steam of the Base-Case is evaluated with/without the CCS plant. For this purpose, a NPV analysis is performed using SBAF estimates for FCI, COM and revenues. NPV is evaluated via Eq. (20), given the total steam price and the annual interest (%) rate *j*, where *t* is the project lifetime plus construction time and *CF* is the cash flow, which depends on the steam price. As a simplification, the steam produced is the sum of the mass flowrates of LPS and MPS, and is referenced hereinafter as Total-Steam, with only one price. The *MSP* is the Total-Steam price giving $NPV = 0$ and using the minimal acceptable rate of return (MARR) as interest rate in Eq. (21). The Total-Steam *MSP*is evaluated instead of the electricity price due to its higher uncertainty, since the Brazilian electricity auctions define electricity contracts price.

$$NPV(j, Total - Steam\ Price) = \sum_{i=1}^{t}\frac{CF_i(Total - Steam\ Price)}{(1 + j/100)^{i}} \tag{20}$$

$$NPV(j = MARR, Total - Steam\ MSP) = \sum_{i=1}^{t}\frac{CF_i(Total - Steam\ MSP)}{(1 + MARR/100)^{i}} = 0 \tag{21}$$

To evaluate the BECCS cost, the CO_2 avoidance cost [38] is used. For sake of simplicity, it is here called Capture Cost (*CC*). *CC* is readily compared to CO_2 emissions tax and represents its minimum value that incentives carbon capture, when the taxation is applied to both BECCS and non-BECCS biorefinery [38]. A minor modification in the original equation is required to use it in the proposed scenario. The MSP (USD/t) is used instead of the electricity price, in Eq. (22), where "Reference" refers to the CHP plant without CCS and CO_2 *Emissions* is the ratio ($tCO2/t^{Total-Steam}$) of CO_2 emitted per Total-Steam produced.

$$CC = \frac{\left(MSP^{CCS} - MSP^{Reference}\right)}{CO_2\ Emissions^{Reference} - CO_2 Emissions^{CCS}} \tag{22}$$

Table 4 shows the required information for NPV and MSP analyses, also used for sensitivity analyses. Sensitivity analyses are performed for some parameters to evaluate possible feasible scenarios.

The Brazil Process Plant Cost Index from Intratec is used to update the publicly available 2018 Brazilian CHP costs. The values are 323.34 and 165.48 for 2018 and 2010, respectively [42]. The publicly costs are initially converted to Brazil currency BRL, then are multiplied by the ratio of the Index of the desired year by the Index of the original year, and finally, are converted back to USD in 2018. Moreover, USA BECCS costs are also updated to 2018 and compared to the results without considering the location-factor.

3. Results

The Base-Case ($X = 138\ t/h$, $Y = 80.565\%$, $Z = 7.929\%$) simulation results are shown in Table 5 and compared to literature values for validation. Boiler and CHP efficiencies and Power-to-Heat ratio are within the reported range, while %mol CO_2 in the dry-basis flue-gas is higher than the theoretical 13.07% [15]. This is due to the difference in carbon content considered for sugarcane bagasse, since it is represented as cellulose in this work. The total power W^{CHP} predicted by SBAF is close to similar literature simulation results, with divergences of ≈1%. Regarding the CCS plant, although the same method from a previous work for coal-fired plants [17] was used, a lower value for the reboiler duty was achieved, due to the higher %mol CO_2 in the sugarcane-biorefinery flue-gas (14%mol versus 7.69%mol), increasing

Table 4
Base-Case assumptions for economic analysis.

Parameter	Value	Reference
Engineering & Construction	1 year	APEA
Project Lifetime	20 years	Assumption
MARR	8%	Assumption
Reference Year	2018	Assumption
Bagasse Cost (BRL/t)	59.22	[39]
Operator/Supervisor Wage (USD/month/worker)	2710.4/5789.4	Assumption
2018 BRL/USD	3.6542	Assumption
FCI of 150 km CO_2 Pipeline	150 MMUSD	[6]
CW & Demineralized-Water costs (C_{CW},C_{DW})	4.29e-8 MMUSD/t	Assumption
Electricity Price (average of 2018 Brazilian auctions)	188 BRL/MWh	[40]
OT	5856 h/a (8 months)	[41]
Salvage Value	0%	Assumption
CO_2-to-EOR	25 USD/t	[36]
Cap-and-trade	40 USD/t	[36]

Table 5
Base-Case simulation results.

