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# Upgrading to biomethane and power production in SOFC-based cogeneration system: An exergo-economic comparison of biogas conversion alternatives


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# Nomenclature

**Symbols**

| | |
|---|---|
| $\overline{\overline{A_c}}$ | cost matrix |
| $\bar{\bar{A}}$ | incidence matrix |
| $C_p^0$ | bathe se cost of component ($) |
| $\dot{m}$ | mass flow (kg/s) |
| μ | specific chemical potential (kJ/kg) |
| C | cost of stream (€/s) |
| C | specific cost of the stream (€/kJ) |
| $C_{bm}$ | bare erected cost or BEC ($) |
| e | specific exergy (kJ/kg) |
| E | total exergy flow (kW) |
| E* | exergy cost (kW) |
| f | capacity factor |
| $F_{bm}$ | bare erected factor |
| $f_{ex}$ | exergo-economic factor |
| $F_m$ | material factor |
| $F_p$ | pressure factor |
| G | Gibbs free energy (kJ) |
| h | specific enthalpy (kJ/kg) |
| i | discount rate (%) |
| I | exergy lost (kW) |
| k* | specific exergy cost |
| n | lifetime of the plant (years) |
| p | pressure (bar or Pa) |
| Q | heat flux (kW) |
| R | elastic module of gas (kJ/kg K) |
| r | relative cost difference |
| s | specific entropy (kJ/kg K) |
| T | temperature (K or °C) |
| W | work (kW) |
| x | mass fraction |
| y | molar fraction |
| $Z_c$ | cost of component (€/s) |
| ε | exergy efficiency (II principle) |
| η | energy efficiency (I principle) |
| λ | exergy factor |
| Ψ | exergy destruction rate |

**Acronyms**

| | |
|---|---|
| CEPCI | chemical engineering plant index |
| CRF | capital recovery factor |
| DBE | distance from break-even |
| EPC | engineering and procurement cost |
| IR | investment return |
| LHV | lower heating value (kJ/kg) |
| NPV | net present value (€) |
| PBT | pay-back time (years) |
| PS | public subsidies |
| TASC | total as-spent cost or capital (€) |
| TOC | total overnight cost (€) |
| TPC | total plant cost (€) |

# Abstract

Biogas is experiencing a period of rapid growth in production in the past decades and another developer is predicted for the next years, the biogas is considered an important source of energy to achieve the 20-20 European goals. The possible applications of the produced biogas are the direct use in specific burner to thermal usage or the direct use in CHP system based on internal combustion engine, the alternative is the upgrading of the biomethane to obtain biomethane with high concentration of methane to grid injection or transport fuel, all these applications can take advantage of public incentive recognized for renewable energy source in European country. With the increase of biogas production all over the world, the technologies to valorize this renewable energy source have increased their importance. This paper aims to compare two possible use of biogas: upgrading with pressurized water scrubbing to obtain biomethane for grid injection or the direct use in advance CHP plant based on SOFC. The analysis starts from a nominal flow of biogas feeding both plants, the simulations of both plants are developed in Aspen-Plus® to obtain the thermodynamic condition of each stream of mass, work and heat. The results of Aspen-Plus® simulations are used to investigate the energy and exergy performances of both plants, based on the economic cost of components calculated with the cost function, the exergo-economic analysis determines the cost of each stream especially the production cost of principal product of systems and eventually the production cost of secondary products. The economic analysis investigates the economic performances of the plants starting from the investment cost, cost of operating and maintenance and the cost of fuels, the earnings of the plants derived from the selling of products of systems and the public subsidies in the Italian market. The sensitivity analysis performed over the two plant aims to evaluate the change of the production cost of biomethane and electricity with the variation of some project parameters like the cost of raw biogas, the investment cost of SOFC module and the size of systems and the consequential change of economic performances. The last section of this work regards the comparison of the exergetic and economic performance of both plants to determine which is the better technological solution to exploit the biogas coming from anaerobic digestion of waste.

# 1 Introduction

Biogas has an important role in the scenario of the renewable energy sources because it is experiencing a period of rapid growth of their production and development of technology for its use. It is estimated that the European target about of 20% of renewable energy in 2020 will be covered partially by bioenergy, at least 25% of which will be biogas, the top five biogas producing country in Europe are Germany, UK, Italy, France and Netherland [1], [2]. In addition, the global power generation capacity will increase from 14.5 GW in 2012 to 29.5 GW in 2022 [1], [2].

