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Energetic, Exergetic, and Techno-Economic Analysis of A Bioenergy with Carbon Capture and Utilization Process via Integrated Torrefaction–CLC–Methanation

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Abstract: Bioenergy with carbon capture and storage (BECCS) or utilization (BECCU) allows net zero or negative carbon emissions and can be a breakthrough technology for climate change mitigation. This work consists of an energetic, exergetic, and economic analysis of an integrated process based on chemical looping combustion of solar-torrefied agro-industrial residues, followed by methanation of the concentrated CO₂ stream with green H₂. Four agro-industrial residues and four Italian site locations are considered. Depending on the considered biomass, the integrated plant processes about 18–93 kg h^{−1} of raw biomass and produces 55–70 t y^{−1} of synthetic methane. Global exergetic efficiencies ranged within 45–60% and 67–77% when neglecting and considering, respectively, the valorization of torgas. Sugar beet pulp and grape marc required a non-negligible input exergy flow for the torrefaction, due to the high moisture content of the raw biomasses. However, for these biomasses, the water released during drying/torrefaction and CO₂ methanation could be recycled to the electrolyzer to eliminate external water consumption, thus allowing for a more sustainable use of water resources. For olive stones and hemp hurd, this water recycling brings, instead, a reduction of approximately 65% in water needs. A round-trip electric efficiency of 28% was estimated assuming an electric conversion efficiency of 40%. According to the economic analysis, the total plant costs ranged within 3–5 M€ depending on the biomass and site location considered. The levelized cost of methane (LCOM) ranged within 4.3–8.9 € kg_{CH₄}^{−1} but, if implementing strategies to avoid the use of a large temporary H₂ storage vessel, can be decreased to 2.6–5.3 € kg_{CH₄}^{−1}. Lower values are obtained when considering hemp hurd and grape marc as raw biomasses, and when locating the PV field in the south of Italy. Even in the best scenario, values of LCOM are out of the market if compared to current natural gas prices, but they might become competitive with the introduction of a carbon tax or through government incentives for the purchase of the PV field and/or electrolyzer.



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1. Introduction

Institutions, policymakers, and teams of experts are depicting scenarios toward a renewable energy transition, simultaneously assessing current progresses to reduce gaps and underline new opportunities [1–4]. Inside the Net Zero Roadmap [1], the International Energy Agency (IEA) stated that achieving net zero emissions by 2050 requires rapid and deep cuts in emissions of both CO₂ and other greenhouse gases, and identified bioenergy as one of the pillars for the clean energy transition. According to IEA, the share of energy supply through bioenergy should grow from the current 6% to 18% in 2050 [1]. Integrating bio-based energy sources with carbon capture and storage processes, currently referred to as bioenergy with carbon capture and storage (BECCS), can result in negative CO₂ emissions,

facilitating the transition towards a net-zero economy. Alternatively, the concentrated CO₂ stream from bio-based energy sources can be exploited for the production of synthetic fuels (e.g., methane, methanol) in processes currently referred to as bioenergy with carbon capture and utilization (BECCU) [5,6]. Considering CO₂ as a valuable feedstock for the production of materials and fuels paves the way for a new market model by replacing the products derived from fossil sources. This new paradigm can reduce the emissions of greenhouse gases [6]. The protection of carbon sinks and the potential impact on land use deriving from biomass supply chains may lead to criticisms on the utilization of bioenergy [7]. These issues might, however, be of secondary importance in the case of residual waste biomass [8,9]. Nevertheless, optimizing the transformation processes of such renewable carbon sources is crucial for preserving ecosystems.

Direct utilization of biomasses is sometimes hindered by their poor properties in terms of heating value, moisture content, and flowability. Moreover, the possible presence of micro-organisms and the high hydrophilicity make their chemical–physical properties susceptible to changes over time, posing potential problems to downstream operations. Torrefaction has often been proposed as a thermochemical treatment to improve the utilization of biomasses as fuel. It is indeed able to reduce the moisture content, increase the heating value following a reduction in H/C and O/C ratios, improve flowability/fluidizability, enhance grindability, and stabilize its chemical–physical structure and composition [10,11]. Torrefaction is conducted by heating the biomass at a temperature within 200–300 °C under an inert atmosphere (e.g., N₂, Ar), for reaction times within 5–30 min. A torrefied gas is also produced, which may contain valuable chemical species but typically diluted by the inert gas stream. Torrefaction has been largely explored in the literature in recent years, and the use of fluidized beds [12], rotary kilns [13], and moving bed [14] reactors has been widely investigated.

The exploitation of biomasses for material/energy generation with simultaneous production of a concentrated CO₂ stream for subsequent storage or utilization can occur through various techniques. These are typically distinguished in post-combustion, pre-combustion, and oxy-combustion. Post-combustion techniques, such as calcium looping, absorption, and adsorption, rely on concentrating the CO₂ stream after a conventional (i.e., with air) combustion process. They have the advantage of being retrofittable, but typically require an external energy supply to regenerate the CO₂ sorbent. Pre- and oxy-combustion techniques necessitate instead of important modifications to already existing industrial plants, if not the design of novel power plants. Among them, chemical looping combustion (CLC) emerges as one of the most promising technologies thanks to its high potential efficiency, and to the absence of an expensive air-separation unit. CLC splits the conventional combustion in a two-step process. In the first step, oxidation of the fuel is performed by means of a solid compound, termed oxygen carrier, which provides the required oxygen by reducing itself. In the second step, the oxygen carrier is regenerated through oxidation with atmospheric air. In this way, direct contact between fuel and air is avoided, and a concentrated CO₂ stream (upon water condensation) is produced after fuel oxidation. The oxygen carrier typically consists of an active metal oxide (e.g., CuO, NiO, Mn₂O₃, Fe₂O₃), dispersed on an inert support (e.g., Al₂O₃, ZrO₂, Fe₂O₃/Fe₃O₄/FeO) that ensures high specific surface/porosity and enhanced mechanical stability [15]. CLC is typically investigated in dual interconnected fluidized bed systems [16,17], but dynamic operation of packed beds has also been proposed and investigated [18].

The conversion of concentrated CO₂ streams into renewable fuels through reaction with green H₂ is largely investigated in the current literature. This approach can increase the penetration of renewable energies, enable thermochemical energy storage of renewable energy into valuable fuels, and help develop a circular carbon and hydrogen economy [19]. Production of a synthetic natural gas stream via the Sabatier reaction (Equation (1)) is one possible strategy:



The reaction is strongly exothermic and thermodynamically favored at temperatures within 300–500 °C, but a catalyst is required to boost reaction kinetics [20]. Ni-based catalysts are among the most investigated thanks to their good combination of activity, selectivity, abundance, and low price. Ru has also been widely investigated as a catalyst thanks to its greater activity at lower temperatures, but its cost is significantly higher [21]. Several reactor concepts have been investigated in the literature to deal with the strong exothermicity of the methanation reaction to both avoid catalyst sintering due to strong temperature increase and/or hotspot formation and boost the reaction conversion degree. Adiabatic packed beds with intercooling and optional water condensation/product recycle are among the most investigated configurations, already implemented on an industrial scale [22–24]. Fluidized bed reactors have been investigated as they can allow for almost isothermal operations and can maximize gas–solid mass transfer, with potential benefits on reaction conversion degree and catalyst sintering. However, the strong volume reduction caused by the chemical reaction can pose problems to fluidization, and comminution/elutriation phenomena can reduce the catalyst lifetime [22,25]. Microstructured packed bed and multi-tubular cooled reactors have also been proposed because their high surface/volume ratio enhances heat and mass transfer coefficients, thus allowing high conversion degrees while reducing hotspot formation [9,26,27].