Process	Variable	Unit	This study	Literature
CHP	Boiler Efficiency	%	88.9%	69-88% [43]
	CHP Efficiency	%	64.9%	61-73% [43]
	Power-to-Heat Ratio	-	0.35	0.3–0.5 [43]
	Flue-Gas CO_2	%mol (dry-basis)	14.98%	13.07% [15]
	Total Power Generated	kWh	56.33	55.7 [29]
CCS with Aqueous-MEA	Reboiler Heat-Ratio	GJ/tCO_2	3.612	3.723 [17]
	Lean Loading	mol^{CO2}/mol^{MEA}	0.2230	0.2185 [17]
	Solvent Capture-Ratio	$kg^{Solvent}/kg^{CO2}$	12.89	13.32 [17]

Table 6
SBAF coefficients from Base-Case simulation.

Area	Coefficient	Value	Unit
CHP (A10 + A20)	$\Delta\hat{H}_{13-14}$	0.4694	GJ/t
	$\Delta\hat{H}_{13-17}$	0.2171	GJ/t
	$\Delta\hat{H}_{19-20}$	0.4583	GJ/t
	$\Delta\hat{H}_{27-26}$	0.01786	GJ/t
	$(\dot{m}_{13}/X)^*$	2.093	t/t
	$(\dot{m}_{CW}/\dot{m}_{20})^*$	63.03	t/t
	$\hat{H}_{16}, \hat{H}_{22}, \hat{H}_{23}$ $\hat{H}_{24}, \hat{H}_{25}, \hat{H}_{26}$	−13.00, -15.78, -15.31–15.71, -14.37, -15.54	GJ/t
	$(\dot{m}_{CO2}/X)^*$	0.7777	t/t
CCS (A30)	$(m^{CCS}_{LPS}/X)^*$	1.072	t/t
	$(W^{CCS}/X)^*$	90.50	kWh/t
	$(m^{CCS}_{CW}/X)^*$	23.11	t/t

Table 7
Estimated parameters for FCI predictors.

	FCI^{CHP} (A10 + A20)				FCI^{CCS} (A30)	
	a	b	c	d	$a^{\#}$	$b^{\#}$
Estimated value	2.368	0.5891	0.1670	0.03545	3.051	0.6808
Average absolute error	1.91%				1.31%	
Error range	[-4.09%, 3.55%]				[-1.7%, 1.8%]	
Correlation coefficient	0.9960				0.9911	

CO_2 fugacity. The CO_2 loading and the solvent capture-ratio also have better results than the conventional coal-plant CCS for the same reason.

The parameters obtained from simulation are shown in Table 6, while the adjusted parameters for the FCI predictor are shown in Table 7. The parameters for the FCI^{CHP} predictor Eq. (3) are kept constant whether CCS is performed or not. The small average errors of 1.9% and 0.139% respectively for FCI^{CHP} and FCI^{CCS}, together with the respective correlation coefficients (0.9960 and 0.9911), indicate a good fit of the FCI predictor models over the computational-experimental data. The scale-factor exponents for FCI^{CHP} and FCI^{CCS} are 0.5891 and 0.6808 respectively, meaning that the former is more benefited from higher capacities.

Fig. 4 shows observed (computational-experiments) and predicted FCI^{CHP} via Eq. (3), already considering the contingencies but without

Table 8
FCI^{CCS} predictor results versus literature results.

Case	Flue-Gas %mol CO_2	Capacity $tCO2^{Captured}$/h	FCI^{CCS} [a] MMUSD/$tCO2^{Captured}$/h	Source
550 MW Subcritical Pulverized Coal	12.88%	500.4	1.30	[38]
550 MW Supercritical Pulverized Coal	12.88%	480.4	1.32	[38]
559 MW NGCC	3.91%	202.1	1.87	[38]
430 MW Subcritical Pulverized Coal	12.80%	378.8	1.08[b]	[34]
430 MW Subcritical Pulverized Coal	12.80%	118.0	1.63[b]	[34]
Base-Case 56.3 MW CHP, X = 138 t/h	13.98%	96.60	1.51	This work
69 MW CHP, X = 169 t/h	13.98%	118.0	1.42	This work
118 MW CHP, X = 118 t/h		202.1	1.20	
221 MW CHP, X = 541 t/h		378.7	0.98	

[a] Only CO_2 capture and compression unit: USA values updated to 2018 via Intratec Process Plant Cost Indexes.
[b] Flue-gas desulfurization unit removed from FCI^{CHP}; Let-down turbine/generator included but not required in sugarcane-biorefinery CHP.