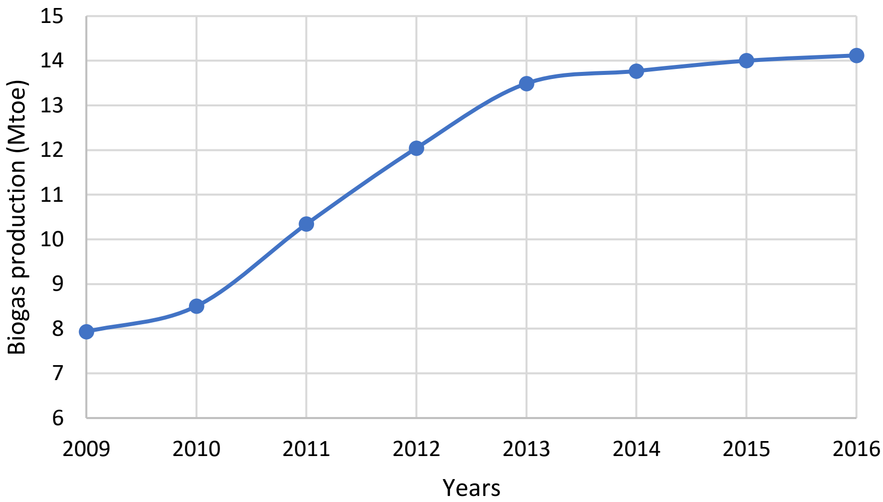


*Figure 1: Biogas production in Europe* [3]

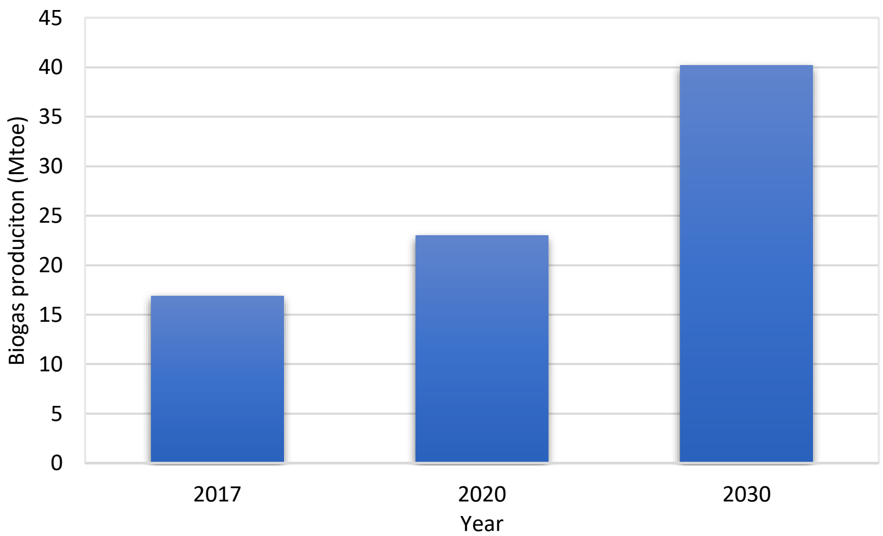


*Figure 2: Predicted biogas production in Europe* [1]

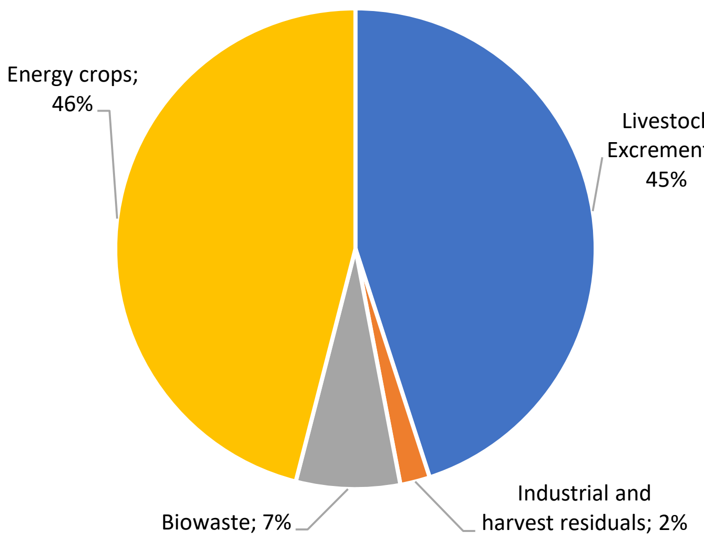


*Figure 3: Typical feedstock in biogas plant* [3]