Integration of methanation with different processes is widely investigated in the current literature. Several studies focus on integration with direct air capture technologies. On this topic, Veselovskaya et al. tested the use of K_2CO_3/Al_2O_3 as a sorbent for CO_2 capture and both a catalyst based on either Ni [28] or Ru [29] for the methanation process. Tregambi et al. investigated instead the use of calcium looping for direct air capture and performed both a modelling [30] and techno-economic [31] study. Techno-economic studies on this integration were also performed by Peters et al. [32] and Harada et al. [33]. Integration with gasification processes from either fossil fuels [34], biomass [35–37], or municipal solid waste [38] is also widely assessed. Moreover, several studies are currently investigating integrated CO_2 capture and methanation by means of dual function materials, so as to couple the exothermic methanation reaction with the endothermic sorbent regeneration step, thus allowing higher overall process efficiencies [39–42]. Closer to the topic of the present work, the integration of combustion processes and methanation is also widely assessed. Fan et al. [43] performed a thermodynamic assessment and process simulation of an integrated methanation–CLC process using a co-feed of biomass and coal. Bailera et al. [44] investigated a hybrid oxy-combustion and power-to-gas system, in order to avoid the penalty ascribed to the CO_2 capture step. Kermani and Houshfar [45] analyzed the energetic and exergetic performance of a biomass (wood chips and wheat straw chopped wood) combustion plant integrated with post-combustion CO_2 separation via absorption and subsequent methanation. Recently, Fleiß et al. [46] successfully demonstrated the operation of an 80 kW_{th} CLC unit fed by bark mulch with direct utilization of the flue gas stream in a fluidized bed methanator and performed a scale-up study to a 100 MW_{th} plant. Finally, Brachi et al. [9] investigated the performance of an integrated CLC–methanation plant relying on solar torrefied agro-industrial biomasses as input fuel through model computations. The potential of the plant was estimated in terms of energy generation, produced methane, and electrical energy consumption due to both torrefaction and H_2 production, with reference to four different biomasses and four different Italian site locations.

BECCU with CO_2 conversion into renewable fuels is thus gaining significant attention to increase the penetration of renewable energies, reduce greenhouse gas emissions, and convert H_2 into more easily exploitable fuels [9,43–46]. This study aims to contribute to the existing literature on the topic by estimating key energetic and economic performances of an integrated process targeting both BECCU and methane production. The objective is to assess the competitiveness of the process with natural gas, and to identify the bottlenecks of the technology, offering insights into areas requiring further improvement. In particular, this study builds upon the modeling work by Brachi et al. [9] and is aimed at performing a comprehensive energetic, exergetic, and techno-economic analysis of the integrated CLC–

methanation process proposed in the previous companion paper. Key process indicators such as exergetic efficiency, total plant costs, and levelized cost of methane are estimated for the different options (i.e., biomass type and site location) and under different parameters, in order to understand the potential commercial applications of the proposed process layout and to analyze how variations in biomass type and site locations affect its performance and potential.

2. Process Description

The layout of the integrated process has been described in the companion research paper by Brachi et al. [9], but is briefly recalled in this section, and sketched in Figure 1, for the readers' convenience. The process starts with the torrefaction of a raw residual biomass. This step increases the biomass heating value, reduces the water content (that can be as high as 60%_{wt} in some cases [9]), and enhances/stabilizes its chemical–physical properties, thus improving its subsequent use as fuel. The energy needed for the torrefaction process comes from a dedicated photovoltaic (PV) array. The biomass is processed at different throughputs during the year, depending on solar energy availability, and then stored in a temporary storage silo during peaks of production. A share of the solar energy used for the torrefaction is recovered through a heat exchanger/condenser of the steam released. A torrefied gas, which may contain valuable fuel compounds but often largely diluted by the inert process gas used, is also produced. The torrefied biomass is then fed to a CLC unit to produce energy and a concentrated CO₂ stream ready for subsequent utilization. On a parallel line, an electrolyzer powered by a dedicated PV field produces a H₂ stream. To account for the intrinsic variability of the renewable solar source, a portion of the stream is temporarily stored in a vessel during peak production periods. It is then mixed in a 4:1 molar ratio with the concentrated CO₂ stream, as required by the stoichiometry of reaction (Equation (1)), and sent to a methanation unit. For the methanation reactors, the use of three in-series adiabatic packed beds with intermediate cooling, such as in [24,30], has been considered here instead of the microreactor originally proposed in the companion paper [9], because of the higher technology readiness level of this solution, already implemented at industrial scale. A synthetic natural gas stream leaves the methanation unit. Moreover, heat is released because of the exothermicity of the Sabatier chemical reaction, which is recovered by means of heat exchangers/condensers.

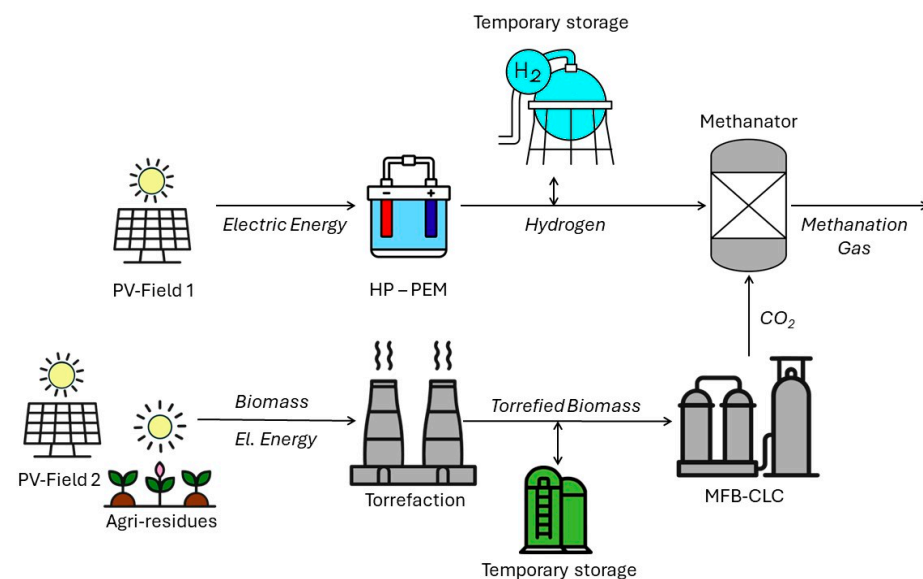


Figure 1. Sketch of the integrated process layout.

Four different agro-industrial residues from the Italian territory have been considered, namely, sugar beet pulp (SBP), olive stones (OS), grape marc (GM), and hemp hurd (HH),

and four different Italian site locations for the PV field, ranging from north to south (i.e., Brescia, Viterbo, Lecce, Caltanissetta).

3. Methodology

3.1. Energetic and Exergetic Analysis

An energetic analysis was performed on the integrated plant to evaluate the enthalpy fluxes through the different subunits and individual equipment of the plant (e.g., reactors, heat exchangers/condensers). However, the energetic analysis does not allow an easy energy distinction according to its quality (e.g., low or high temperature). Conversely, the exergetic analysis considers how energy is transformed and degraded during the process, identifying the potential work that can be extracted from each stream when brought into chemical and physical equilibrium with the environment. This analysis results in useful considerations and opportunities for optimization, which can enhance the overall efficiency of the system. Exergetic analysis then allows a first classification of the biomasses analyzed, regardless of the site location. Exergetic efficiencies were computed with reference to both the individual plant subsections (e.g., torrefaction, CLC, methanation), and the overall process, as the ratio between output and input overall exergy fluxes. Moreover, a round-trip electric exergetic efficiency ($\eta_{\text{round-trip}}$) was evaluated for the methanation process, by converting the output chemical and thermal exergy to electricity through an efficiency factor (η_{el}) and then comparing the outcome with the electrical input power to the electrolyzer. The round-trip electric exergetic efficiency is associated with the concept of renewable energy storage [47] since it indicates how much renewable electricity can be stored into methane, and then reconverted into electricity.

For the analysis, it is important to evaluate the exergetic flows entering and exiting the selected control domains. Within the integrated process, exergetic flows are associated with inflow/outflow of matter, heat, and work transfer [48]. Exergetic flows from heat transfer at constant temperature were computed as:

$$\dot{E}x_q = \tau \dot{Q} \quad (2)$$

where $\dot{Q}$ is the outflow heat from the control region and τ is the Carnot efficiency, evaluated for a Carnot machine operating between the environment, at temperature T_0 , and the process, at temperature T .