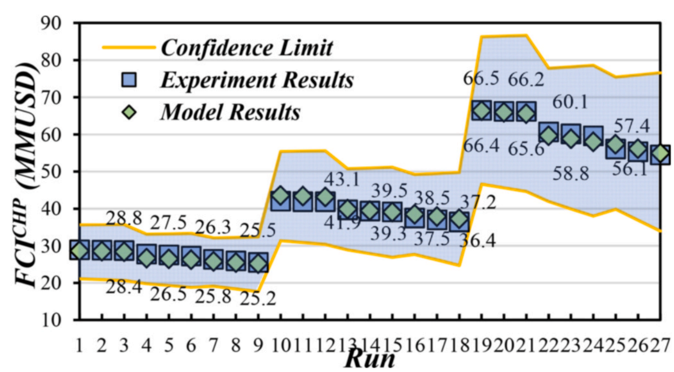

Fig. 4. Observed and predicted FCI^{CHP} results (no CCS).

CCS. One can see that X is the most impacting factor on FCI^{CHP}, which is the same in the 1–9, 10–18 and 19–27 runs. Furthermore, factor Y is more important than Z since there is a greater displacement for each triad of runs than inside each one. Factor Z only plays an important role when high X and Y are present. The estimated FCI^{CHP} is within 25–30 MMUSD for X = 33.75 t/h of bagasse, within 35–45 MMUSD for X = 67.5 t/h and within 55–65 MMUSD for X = 135 t/h. The confidence limits indicate the range wherein values of FCI^{CHP} with 15% maximum error (Class 4 project) are located with 95% of probability. If more accurate data is provided, the 95% confidence band would shrink. Fig. 5 depicts predicted versus observed FCI^{CHP} for all runs, together with 15% error bars and without CCS. Points are close to the diagonal, translating a good performance of the FCI^{CHP} predictor.

Fig. 6 depicts FCI^{CCS} versus X, including the investment of a 150 km CO_2 pipeline. It is notorious that FCI^{CCS} is considerably higher than FCI^{CHP}. For example, at X = 67.5 t/h of bagasse, FCI^{CCS} is 200–300 MMUSD, while FCI^{CHP} is within 30–60 MMUSD (Fig. 4) without CCS. The pipeline FCI accounts for a good portion of FCI^{CCS}, ranging from 50% to 75%; but the higher the capacity, the lower the unitary (i.e., per tCO2) pipeline contribution.

Table 8 compares sugarcane-biorefinery FCI^{CCS} per capture capacity

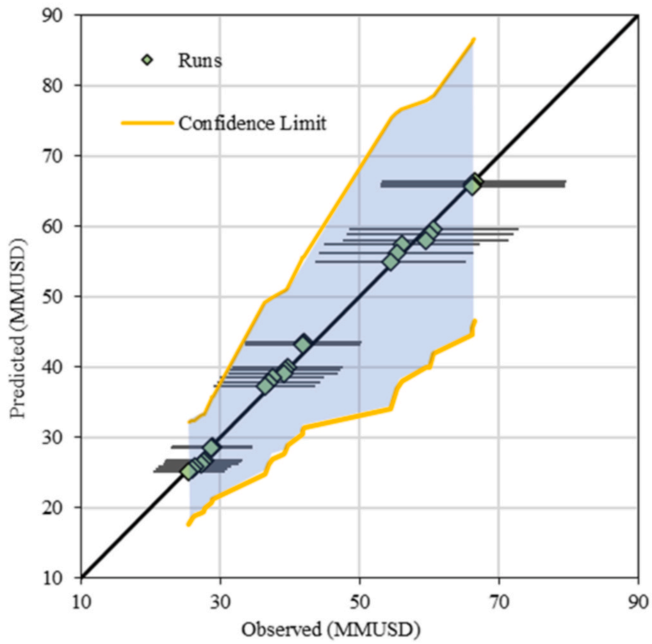

Fig. 5. Observed versus predicted FCI^{CHP} (A10 + A20, no CCS).