Biogas is produced by microbial anaerobic digestion of the organic mass through four processes: hydrolysis, acidogenesis, acetogenesis, and methanogenesis. It is composed mainly of methane ($CH_4$) and carbon dioxide ($CO_2$), in different concentration depending on the source production, and other trace gas [4]. Some examples of biogas composition are reported in *Table 1.*

| Components | Municipal waste | Wastewater | Agricultural/animal waste | Food industry waste | Landfill |
|---|---|---|---|---|---|
| $CH_4$ (%vol.) | 50-60 | 61-65 | 60-75 | 68 | 45-70 |
| $CO_2$ (%vol.) | 34-38 | 36-38 | 19-33 | 26 | 35-40 |
| $N_2$ (%vol.) | 0-5 | <1 | <1 | - | <3 |
| $O_2$ (%vol.) | - | <0.5 | <0.5 | - | <0.2 |
| $H_2$ (%vol.) | - | - | - | - | 0-5 |
| CO (%vol.) | - | - | - | - | 0-3 |
| $H_2S$ (ppm) | 70-650 | 700-2800 | 2100-7000 | 2800 | 10-200 |

*Table 1: biogas composition* [5]

Biogas contains also other trace components like siloxane, moisture and chlorine that must be removed before the use, the Lower Heating Value of the produced gas is strongly dependent from the concentration of methane in the mixture. The possible uses of the biogas are: direct use to produce thermal power in the specific burner, direct use in CHP generation by the internal combustion engine or micro-turbine, injection in the natural gas grid and the use as a transport fuel.

| Type | Municipal solid waste | Food industry waste | Agricultural/animal waste | Sewage sludge |
|---|---|---|---|---|
| Biogas yield per ton of fresh matter ($m^3$) | 100 | 110 | 50 | 47 |

*Table 2: Biogas yield from different substrates* [3]

## 1.1 Upgrade to biomethane

The raw biogas is not suitable to be used as a substituted natural gas or biofuel for vehicles due to the low concentration of methane respect to the natural gas and the presence of contaminants compounds. Firstly the biogas needs to be cleaned from trace gas, this process is usually made with a cleaning bed composed by metal ion impregnated activated carbon [2], then the biogas undergoes a process called "biogas upgrading to biomethane" to obtain a product with a concentration of $CH_4$ similar to natural gas. There are different technologies of upgrading with different performance in terms of purity of methane, losses and efficiency [1], [2], [6]:

- Pressurized water scrubbing (PWS): is the most used and consolidated technology, it can achieve a purity of 80-99% with low losses 3-5%, in some case lower than 2%. The energy consumption, only electric power, is mainly used to compress raw gas and processing water.
- Physical absorption (PA): is similar to PWS but instead of water it is used an organic solvent to have a higher solubility of the $CO_2$, the system needs thermal power to regenerate the solvent.
- Chemical absorption (CA): the solvent, usually amines, react selectively with carbon dioxide to remove it from the biogas. The technology can reach a high purity of methane (99%) and $CO_2$ but a large amount of thermal energy is needed to regenerate solvent.
- Cryogenic separation (CS): the methane and carbon dioxide are separated by condensation and distillation. This technology is still under development and has a high energy demand and it is economical expensive, but it allows to obtain a high concentration of methane in the product, and also a high concentration of carbon dioxide in byproduct. CS can be useful if the goal is to produce liquefied biomethane for transport.
- Pressure swing adsorption (PSA): the process is based on the selective adsorption on the solid surface of the molecule based on their size. The concentration of $CH_4$ is between 96% and 98% but higher is the purity request higher are the losses of methane.
- Membrane separation (MS): is based on the selective permeability of membrane which can be crossed by $CO_2$ and not by $CH_4$. MS is considered cheap, simple and efficiency but is not possible to achieve high methane concentration without many stages.