Exergetic flows associated to inflow/outflow of matter are computed as the sum of physical and chemical exergy, thus neglecting potential and kinetic exergy variations. Specific physical exergy was calculated as:

$$ex_{ph} = \Delta h - T_0 \Delta s \quad (3)$$

where Δh and Δs are the specific enthalpy and entropy variations, respectively. In particular, for an ideal gas mixture [49]:

$$(ex_{ph})_M = (T - T_0) \sum_{i=1}^N x_i \tilde{c}_{p,i}^{\varepsilon} + \tilde{R} T_0 \ln \left(\frac{P_i}{P_0} \right) \quad (4)$$

where x_i is the molar fraction of the i chemical species and $\tilde{c}_{p,i}^{\varepsilon}$ is its mean molar isobaric exergy capacity, $\tilde{R}$ is the universal gas constant, P_i is the partial pressure of the i chemical species, and P_0 the reference pressure. Mean molar isobaric capacity were computed as [48]:

$$\tilde{c}_p^{\varepsilon} = \frac{1}{T - T^*} \left[\int_{T^*}^T \tilde{c}_p dT - T_0 \int_{T^*}^T \frac{\tilde{c}_p dT}{T} \right] \quad (5)$$

where $\tilde{c}_p$ is the molar mean specific heat of the chemical species considered and T^* the reference temperature. Thermodynamic parameters were obtained from the NIST-JANAF tables [50]. The specific chemical exergy of a mixture was instead computed as:

$$ex_{ch,M} = \sum_i x_i ex_{ch,i} + \tilde{R}T_0 \sum_i x_i \ln x_i \quad (6)$$

where $ex_{ch,i}$ is the standard specific chemical exergy of the given species, obtained from tabulated values [48].

The thermodynamic state of the streams leaving the system toward the environment was set at 50 °C and 1 atm. The remaining exergy is considered as an irreversibility because it is assumed to come into equilibrium with the environment without delivering useful work. The environment is considered for each studied case at a temperature $T_0 = 25$ °C and a pressure of 1 atm.

Following the sketch outlined in Figure 1, different fluxes enter or leave the system depending on the specific sub-unit considered. Concerning the torrefaction, two exergetic fluxes enter the system. One is the electric exergy flow delivered by the PV field, equal to the power required by the torrefaction process as detailed in Brachi et al. [9]. The other is the chemical exergy of the incoming raw biomass. Following literature correlations [51], the latter was computed as:

$$ex_{ch,biomass} = \beta_{biomass} LHV_{biomass} \quad (7)$$

where $LHV_{biomass}$ is the lower heating value of the biomass and $\beta_{biomass}$ is a coefficient that depends on the biomass composition:

$$\beta = \frac{1.0412 + 0.2160\left(\frac{H}{C}\right) + 0.2499\left(\frac{O}{C}\right)\left[1 + 0.7884\left(\frac{H}{C}\right)\right] + 0.0450\left(\frac{N}{C}\right)}{1 - 0.3035\left(\frac{O}{C}\right)} \quad (8)$$

where H , C , N , and O are the weight fractions of the given elements within the biomass, as obtained from elemental analysis [9]. The exergetic fluxes leaving the system are as follows: (i) outflow of torrefied biomass; (ii) heat transfer from condensation and cooling of steam produced from biomasses drying; and (iii) outflow of torrefaction gas (termed torgas in the following). The exergy content of the torrefied biomass was assessed mirroring the approach employed for the raw biomass. The exergy content of the steam was calculated following Equations (3)–(5). The chemical exergy content of torgas was instead evaluated as the difference between the exergy content of the torrefied biomass and that of the dry biomass [52,53], thus neglecting the heat of the torrefaction reaction. Physical exergy content was computed assuming a specific heat capacity of $1600 \text{ kJ kg}^{-1} \text{ K}^{-1}$ [54] and applying Equations (4) and (5). It is, however, worth noting that the chemical exergy of the torgas is not easily exploitable, since valuable fuel compounds might be excessively diluted by the inert stream used for the process [51]. In the literature, it is sometimes co-fed with biomass to reduce the heat required for the torrefaction. If the torgas chemical exergy is not used, it becomes an irreversibility in the exergetic analysis. In the discussion of the results, both scenarios will be considered: one in which the torgas chemical exergy is considered to be exploitable, and the other in which it is not.

Concerning the CLC, one exergetic flux enters the system, which is related to the torrefied biomass. Outlet fluxes are instead due to the following: (i) heat transfer extracted at constant temperature (i.e., 850/900 °C) from the CLC, as detailed in Brachi et al. [9] and evaluated through Equation (2); and (ii) heat transfer from the cooling of exhaust flue gases, computed through Equations (3)–(5).

The electrolytic cell has only one exergetic input flux, related to the electrical power required, which has been estimated in the previous companion work [9]. Exergetic outflows are instead those related to the output mass flow of H_2 and O_2 , for which the physical and

chemical contributions were computed through Equations (3)–(6), considering an outlet pressure from the electrolyzer of 20 bar and a temperature of 50 °C.

Finally, concerning the methanation, incoming fluxes are due to the compressed CO₂ and H₂ streams. Outlet fluxes are instead related to the following: (i) sensible heat recovery from the stream of synthetic natural gas produced. Heat recovery occurs through heat exchanges located both between the consecutive reaction steps and at the end of the process; (ii) outflow stream of synthetic natural gas produced at the end of the methanation unit. Also for the methanation unit, Equations (3)–(6) apply for estimating the exergy fluxes.

3.2. Techno-Economic Analysis

This section presents the economic model assumptions and the main correlations utilized to estimate the economic characteristics of the integrated plant. Unless otherwise stated, capital cost equations indicate the purchase cost of the single equipment. Direct costs required for the equipment installation are set to 80% of the purchase cost [55]. For the indirect costs, engineering procurement and construction costs are set to 14% of the purchase plus direct costs (total direct plant costs), while contingency and owner's cost are set to 5% and 10%, respectively, of the total direct plant plus engineering costs [55]. The levelized cost of methane (LCOM) was evaluated according to Equation (9):

$$LCOM = \frac{TPC \cdot FCF + FOM}{W_{CH_4, year}} + VOM - REV \quad (9)$$

where TPC are the total plant costs, estimated by summing the purchase, direct, and indirect costs of all the equipment, FCF is the fixed charge factor, which annualizes the capital expenditures, FOM are the operating and maintenance (O&M) fixed costs, $W_{CH_4, year}$ is the yearly produced methane, VOM are the O&M specific variable costs, and REV is the specific revenue obtainable by selling the thermal/electrical energy produced within the integrated plant. The raw agro-industrial residues are considered to be available free of charge, and possible transportation costs are not considered.

FCF is evaluated through Equation (10):

$$FCF = \frac{PIR \cdot (1 + PIR)^\gamma}{(1 + PIR)^\gamma - 1} \quad (10)$$

where PIR is the project interest rate and γ the plant lifetime. FOM are assumed as a fixed percentage of the total plant costs.

The total plant costs include the main equipment of the integrated plant. The individual costs considered for each sub-section are detailed in the following:

- For the torrefaction, feeding (made of a front-end loader, hopper, and conveyor), drying, reaction, and storage units, plus the PV field required for the energy supply, and a heat exchanger/condenser to recover a share of the steam sensible/latent heat;
- For the chemical looping, reactors, cyclone, oxygen carrier, heat exchangers, and steam turbine;
- For the methanation, CO₂ compressor, reactors, heat exchangers/condensers, and catalyst;
- For the H₂ production, PV field, polymer electrolyte membrane (PEM) electrolyzer, and H₂ temporary storage vessel.

The cost equations of the different components are detailed in Table 1, and specific comments are given in the following when necessary. Capital expenses were updated to the year 2022 by means of the chemical engineering plant cost index (CEPCI). The reference cost of the PV field and PEM electrolyzer were instead not updated because it is believed that the increase due to inflation would be more than compensated by the technological advancements in these specific fields, and by an increased mass production of these devices. A money exchange rate of 0.9 € \$⁻¹ was considered.

Table 1. Capital cost equations for the main equipment of the integrated plant. Some equations rely on the six-tenths factor rule applied to a reference cost retrieved from the literature.