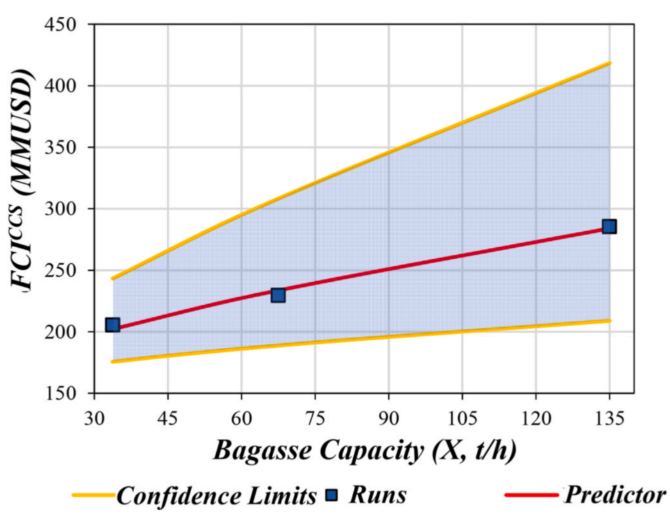

Fig. 6. Observed and predicted FCI^{CCS} (A30) versus bagasse capacity.

(MMUSD/$tCO2^{Captured}$/h) with 2018 USA plants. The comparison only considers the FCI of the CO_2 capture and compression unit. The FCI^{CCS} predictor of this work is close to the results of [34], showing divergences of 15% and 10% for capacities of 118 and 379 $tCO2^{Captured}$/h, respectively. However, specific values from Ref. [34] are lower than the counterparts of [38]: 1.08 MMUSD/$tCO2^{Captured}$/h against 1.3 MMUSD/$tCO2^{Captured}$/h, even at high capture capacities of 378.8 $tCO2^{Captured}$/h and 500.4 $tCO2^{Captured}$/h, respectively. Fig. 7 shows COM^{CHP} with/without CCS and COM^{CCS} versus simulation runs. FCI^{CHP} increases considerably with CCS up to 34% in the analyzed scenarios, for the same power output. The numbers of operators/supervisors are 2/1 for the CHP and 5/1 for the CCS plant, resulting in COL values of 0.67 MMUSD/a and 1.16 MMUSD/a, respectively. It can be seen that the trend of COM^{CHP} is similar to the FCI^{CHP} analogue; i.e., X impacts more than Y, which impacts more than Z. The underlying reason is the fact that FCI is a relevant factor in COM with a minimum contribution of 86%; on the other hand, a minor (but not insignificant) tributary to COM is COL, the cost of labor.

Fig. 8 shows the FCI^{CHP} behavior (no CCS included) against power exported for a typical sugar-bioethanol sugarcane-biorefinery, and its comparison with the literature. The specific FCI^{CHP} (USD/kW) is also shown in the right axis. This time, the fitted FCI^{CHP} equation does not depend on (Y,Z) because FCI^{CHP} was handled to match a typical biorefinery configuration requirements; i.e., the change in X does not change the other factor values of the Base-Case ($Y = 80.5653\%$, $Z = 7.9287\%$). The FCI^{CHP} in Brazil for 500 t/h sugarcane using an 82 bar boiler was estimated at 126 MMBRL in 2010 [33], or a 2018 updated value of 65.6 MMUSD. Brazilian 2018 electricity auctions gave an average specific FCI^{CHP} of 4000 BRL/kW with 50 MW of power average [41]. The SBAF model agrees with the available public data since they are within the considered error margin for a preliminary study. However, the specific FCI^{CHP} from electricity auctions in Brazil vary significantly from 1000 BRL/kW to 8000 BRL/kW, even considering only the new ventures, because sometimes they are installed next to a pre-existing plant sharing infrastructure and reducing costs [41]. Even so, the average 2008 specific FCI^{CHP} of 4000 BRL/kW for 50 MW of power average [41] shows that SBAF reasonably represents the Brazilian scenario.

The MSP of the Base-Case (no CCS) Total-Steam was found at 11.3 USD/t. Fig. 9 shows the composition of the destination of total revenue for the Base-Case without CCS considering a Total-Steam price of 23.7 USD/t. Fig. 9 shows that 18.1% of revenues is destinated to the Return of Investment (*ROI*) in order to achieve the investment *MARR*, which is obviously the most important revenue destination after raw-materials (basically sugarcane) compromising 27.5% of revenues. The smallest revenue destination is COL, accounting only for 2.6% of revenues.