For the injection in the natural gas grid the biomethane have to respect the limiting value of physical properties and concentration of components prescribed by the national regulation, for Italy the values are reported in *Table 3* and *Table 4*:

| Properties | Value | |
|---|---|---|
| HHV | 34.95 – 45.28 | $MJ/Sm^3$ |
| Wobbe index | 47.31 – 52.33 | $MJ/Sm^3$ |
| Relative density | 0.5548 – 0.8 | |
| Water dew point | ≤ -5 | °C |
| Hydrocarbon dew point | ≤ 0 | °C |

*Table 3: Physical properties* [7]

| Components | Concentration | |
|---|---|---|
| Oxygen | 0,6 | % vol. |
| Carbon dioxide | 3 | % vol. |
| $H_2S$ | ≤ 6.6 | $mg/m^3$ |
| Total sulfur | ≤ 150 | $mg/m^3$ |

*Table 4: Admissible compounds* [7]

The components not reported in the *Table 4:* methane, ethane, propane, hydrogen and nitrogen, their limiting concentration in the mixture for the injection in the grid is automatically defined by the value of the Wobbe index defined in *equation* (*1*) where HHV is the higher heating value of the mixture and ρ is the relative density:

$$I_W = \frac{HHV}{\sqrt{\rho}} \tag{1}$$

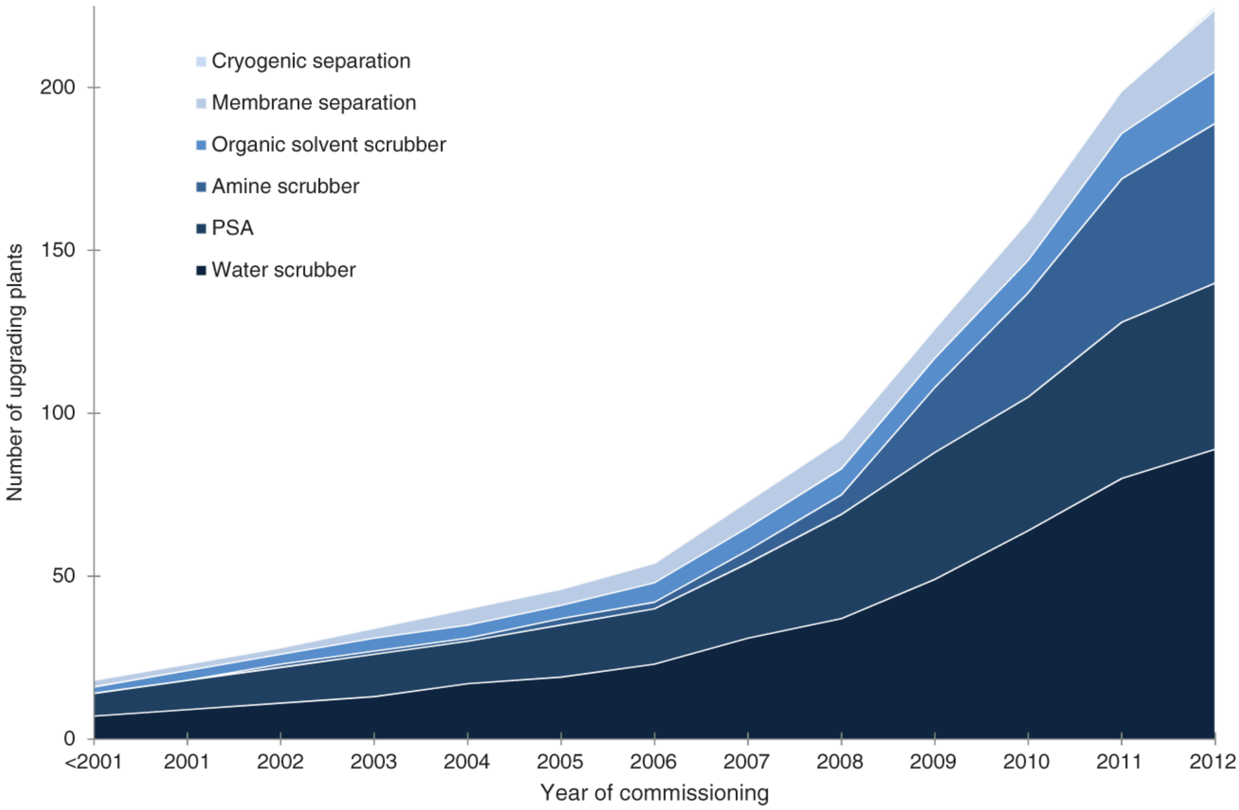


*Figure 4: biogas upgrading plant in Europe* [1]

# 1.2 SOFC-based CHP system

In the case of use of biogas in combined heat and power generation is not mandatory the upgrading of biogas, is only necessary the pretreatment of the fuel to remove the pollutant substance that can cause the problem of corrosion. The main technology for CHP application fed by biogas is the internal combustion engine that operates with an electrical efficiency lower than 42% and a thermal efficiency between 42% and 43% as reported in the technical paper of Jenbacher [8].