	Component (Scaling Parameter)	Correlation	Ref.	Year
PV	Photovoltaic field, installed (overall peak power, P_{PV} [kW])	$C_{PV}[\$] = 564 \cdot P_{PV}$	[56]	–
TORREFACTION	Feeding unit (biomass feed rate, F_B [t h ^{−1}])	$C_{TFU}[\text{€}] = 828,135 \cdot \left(\frac{F_B}{30}\right)^{0.6}$	[57]	2018
	Drying unit (biomass feed rate, F_B [t h ^{−1}])	$C_{TDU}[\$] = 3,813,728 \cdot \left(\frac{F_B}{166.7}\right)^{0.75}$	[58]	2002
	Reactor (dry biomass feed rate, F_{DB} [t h ^{−1}])	$C_{TR}[\$] = 8,363,200 \cdot \left(\frac{F_{DB}}{726}\right)^{0.7}$	[58]	2020
	Storage (torrefied biomass mass, M_{TBM} [t])	$C_{TBS}[\text{€}] = 31.29 \cdot \left(\frac{M_{TBM}}{1}\right)^{0.6}$	[59]	2014
	Condenser (heat transfer surface, A_C [m ²])	$C_C[\$] = 8500 + 490 \cdot A_C^{0.85}$	[60,61]	2012
CHEMICAL LOOPING	Reactors and oxygen carrier (input power, $P_{CLC, input}$ [MW])	$C_{CLCR}[\text{€}] = 950,000 \cdot \left(\frac{P_{CLC, input}}{200.67}\right)^{0.6}$	[62]	2018
	Cyclones (input power, $P_{CLC, input}$ [MW])	$C_{CLCC}[\text{€}] = 17,000 \cdot \left(\frac{P_{CLC, input}}{200.67}\right)^{0.6}$	[62]	2018
	High-temperature heat exchangers (input power, $P_{CLC, input}$ [MW])	$C_{CLCHX}[\text{€}] = 60,000 \cdot \left(\frac{P_{CLC, input}}{200.67}\right)^{0.6}$	[62]	2018
	Oxygen carrier (input power, $P_{CLC, input}$ [MW])	$C_{CLCOC}[\text{€}] = 980,000 \cdot \left(\frac{P_{CLC, input}}{200.67}\right)^{0.6}$	[62]	2018
	Steam turbine (output power, P_{ST} [kW])	$C_{ST}[\$] = 3744.3 \cdot P_{ST}^{0.7} - 61.3 \cdot P_{ST}^{0.95}$	[61,63]	2016
METHANATION	CO ₂ dryer and compressor, installed (compressor power, P_C [kW])	$C_C[\$] = 1,4800,000 \cdot \left(\frac{P_C}{13000}\right)^{0.67}$	[64]	2002
	Packed bed reactors	see detailed methodology in ref.	[31]	–
	Catalyst	see detailed methodology in ref.	[31]	–
	Mid-temperature heat exchangers (heat transfer surface, A_{HX} [m ²])	$C_{HX}[\$] = 2290 \cdot (A_{HX})^{0.6}$	[61,63]	2012
	Condenser (heat transfer surface, A_C [m ²])	$C_C[\$] = 8500 + 490 \cdot A_C^{0.85}$	[60]	2012
H ₂	Electrolyzer (peak power from PV field, P_E [kW])	$C_E[\text{€}] = 375 \cdot P_E$	[65,66]	–
	Temporary storage vessel, installed (storage volume, V_{H_2} [Nm ³])	$C_{HST}[\text{€}] = 33 \cdot V_{H_2}$	[67,68]	–

For the PV field, fixed modules with optimized slope and orientation and made of crystalline silicon have been considered, rather than the two-axis tracking modules originally proposed in the companion paper [9], because of their lower investment cost. To estimate the peak power (P_{PV}) required for the capital cost estimation, the methodology described in [31] has been applied, using monthly averaged solar data retrieved from the PVGIS database [69,70]. The energy required is that for either the torrefaction process, or the H₂ production [9]. The efficiency of the PV modules (η_{PV}) is set to 20%, and overall system losses to 14% (η_F). Solar data for the four different locations under the given assumptions are depicted in Figure 2.

For the torrefaction unit, since the system works at different throughputs during the year [9], the maximum biomass feed rate over the year was considered in order to estimate the capital cost. This was computed by taking as reference the month of the year with the highest irradiation and scaling the nominal biomass feed rate [9] according to the ratio between the actual energy available during a typical day in the considered month and the nominal daily energy required. The mass of torrefied biomass to store (M_{TBM}) was computed by assessing the mass of torrefied biomass that, during each typical day, the

torrefaction system does not produce or overproduce (because of the lack or excess of input solar energy, respectively) vs. that required for the nominal operation of the CLC plant. The detailed methodology is explained in Tregambi et al. [71], where it is used to size the calcium oxide/carbonate storage silos for a calcium looping process. Heat exchangers/condensers were sized by considering the power recoverable from the cooling and condensation of the steam produced during biomass drying (P_{TD}), and estimating an average heat transfer surface assuming an overall heat transfer coefficient of $350 \text{ W m}^{-2} \text{ K}^{-1}$ and a mean temperature difference for the heat transfer operation of $25 \text{ }^{\circ}\text{C}$. The outlet temperature from heat exchangers/condenser is set at $50 \text{ }^{\circ}\text{C}$, as for the previous exergetic analysis.

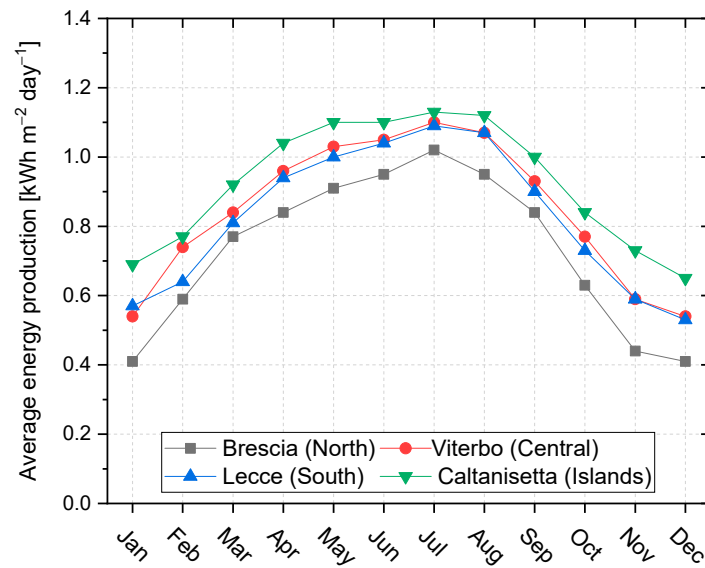


Figure 2. Solar data for the four different locations considered.

For the CLC unit, the design of the steam turbine was performed by also including the heat recovered in the previous torrefaction and subsequent methanation sections, besides that produced in the CLC unit itself. The output power of the steam turbine (P_{ST}) was evaluated by multiplying the input exergy flux for an electric conversion efficiency (η_{el}), set equal to 0.40.

For the methanation reactors, the power of the CO_2 compressor was estimated considering a three-stage isentropic compression with intercooling, while the heat transfer surface was estimated by considering the overall power produced by the methanation section (P_M), and the same parameters used for the heat exchangers/condensers of the torrefaction unit.

Finally, for the H_2 production unit, the peak power delivered to the electrolyzer was evaluated by reducing the PV field power because of the system losses [31], while for the H_2 temporary storage vessel, the same approach used for the biomass storage tank was implemented [31,71].

The O&M variable costs account for replacing the oxygen carrier, catalyst, and electrolyzer, that feature a shorter lifetime than the plant one:

$$VOM = \frac{C_{cat}}{lt_{cat} \cdot W_{CH_{4,y}}} \cdot \left(\frac{\gamma - lt_{cat}}{\gamma} \right) + \frac{C_{OC}}{lt_{OC} \cdot W_{CH_{4,y}}} \cdot \left(\frac{\gamma - lt_{OC}}{\gamma} \right) + \frac{C_E}{lt_E \cdot W_{CH_{4,y}}} \cdot \left(\frac{\gamma - lt_E}{\gamma} \right) \quad (11)$$

where C_{cat} , C_{OC} , and C_E are the costs of catalyst, oxygen carrier, and electrolyzer, respectively, and lt_{cat} , lt_{OC} , and lt_E their expected lifetime. For their replacement, direct costs are halved, whereas indirect costs do not apply.

Finally, the revenue accounts for the electrical energy produced from the steam turbine, which includes the heat recovered from the torrefaction plus CLC and methanation unit, reduced by that required for the operation of the CO₂ compressor:

$$REV = \frac{(P_{ST} - P_C) \cdot 8760 \cdot h \cdot CF \cdot PoE}{W_{CH_4,y}} \quad (12)$$

where C_F is the capacity factor and PoE is the price of energy. In this work, a capacity factor of 0.85 was considered, whereas the price of energy was varied within 0.10–0.14 € kWh^{−1}. Table 2 summarizes some of the main input parameters required for the economic analysis.

Table 2. Main process and economic parameters.