Fig. 10 shows one-dimensional *MSP* sensitivity analyses regarding changes in parameters and factors (i.e, considering one parameter at a time). Bagasse cost (Base-Case = 11 USD/t), electricity price (Base-Case = 51.45 USD/MWh), bagasse capacity (Base-Case = 138 t/h), Capacity Factor (Base-Case = 100%), OT (Base-Case = 5856 h/a) and FCI^{CHP} (Base-Case = 58.72 MMUSD) were varied by ±50%, except for the Capacity factor, which was varied from −50% to 0%. One can see that an increase in electricity price has a higher effect on *MSP* than a decrease in

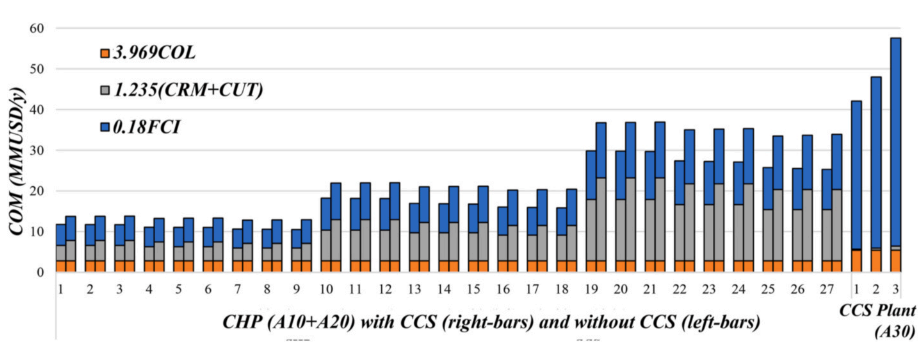

Fig. 7. COM^{CHP} with/without CCS and COM^{CCS} vs simulated runs.

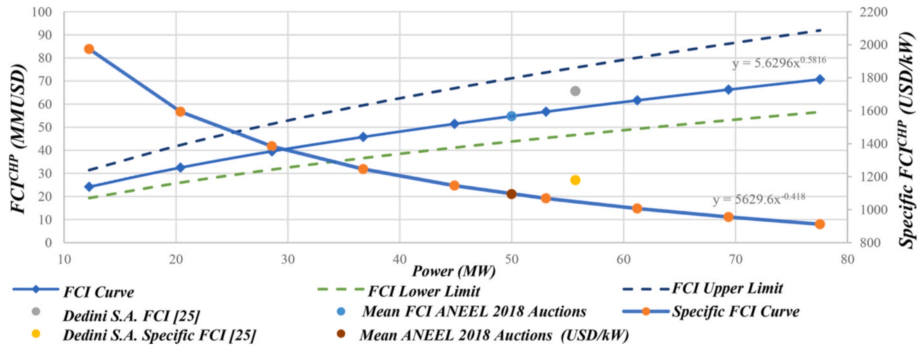

Fig. 8. Absolute FCI^{CHP} (MMUSD) and specific FCI^{CHP} (USD/kW) of typical sugar/ethanol biorefineries (Brazil, 2018) and comparison with literature.

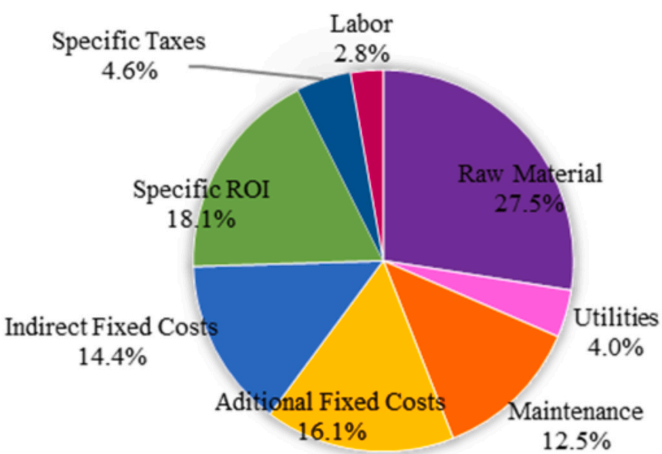

Fig. 9. Revenue destination: Base-Case (no CCS).