A promising alternative to the internal combustion engine or micro-turbine for the utilization of biogas in distributed cogeneration power plant are the fuel cells which produce electric power using the electrochemical reaction splitting the reaction of combustion in two semi-reaction: oxidation of the fuel at the anode and reduction of the oxygen at the cathode, the two electrodes are separated by an electrolytic membrane that allows the passage of positive cations but not permits the passage of electrons.

For the direct utilization of biogas in stationary power generation the best options are the high-temperature fuel cells like Solid Oxide Fuel Cell (SOFC) and Molten Carbonate Fuel Cell (MCFC) which operate at temperature between 600°C and 900°C, they can achieve higher electrical efficiency (>50% [9], [10], [11]) and manage the carbon contained in the fuel thanks to higher fuel flexibility [2]. $H_2S$ must be removed to avoid poisoning of the nickel-based catalyst, also the removal of siloxane is necessary to avoid deposition of $SiO_2$ on the anode.

The SOFC is a technology under development and it is not jet at commercialized scale, there are technical and economic issues, especially related to lifetime and cost, to resolve before SOFC can become competitive in the global power market. In the past decades, the SOFC technology made a great advance and many theoretical and experimental types of research have been developed to investigate the technical solution and configuration of the plants with the goal to reach the thermo-economic optimization on the power plant based on SOFC. The principal context of research regards the advantages of partial recirculation of the anodic and cathodic products [5], the thermal integration of the plant through a net of heat exchanger [10], comparison between atmospheric and pressurized plant combined with turbo-gas bottoming cycle [10], [11] and the possibility to add the carbon capture and storage (CCS) concept to the SOFC power plant [10], [11].

Upgrading and Solid Oxide Fuel Cell represent two different ways to valorize the biogas, an important renewable energy source. A possible guideline to compare upgrading and SOFC is reported in reference [12] where the direct use of biogas in CHP plant and upgrading are compared on the base of biogas yield: if the yearly production of biogas is higher than 3.5 million $m^3$/y the best solution is the upgrading due to the risk of limited demand for CHP product especially the thermal power.

# 1.3 Comparison method

The scope of this analysis is to compare the two technologies from the technical and economical point of view through the exergo-economic analysis. This methodology gives a technical evaluation of the two plants based on exergy and economic performances and allows to compare productive plant with two different products from the energy point of view like biomethane and electric power. The analyzed biogas upgrading plant is based on PWS with air stripping to regenerate the process water, instead, the SOFC plant operates at ambient pressure with partial recirculation of anodic products and a partial external reforming of feeding biogas. Both plants have the same biogas volumetric flow equal to 120 $Sm^3/h$ with a standard composition of 60% methane and 40% carbon dioxide, the model does not consider the presence of other substances in the biogas mixture, all the contaminants are removed in the cleaning section. With the selected composition of the biogas, its LHV results in 20.38 $MJ/Sm^3$, considering the volumetric flow feeding the plants the chemical energy entering is equal to 680 kW. The volumetric flow of biogas feeding the plants analyzed can correspond to about one ton per hour of municipal solid waste or food industrial waste or about two tons per hour of animal waste of sewage sludge (following the *Table 2*). The selected size of biogas production is smaller than the average size of the existing biogas production plant (>500 $Sm^3/h$ [3]), the small size is useful to investigate the possibility to exploit the biogas produced by the diffuse small plant. Both plants are modeled with AspenPlus® to simulate the thermodynamic behavior and to obtain the value of the principal thermodynamic variable of each stream of mass.

# 2 Case study

## 2.1 Upgrade to biomethane

Figure 5 shows the layout of the analyzed biogas upgrading plant.

Before entering upgrading plant, the biogas (*biogas0*) passes through CLEAN-UP bed: this process is needed in order to remove from the raw biogas trace compounds like sulfur, chlorine, siloxane and also water that can decrease the efficiency of the plant and degrade equipment. There are several cleaning techniques but the most used is the adsorption with active carbon treated with metal ions [13]. Clean biogas is then mixed with the gas recirculated from FLASH tank.