Symbol	Description	Value	Dimension
CF	Capacity factor of the integrated plant	0.85	–
lt_{cat}	Catalyst lifetime	5	y
P_C	CO ₂ compressor power	1.2–1.6	kW
lt_{EC}	Electrolyzer lifetime	10	y
PoE	Energy price	0.10; 0.12 *; 0.14	€ kWh ^{−1}
FOM	O&M fixed costs	1	%
U_{HXs}	Heat transfer coefficient in heat exchangers	350	W m ^{−2} K ^{−1}
$P_{CLC,input}$	Input power of CLC (based on HHV of torrefied biomasses)	62.4–81.5	kW
γ	Integrated plant lifetime	25	y
ΔT_{HXs}	Mean temperature difference for heat exchangers	25	°C
lt_{OC}	Oxygen carrier lifetime	5	y
P_{PV}	Peak power of photovoltaic field	12–370 (torrefaction) 919–1463 (H ₂ prod.)	kW _p
P_{ST}	Power output of steam turbine	21.3–32.5 kW	kW _e
P_M	Power recovered from methanation	25.1–30.6	kW _{th}
P_{TD}	Power recovered from steam cooling and condensation after biomass torrefaction	1.5–47.0	kW _{th}
PIR	Project interest rate	4.75; 6.75 *; 8.75	% y ^{−1}
η_{PV}	PV modules efficiency	0.20	–
η_F	System losses of overall PV field	0.14	–
η_{el}	Thermal-to-electric conversion efficiency	0.40	–
$W_{CH_4,y}$	Yearly produced methane	55.4–69.8	t y ^{−1}

* Base case.

4. Results

4.1. Energetic and Exergetic Analysis

Table 3 reports the absolute values of energy, exergy, and mass fluxes for each process unit of the integrated plant, with reference to SBP. Data related to the other biomasses are reported in the Supplementary material (Tables S1–S3). Differences between energy and exergy flow reveal the quality of each flux. Typically, exergy fluxes are lower than corresponding energy ones; only in the case of chemical exergy can this trend be sometimes reverted, because the ratio between the chemical exergy and the heating value of a fuel can be greater than unity [48]. In particular, this only happens with reference to the input of the torrefaction step in Table 3. Differences in output energy and exergy for torrefaction are mainly due to the energy recoverable at low temperatures (50–300 °C) from the torrefaction process, resulting in an ideal work that is lower than its corresponding energy output. Similar considerations apply to the CLC unit, where however heat is released at higher temperatures (850/900 °C from the reactors and 50–900 °C from heat exchangers). For the electrolysis unit, no differences are foreseen between energy and exergy input flows since the power flow is electrical. On the other hand, the electrolysis exergy outflow is lower than the energy output because the specific chemical exergy of H₂ is lower than its

corresponding specific enthalpy of combustion. The methanation input is slightly higher than the electrolysis output because of the energy requirements for CO₂ compression. The methanation output includes both the heat fluxes and the heating value of the synthetic natural gas, whose average composition is 92%_{db} CH₄, 6%_{db} H₂, 2%_{db} CO₂. The output heat flow from the methanator is at an intermediate temperature, within 50–472 °C. The quality of this heat and the differences between the specific chemical exergy and specific heating value of the mixture explain the lower exergy output flow for the methanation. Analysis of Table 3 also allows for an interesting consideration of the water needs of the process. Ideally, if the methanation gas is oxidized locally, the water required for the electrolysis is already available. However, the synthetic natural gas stream produced could be transported and exploited elsewhere. Therefore, the water demand of the electrolysis unit must be supplied from the environment. Local sources could be unavailable in regions where the amount of water in natural reserves, such as rivers, lakes or aquifers is limited or even insufficient for agriculture and human settlements. This critical problem, tied to the process, might be overcome, and, in some cases, completely solved, if the water output from the other units (i.e., torrefaction and CLC) is recovered and recycled. In particular, as detailed in Table 3, for SBP the output water from torrefaction is already more than enough to cover the water requirements of the electrolysis unit, due to the high moisture content of SBP. Similar considerations apply to GM but also including the water recovered from the methanation process (Table S2). For OS and HH, due to the lower moisture content of the raw biomasses, the output water is less than that required by the electrolysis unit (Tables S1 and S3, respectively). However, this recycling can reduce the net water demand of the process by approximately 65% for both biomasses.

Table 3. Energy, mass, and exergy flows for each unit of the integrated process treating SBP.

Process Unit	Flow Direction	Energy Flow [kW]	Exergy Flow [kW]	Flow Composition	Mass Flow [kg h ^{−1}]
Torrefaction	In	178.9	185.5	Wet Biomass	93.4
	Out (with; w/o torgas)	167.4; 110.5	141.1; 83.6	Torrefied Biomass	10.0
				Water (L)	68.3
				Torgas	14.0
CLC	In	63.6	65.5	Torrefied Biomass	10.0
	Out	63.4	41.3	Water (L)	2.5
				CO ₂	21.5
Electrolysis	In	182.6	182.6	Water (L)	35.2
	Out	155.3	133.6	O ₂	31.3
				H ₂	3.9
Methanation	In	156.5	134.8	CO ₂	21.5
	Out	149.3	130.7	H ₂	3.9
				Methanation gas	8.5
				Water (L)	17.4

Table 4 shows instead the exergetic efficiency with reference to either the individual plant subsection, or the overall integrated process. Data suggest that varying the fed biomass leads to a sensible change in the individual torrefaction efficiency. Without considering the torgas contribution, it ranged indeed within 34–68% for the different biomasses. The higher value was obtained for OS, followed by GM, SBP, and HH. Such variability is probably due to differences in the moisture content of raw biomasses and in the mass yield of the torrefaction process. Valorization of torgas, if technically feasible for the considered biomasses, would make the torrefaction efficiency increase by about 60% for HH and about 30% for the other biomasses. Such results might also suggest that a too-high torrefaction temperature was applied for HH. The exergetic efficiency of CLC showed instead only minor differences. Efficiency values range indeed within 61–67%, and variations are mainly

due to differences in the biomasses' composition and heating values. Concerning the electrolysis, the exergetic efficiency is independent of the biomass, as the value obtained arises from the model considered for the electrolyzer [9], which is assumed to behave equally in all the analyzed cases. The efficiencies of the methanation unit also show mostly consistent levels of performance throughout, as expected. In fact, at this stage, the methanation reactors simply process a stream of CO_2 and H_2 , and any information regarding the starting biomass is lost. The reactors achieve a conversion degree of 98%, and the composition and temperature of inlet and outlet gases are the same regardless of the biomass.

Table 4. Global and individual units' efficiencies as inferred from the exergetic analysis.

Efficiency	SBP	OS	GM	HH
Torrefaction w/o torgas	45%	68%	57%	34%
Torrefaction with torgas	76%	95%	86%	94%
CLC	63%	61%	67%	65%
Electrolysis	73%	73%	73%	73%
Methanation	97%	96%	95%	96%
Global w/o torgas	51%	60%	57%	45%
Global with torgas	67%	70%	69%	77%

Altogether, global exergetic efficiencies also show a fair variability with respect to the biomass fed. OS stands out as the most efficient biomass option, with an exergetic efficiency of 60%. GM follows the chart (57%), followed by SBP (51%) and HH (45%). When considering also torgas valorization, HH becomes instead the preferable option (77%), followed by OS (70%), GM (69%), and SBP (67%). In the scenario without torgas valorization, global efficiencies are less uniform when varying the biomass type, which is due to a corresponding higher variability in the torrefaction efficiency. In general, the efficiency of the torrefaction unit shows the highest degree of variability, a phenomenon that will be further elucidated after the introduction of the Sankey diagrams. Moreover, the introduction of Sankey diagrams makes it easier to understand where the irreversibilities of the process are nested.