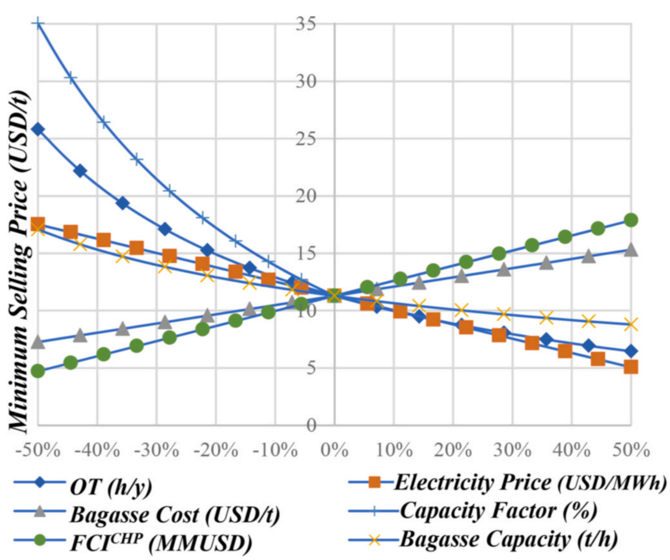

Fig. 10. Base-Case (no CCS): Total-Steam MSP sensitivity to changes in electricity price, bagasse price, bagasse capacity, capacity factor, OT and FCI^{CHP}.

bagasse cost, 5.1 USD/t against 7.3 USD/t for a 50% change. An increase in CHP capacity is only recommended when there is a real demand for the extra power generated, since MSP sensitivity to the Capacity Factor is the greatest among the analyzed parameters. For example, a capacity increase from the Base-Case to 207 t/h bagasse capacity decreases the *MSP* from 11.1 to 8.8 USD/t; but if the Capacity Factor is 40% lower, it can double the MSP. The decrease in FCI^{CHP} – e.g., CHP sharing some infrastructure or revamping an existing plant – can have a greater MSP impact. For each MMUSD saved in FCI^{CHP}, a MSP decrease of 0.22 USD/t is achieved. Moreover, some plants burn alternative biomass fuels in the off-season to increase its yearly OT, reducing *MSP* to 6.5 USD/t for CHP plants available all over the year.

For the Base-Case with CCS, the increase of LPS requirement due to the stripping heat-ratio in the post-combustion aqueous-MEA plant is estimated as 157.4 t/h. Thus, the bagasse requirement increases 57.7 t/h to keep the same power output (56.3 MW). The new requirement values of bagasse, LPS, and MPS (respectively 195.7 t/h, 371.7 t/h and 18.45 t/h) result in$X = 195.7\ t/h$, $Y = 95.2\%$, $Z = 4.7\%$.

The captured CO_2 of 96.6 t/h represents 90% of the original CO_2 emitted and 63.5% of the actual CO_2 produced, since there is an increase in the bagasse requirement and, consequently, in the CO_2 produced (152.2 t/h). The CO_2 exportation revenue is 36.8 MMUSD/a, wherein 38.5% is for EOR purposes and the remainder for Cap-and-Trade market. The gross CHP produced power is 75.9 MW, with a net value of 56.28 MW, discounting the power of CO_2 pump and compressors. This CHP net power with CCS is practically equal to the CHP net power without CCS (56.33 MW). The new FCI^{CHP} with CCS is 71.8 MMUSD, a 22.3% increase over the Base-Case without CCS, while FCI^{CCS} reaches 256 MMUSD, wherein the CO_2 pipeline accounts for ≈60% of FCI^{CCS}. Table 9 compares both scenarios in terms of technical and economic metrics.

In the Base-Case with CCS, considering the CO_2 revenue, the *CC*is 262 USD/tCO_2, perfectly within the literature range 88–288 USD/tCO_2, while the MSP is 69.5 USD/t. The CC proximity to the upper limit has probably to do with the technology chosen; namely, post-combustion capture with 30%w/w aqueous-MEA, while the lower limit is normally associated to oxy-fuel technologies [24]. This is also the increase of total revenue that must be fulfilled to pay for CCS operations while keeping the same NPV of the Base-Case without CCS; otherwise, the NPV would become negative in 647 MMUSD, considering the steam value of 11.3 USD/t. Fig. 11 shows the composition of the total revenue destination for the Base-Case with CCS. The main difference from the Base-Case without CCS (Fig. 10) is the increase in the Specific ROI composition (18.1%–28.3%) to achieve the*MARR*, due to a higher FCI (FCI^{CHP}+FCI^{CCS}), resulting in higher COM as well.