The core of the plant is the SCRUBBER where the stream *biogas3* enters in counter-flow with nebulized water (*water2)*; since carbon dioxide has a higher solubility in water compared to methane, the gas exiting this component (*biomtn1)* is rich in methane and poor in carbon dioxide. The process is enhanced at high pressure and low temperature, for this reason, the biogas is pressurized in compressor K1 up to 10.5 bar and cooled to 60°C in COOL before entering the SCRUBBER. The water necessary to the process is pressurized by PUMP1.

The liquid stream (*liquid1)* coming from the SCRUBBER undergoes a separation process in the FLASH tank to recover most of the remaining $CH_4$ which is dissolved. Pressure drops down from 10 to 3 bar and this produces a flash expansion that separates the liquid from the dissolved gas. The gas that contains methane (*gasrec1*) is then recirculated through a circuit made by valve V1 to decrease the pressure and dryer DRY1 to remove water. The stream *liquid2* goes to the valve V2 and after to the STRIPPER, operating at ambient pressure where an air flow (*air1*), taken from environment by the blower K2, strips the carbon dioxide from the water, which is regenerated, and it is ready to be reused in the SCRUBBER; the water recirculation is guaranteed by PUMP2. The air that now is rich in carbon dioxide is released to the environment (*gas-out*).

The gas stream exiting the SCRUBBER enters the dryer DRY2 to eliminate all the water vapor, making the resulting gas ready to be injected in the natural gas network as biomethane (*biomtn2*). In the plant, biomethane is produced at a pressure of 10 bar that is considered sufficient for the injection in the national gas net. The process of pressurized water scrubbing (PWS) can produce a biomethane with a concentration of methane higher than 98% and losses of methane lower than 0,7%.