Figure 3 illustrates the Sankey diagram depicting the exergy flow for all the biomasses considered in this study. Dimensionless quantities are presented, and the total input exergy is indicated in the heading of each diagram. The proposed process layout is distinguished into two parallel processes, linked by the CO_2 stream from the CLC to the methanation unit. In the Sankey diagrams, the stream of compressed CO_2 is not linked to the CLC because it is assumed that the exergy for its compression comes from an inter-refrigerated three-stage compressor, which has not been reported in the diagrams to enhance clarity. A comparison of the presented diagrams reveals that OS is characterized by the lowest exergetic demand in terms of electricity for biomass drying, thanks to its low moisture content. SBP is at the opposite extreme, with the highest exergetic demand dictated by the highest moisture content. Since low-temperature heating processes using electric power are highly degrading in terms of energy quality, when a higher amount of electrical energy is involved, the irreversibility rate is higher. In the case of SBP (Figure 3a), the irreversibility rate is 12% of the total input exergy, of which 15.3% is electric exergy. The irreversibility decreases to 1.6% when OS is treated (Figure 3b), but in this case, only 0.9% of the total input is in the form of electric exergy. The amount of exergy in the torgas stream is directly related to the energy yield of the torrefied solid, given by the ratio between the energy content of the torrefied solid with respect to that of the parent biomass. The lower is the energy yield of the torrefied solid (given by the product between the LHV of the torrefied biomass and its mass yield), and the higher will be the energy and exergy of the torgas stream. This difference strands out clearly when comparing, for example, the Sankey diagrams relative to OS and HH (Figure 3b,d). In fact, the amount of exergy held in torgas for HH is 31.2%, with a mass yield in torrefied solid of 24%. In contrast, torgas of OS contains only 9.4% of the input exergy, and the mass yield of the torrefied solid is remarkably higher, equal to

62%. With these results in mind, it might be useful to investigate in future works the effect of torrefaction conditions: lower temperatures could bring to higher yields in torrefied solid product and correspondingly lower amounts of torgas [54], allowing higher process efficiency even without torgas valorization, which is not always technically feasible. The differences between the Sankey diagrams show that the proposed process configuration achieves the ability to mitigate the effect of variations in biomass chemical and physical properties, aligning with the core objective of utilizing agro-industrial residues, increasing their competitiveness with other types of fuels, and enhancing the value of waste biomasses. It remains true that better biomass quality translates into improved operational efficiency. The biomasses characterized by lower moisture content, higher carbon and hydrogen levels, and reduced nitrogen content, resulted in higher global exergetic efficiencies.

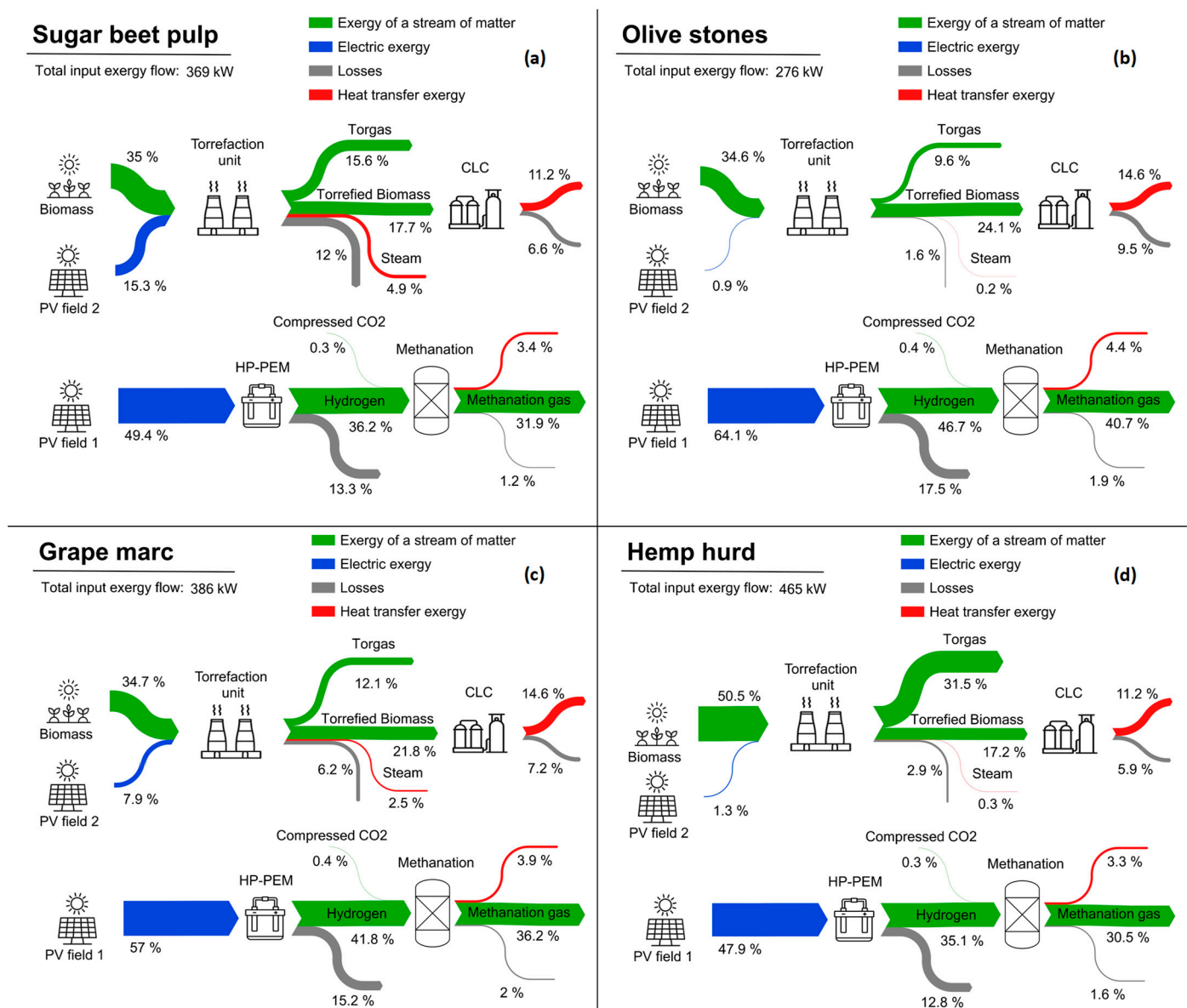


Figure 3. Sankey diagrams illustrating the exergy flow for the four biomasses. The exergetic flow related to torgas should be considered as losses if it is not valorized.

Finally, Figure 4 shows the round-trip efficiency of the integrated process as a function of the electric conversion efficiency. This is the maximum achievable round-trip efficiency if considering both thermal power released during the methanation, and conversion of methane into electricity, through a power plant. The electric conversion efficiency has been

varied within 0.3–0.55 [9,49]. Correspondingly, it is possible to observe that the round-trip efficiency ranged within 20–38%. The influence of biomass is barely detectable, as all the values are mostly overlapped. This index indeed focuses on the methanation process, mostly neglecting the source of the concentrated CO₂ stream.

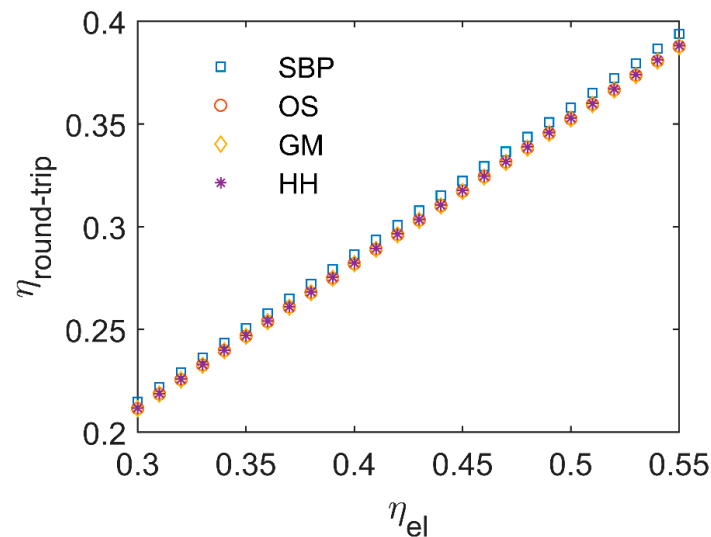


Figure 4. Round-trip efficiency as a function of the electric conversion efficiency.

4.2. Techno-Economic Analysis

Figure 5 shows the total plant costs, and their breakdown in the main plant subsections, for the four biomasses in a fixed site location (base case scenario). It is observed that, in all the cases, the H₂ temporary storage vessel is the most expensive equipment (39–44% of the total plant costs), followed by the electrolysis system (21–23% of total plant costs), and the PV field for the H₂ production (18–20% of the total plant costs). The methanation, chemical looping, and torrefaction units each account for about 3–6% of the total plant costs. The cost of the PV field for the torrefaction process is almost negligible for OS and HH, which feature a very low moisture content (<15%_{wt}), but it accounts for 3–6% of the total plant costs for SBP and GM, which feature instead a much higher water content (>50%_{wt}), in agreement with the exergetic analysis discussed before. Altogether, the total plant costs range within 3.3–4.2 M€ for the considered site location: lower values are obtained for the biomasses with a lower content of fixed C and moisture, and vice versa. Indeed, the fixed C fraction determines the amount of CO₂ produced, hence the H₂ required, whose production directly affects the investment cost of the PV field, electrolyzer, and methanation unit. The moisture content, instead, affects the cost of the torrefaction unit, and of the PV field as well, because of the energy required by the torrefaction process. It should, however, be underlined that for biomasses that produce more CO₂, despite larger investment costs, a higher amount of synthetic natural gas will be produced eventually.