Sensitivity analyses were performed for the *CC*of the Base-Case with CCS by means of ±50% variations on the following items: FCI^{CHP}, CO_2

Table 9
Comparison of CHP performances with/without CCS.

Variable	No CCS	With CCS
Inputs (LPS(t/h); MPS(t/h); X(t/h))	214.3; 18.45; 138.0	371.7; 18.45; 195.71
Y (%); Z(%)	80.6%; 7.93%	95.2%; 4.73%
FCI^{CHP}+FCI^{CCS} (MMUSD)	58.7 + 0	71.8 + 256
COM^{CHP} (MMUSD/y)	25.0	95.4
$CO_2^{Flue\text{-}Gas}$ (t/h)	107.3	152.2
$CO_2^{Captured}$ (t/h)	0	96.6
CCS Efficiency (%)	0%	63.5%
W^{CHP}(MW)	56.3	56.3

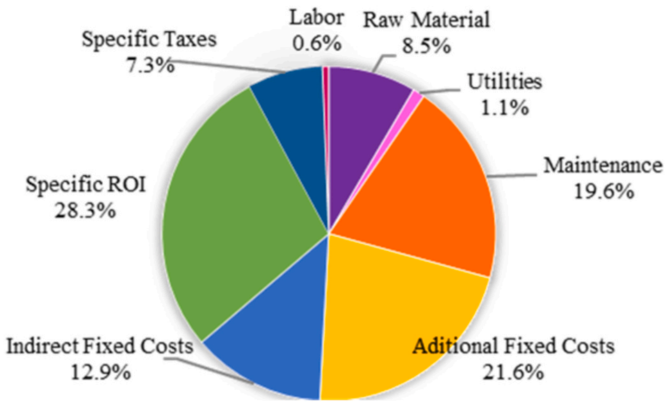

Fig. 11. Revenue destination: Base-Case with CCS.

revenues (cap-and-trade and CO_2-to-EOR), OT, electricity price and bagasse capacity without CCS – 138 t/h in the Base-Case without CCS, even if extra bagasse is required to perform CCS, since this is an input to SBAF model and the total bagasse is adjusted to give the same power output. To perform the sensitivity analysis, the following procedure is used: (i) run SBAF without CCS at the new perturbed condition (ignoring changes in FCI^{CHP} or in CO_2 revenue, if they are the perturbed factors); (ii) evaluate and record the Total-Steam MSP; (iii) run SBAF for the new perturbed condition with CCS and adjust *X* to achieve the same power output; (iv) estimate the new MSP; (v) evaluate CC; (vi) repeat for all perturbed conditions.

Results of sensitivity analyses are shown in Fig. 12. CHP with CCS is only economically advantageous in the event of carbon taxation (40–80 USD/tCO_2 [6]) under extreme conditions (as in a total FCI lower than 240 MMUSD) resulting in CC = 50 USD/tCO_2. CC has a low sensitivity to CO_2 revenues of 3.3 USD/tCO_2 for each MMUSD/a increased; i.e., for a 100% increase of CO_2 revenues, other factors unchanged, *CC* would still be 141 USD/tCO_2. OT and bagasse capacity can also contribute to feasibility: a 50% increase in bagasse capacity reduces CC to 179.4 USD/tCO_2, while reaching 139.7 USD/t for 50% OT increase. *CC* is insensitive to electricity price because Total-Steam *MSP* changes to counterbalance electricity price changes, maintaining total revenue constant.

A combination of changes of −20% in FCI^{CHP}, +25% in CO_2 revenue, +20% in OT and +50% in bagasse capacity – all of them plausible changes in the short-term – result in $CC = 78.5\ USD/tCO_2$ starting to be more advantageous than a taxation of 80 USD/tCO_2. However, the actual case is far from feasible and capturing only CO_2 from the fermentation step should be considered first as a BECCS option. Moreover, considering 378.8 t/h of CO_2 capture and OT = 8000 h/a (i.e., like any conventional power plant), *CC* becomes only 17.2 USD/tCO_2. This shows that the main drawback of the sugarcane-biorefinery enterprise is its low agricultural-based capacity and the consequent low OT, resulting in a high downtime of an expensive BECCS plant.