Figure 6 shows the same data as Figure 5 but varying the site location for the same biomass (SBP). It is observed that the total plant costs decrease from about 4.5 M€ to about 3.3 M€ when moving from the north of Italy (Brescia) to the south of Italy (Caltanissetta). The trend clearly arises from the higher solar irradiation that characterizes the south of Italy, as inferable from the analysis of Figure 2. In particular, both the higher solar irradiation and its slightly lower fluctuation over the different months of the year, strongly reduce the cost of the PV field, electrolyzer, and H₂ temporary storage tank.

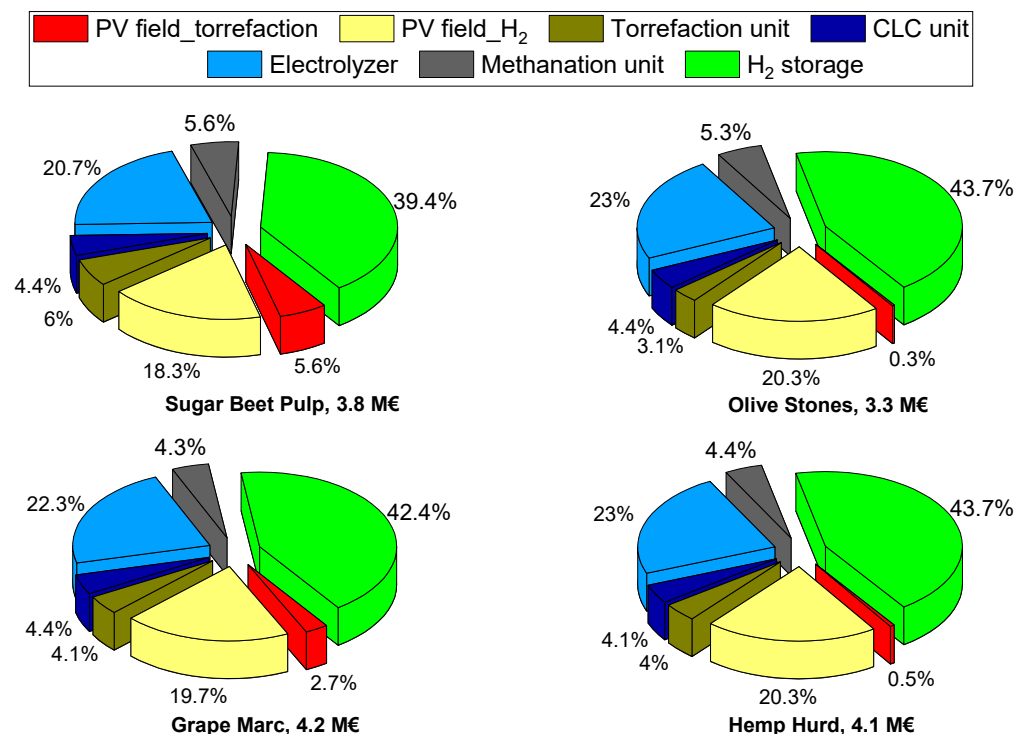


Figure 5. Total plant cost (and its breakdown) for the four biomasses, Viterbo.

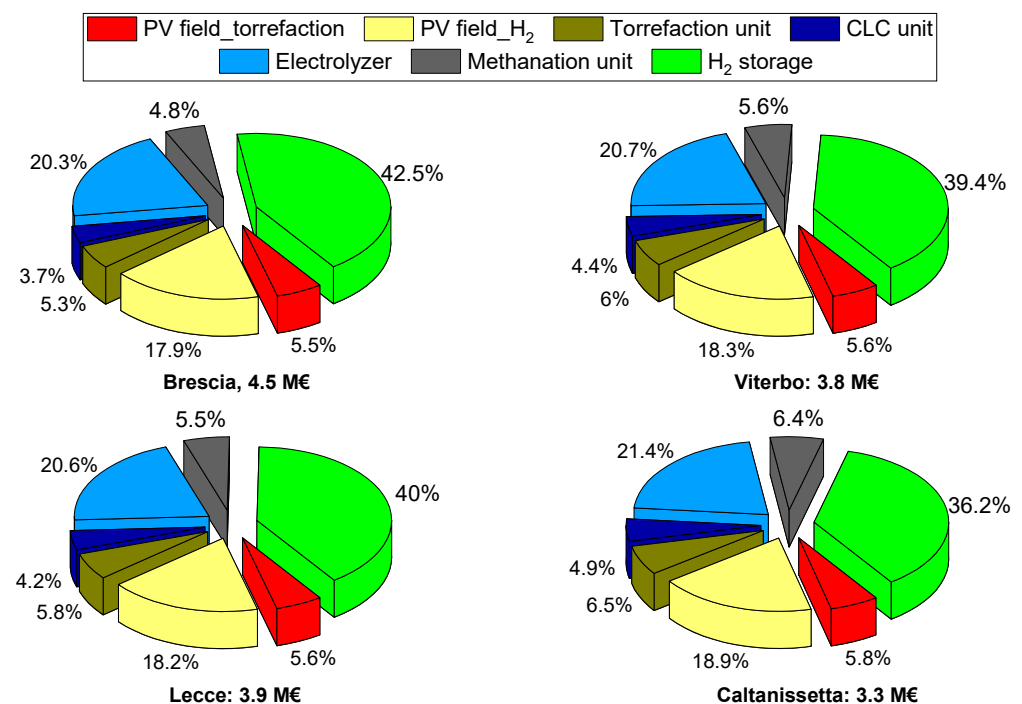


Figure 6. Total plant costs (and related breakdown) for the four site locations, sugar beet pulp.

Figures S1–S3 in the Supplementary Material show the total plant costs, and related breakdown, also for the other combinations of biomasses/site locations. Altogether, the total plant costs range within 2.9–3.9 M€ for olive stones, 3.7–5.0 M€ for grape marc, and 3.6–4.9 M€ for hemp hurd. Again, lower costs are obtained for the south of Italy and higher for the north, as already observed in Figure 6.

Figure 7 shows instead the levelized cost of methane for all the different cases investigated. Error bars refer to a sensitivity analysis on PIR and PoE values with respect to the base-case (see Table 2). From the analysis of the chart, the following can be inferred:

1. Regardless of the site location, the use of SBP returns the highest levelized costs of methane. Values reduce by about $0.5\text{--}0.6 \text{ € kg}_{\text{CH}_4}^{-1}$ when considering the other biomasses;
2. Following the trend already observed in Figure 6, the levelized cost of methane decreases when the site location shifts towards the south of Italy. It decreases by about $1.0\text{--}1.2 \text{ € kg}_{\text{CH}_4}^{-1}$ when moving from the north to the center and early south of Italy, and by $1.9 \text{ € kg}_{\text{CH}_4}^{-1}$ when moving to the deep south;
3. By varying PIR and PoE values, the levelized cost of methane varies by about $0.9\text{--}1.3 \text{ € kg}_{\text{CH}_4}^{-1}$ depending on the specific case. Lower values are obtained when the PIR is minimum and PoE is maximum, and vice versa;
4. In any case, in all the considered scenarios, the levelized cost of methane is out of the market, as it ranges within $4.3\text{--}8.9 \text{ € kg}_{\text{CH}_4}^{-1}$. These values mostly originate from the capital expenditures and O&M fixed costs (i.e., first addend of Equation (9)), whereas the O&M variable costs account for about $0.54\text{--}0.68 \text{ € kg}_{\text{CH}_4}^{-1}$, and the revenue determines a saving of about $0.23\text{--}0.50 \text{ € kg}_{\text{CH}_4}^{-1}$ depending on the value of PoE considered.

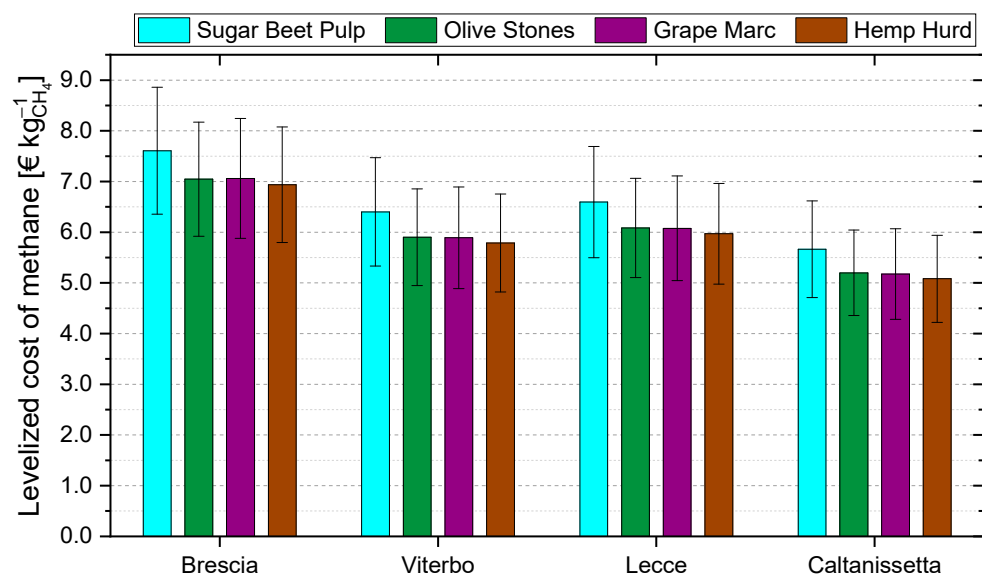


Figure 7. Levelized cost of methane for the different cases. Error bars refer to the sensitivity analysis.

Through a combined analysis of Figures 5 and 7, it appears that a significant contributor to the cost of methane is represented by the H_2 temporary storage vessel, which is required to balance the fluctuations of solar radiation over the different months of the year. Therefore, strategies to avoid or reduce the temporary H_2 storage should be considered. Some of them could be:

- Store CO_2 and operate the methanation reactors at different throughputs over the year according to the H_2 availability;
- Sell a share of the energy produced during the sunniest months and buy it back during the months with less solar irradiation;
- Combine the utilization of different renewable energies, such as solar and wind power;
- Oversize the PV field and sell the extra energy to the market.

A detailed simulation of the different possibilities is beyond the scope of the present work. However, it is interesting to assess what the levelized cost of methane would be if overcoming the problem of the temporary H_2 storage. This can be appreciated in Figure 8,

which shows the same data as Figure 7 but neglecting the investment cost of the temporary H_2 storage vessel. These values represent the lower benchmark for the levelized cost of methane for the process investigated in this study.

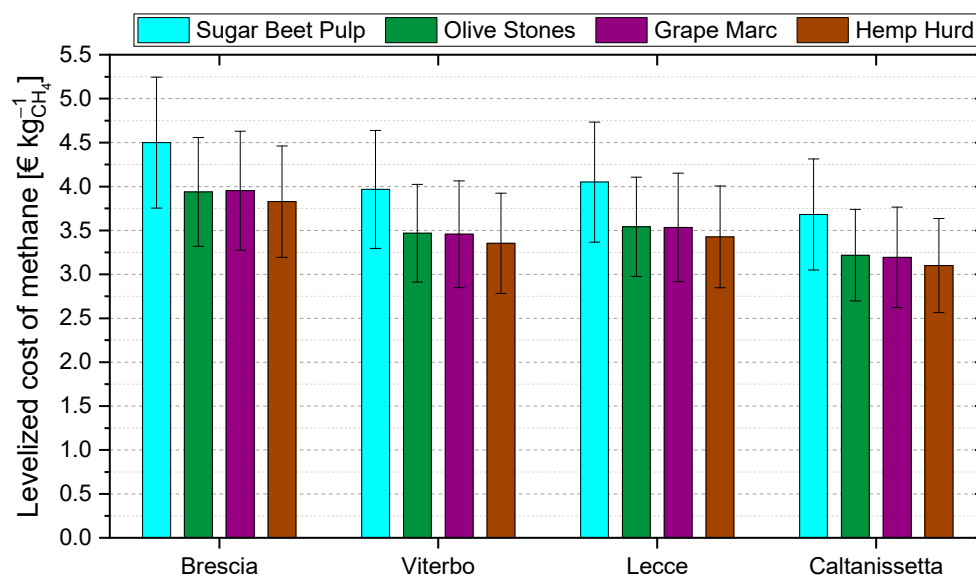


Figure 8. Levelized cost of methane for the different cases neglecting the H_2 temporary storage vessel. Error bars refer to the sensitivity analysis.

Data in Figure 8 reveal a consistent decrease in the values of LCOM, which range now within 2.6–5.3 $\text{€ kg}_{CH_4}^{-1}$ depending on the option considered. The removal of the temporary H_2 storage tank brought indeed to a reduction of approximately 3.1 $\text{€ kg}_{CH_4}^{-1}$ for the north of Italy, 2.4–2.5 $\text{€ kg}_{CH_4}^{-1}$ for the center and early south of Italy, and 2.0 $\text{€ kg}_{CH_4}^{-1}$ for the deep south. The decrease is slightly lower for the south of Italy because of the lower fluctuation of solar radiation over the year in these regions, as discussed above. Altogether, when considering the north of Italy as site location, the levelized cost of methane is still high, ranging within 3.2–5.3 $\text{€ kg}_{CH_4}^{-1}$ depending on the biomass considered and on the values of *PoE* and *PIR* selected. However, when moving to the southern part of Italy, LCOM values become much more interesting. For Caltanissetta, they range indeed within 2.6–4.3 $\text{€ kg}_{CH_4}^{-1}$ depending on the option considered. HH and GM are the biomasses that return the lower values of levelized cost of methane, followed by OS and SBP. Although these values are still slightly out of the market, the introduction of a carbon tax or the availability of government incentives for the purchase of the PV field and/or electrolyzer might push them toward lower values, making the technology competitive towards current natural gas prices.

5. Conclusions

The energetic, exergetic and techno-economic performances of a bioenergy with carbon capture and utilization process have been investigated. The integrated process is based on chemical looping combustion of solar-torrefied agro-industrial residues, followed by methanation of the concentrated CO_2 stream with green H_2 . Four different residues have been considered—sugar beet pulp, olive stones, grape marc, and hemp hurd—along with four different Italian site locations for the PV field location.

Depending on the considered biomass, the integrated plant processes about 18–93 kg h^{-1} of raw biomass and produces 55–70 t y^{-1} of synthetic methane. The highest overall exergetic efficiencies are obtained for olive stones (60%) and grape marc (57%), followed by sugar beet pulp (51%) and hemp hurd (45%). Such efficiencies increase up to 67–77% when considering torgas valorization. For biomasses featuring a high moisture content (sugar beet pulp and grape marc), a relevant share or even all the water needed by the

electrolyzer might be obtained from biomass drying, allowing a more sustainable use of water resources. A round-trip electric efficiency of about 28% has been computed for the methanation process, assuming an electric conversion efficiency of 40%.

The economic analysis revealed total plant costs within 2.9–5.0 M€. The use of grape marc and hemp hurd requires higher investment costs, but with a correspondingly higher synthetic methane production. The H₂ temporary storage vessel is the most expensive component (39–44% of the total plant costs), followed by the electrolysis system (21–23%) and the PV field for the H₂ production (18–20%). The methanation, chemical looping, and torrefaction units each account for about 3–6% of the total plant costs. The levelized cost of methane is within 4.3–8.9 € kg_{CH₄}^{−1}, with lower values for the south of Italy and when using hemp hurd or olive stones as input biomass. By removing the H₂ temporary storage vessel, the costs decrease to 2.6–4.3 € kg_{CH₄}^{−1} for the south of Italy. Despite these values still being out of the market, the introduction of a carbon tax or government incentives might further reduce them, making the technology compete with current natural gas prices.

Supplementary Materials: The supporting information can be downloaded at <https://www.mdpi.com/article/10.3390/en17112690/s1>. Table S1. Energy, mass, and exergy flows for each unit of the integrated process treating OS. Table S2. Energy, mass, and exergy flows for each unit of the integrated process treating GM. Table S3. Energy, mass, and exergy flows for each unit of the integrated process treating HH. Figure S1. Total plant cost (and its breakdown) for the four site locations, olive stones. Figure S2. Total plant cost (and its breakdown) for the four site locations, grape marc. Figure S3. Total plant cost (and its breakdown) for the four site locations, hemp hurd.

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List of Abbreviations

BECCS	bioenergy with carbon capture and storage
BECCU	bioenergy with carbon capture and utilization
CEPCI	chemical engineering plant cost index
CLC	chemical looping combustion
GM	grape marc
HH	hemp hurd
IEA	international energy agency
LCOM	levelized cost of methane
O&M	operating and maintenance
OS	olive stones
PEM	polymer electrolyte membrane
PV	photovoltaic
SBP	sugar beet pulp

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