4. Conclusions

This work analyzed investments in CHP units to supply heat and power to typical Brazilian sugarcane-biorefineries with/without CCS. A robust Sugarcane-Biorefinery Analyzer Framework (SBAF) was developed to assist in sugarcane-biorefinery BECCS decision making. SBAF solves CHP/CCS mass-energy balances, simultaneously estimating net electricity exportation, CO_2 emissions, CO_2 revenues, besides COM^{CHP} and COM^{CCS} together with surface response models for FCI^{CHP} and FCI^{CCS}. With SBAF it is possible to predict FCI^{CHP}, FCI^{CCS}, COM^{CHP} and COM^{CCS} of both scenarios (with or without CCS) in a simple way, regarding only bagasse availability and LPS/MPS requirements. The average errors of FCI^{CHP} and FCI^{CCS} models against the observed values

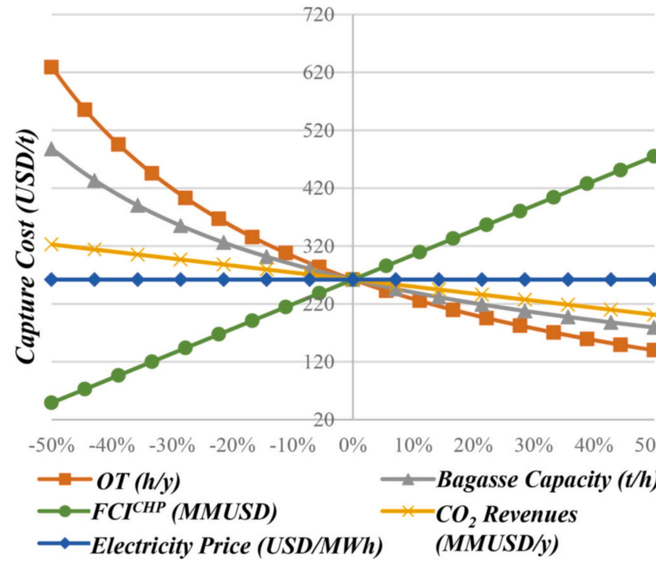

Fig. 12. Base-Case with CCCS: Capture Cost sensitivity to changes in FCI^{CHP}, CO_2 revenues, bagasse capacity without CCS, electricity price and OT.

were 1.9% and 1.3% with correlation coefficients of 0.996 and 0.991, respectively. When SBAF is used to calculate the requirements of a typical ethanol/sugar sugarcane-biorefinery with electricity surplus, its results are compatible with the Brazilian auctions and with literature data.

The CHP capacity in terms of bagasse consumption is the most important input factor in FCI^{CHP}, followed by the SHPS percentage (Y) converted into MPS/LPS, while the less important factor is the SHPS percentage converted into MPS and bleed-steam. The decision to invest in a BECCS sugarcane-biorefinery, under constant electricity exportation, can increase FCI^{CHP} as high as 34%, while also similarly increasing COM^{CHP} fueled by its dependency on FCI^{CHP}, which responds for more than 40% of COM^{CHP}. The FCI^{CCS}, by its turn, has the pipeline investment as its biggest burden, which accounts for more than 50% of FCI^{CCS}, while FCI^{CCS} is, at least, 4 times higher than FCI^{CHP} without CCS. These values evince a certain lack of viability, due to a considerable increase in investment and costs, but deprived of a proper revenue increase from the captured CO_2 due to its low commercial value of at most 100 USD/tCO_2.

The CC of the Base-Case with CCS is estimated as 262 USD/tCO_2, within the literature range 88–288 USD/tCO_2. Sensitivity analyses were performed and almost all perturbed scenarios resulted in CC within the literature range. Such analyses show that CC can drop to 50 USD/tCO_2 if FCI^{CH}+FCI^{CCS} reduce by 50%, the most sensible factor. A further combination of factor changes can reduce CC to less than 80 USD/tCO_2. Moreover, assuming OT and capture capacity typical of conventional power plants leads to CC of only 17.2 USD/tCO_2, showing that they represent the biggest challenges for BECCS sugarcane-biorefineries.

Author credit statement

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Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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