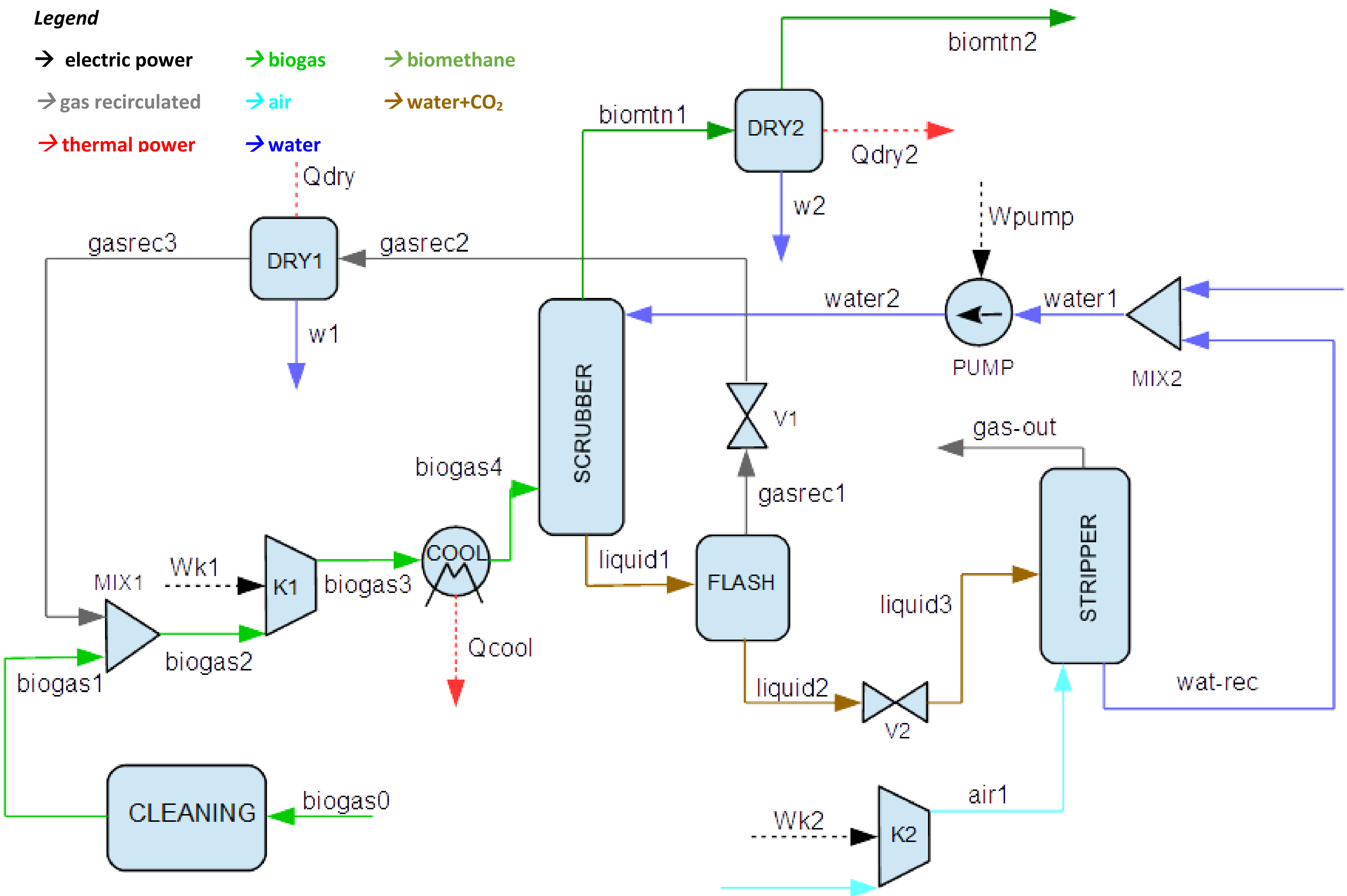


*Figure 5: Biogas upgrading plant*

The external assumption regarding the efficiency of pumps and compressor used in AspenPlus® simulation are the following:

- isentropic efficiency of compressor $\eta_{id}$ = 0.82
- electrical efficiency of compressor $\eta_{el}$ = 0.97
- hydraulic efficiency of pump $\eta_{is}$ = 0.85
- electrical efficiency of pump $\eta_{el}$ = 0.93

For the system described in *Figure 5,* the productive structure is reported in *Table 5,* the thermodynamic results of the Aspen–Plus® simulation are reported in the appendix *Table 37.*

| COMPONETS | FUEL | PRODUCT | LOSS |
|---|---|---|---|
| CLEAN-UP | biogas0 | biogas1 | |
| MIX1 | biogas1 + gasrec2 | biogas2 | |
| K1 | Wk1 | biogas3 - biogas2 | |
| COOL | biogas3 | biogas4 | Qcool |
| PUMP1 | Wpump1 | water2 - water1 | |
| SCRUBBER | biogas4 + water2 | biomtn1 + liquid1 | |
| DRY2 | biomtn1 | biomtn2 | w2 + Qdry2 |
| FLASH | liquid1 | liquid2 + gasrec1 | |
| V1 | gasrec1 | gasrec2 | |
| DRY1 | gasrec2 | gasrec3 | w1 + Qdry |
| V2 | liquid2 | liquid3 | |
| DRY1 | gasrec2 | gasrec3 | Qdry1 + w1 |
| K2 | Wk2 | air1 - air | |
| STRIPPER | liquid2 + air1 | pwat | gas_out |
| MIX2 | wat_rec + water0 | water1 | |
| global | biogas0 + Wk1 + Wk2 + Wpump1 | biomtn1 | Qcool + w1 + Qdry1 + w2 + Qdry2 + gas_out |

*Table 5: productive structure of biogas upgrade to biomethane plant*

## 2.2 SOFC-based CHP system

The layout of the biogas exploitation in CHP mode is shown in Figure 6.

The system is a combined heat and power (CHP) based on Solid Oxide Fuel Cells (SOFC) fed by biogas at atmospheric pressure. The blowers are needed only to compensate the pressure losses of the downstream components. The fuel for the plant is the stream *biogas0* that enters the blower FUEL-CMP.

The biogas passes through the clean-up bed (CLEANING) to remove the trace compounds like siloxane, chlorine, sulfidic acid and water that may cause the fast degradation of the fuel cell by the poisoning of the catalyst or the obstruction of the micro-pore of the electrode.

The clean biogas (*biogas2)* is pre-heated in FUEL-HX from 28°C (outlet temperature from the blower) to 646°C and mixed with part of the anodic exhaust flow (*an-rec2*). The outlet temperature of FUEL-HX is automatically calculated by the software in order to reach, at the SOFC anode inlet, a temperature of 750°C, taking into account also the amount of heat provided by the recirculation stream (*an-rec2).* The anodic recirculation is used to achieve a steam to carbon ratio (S/C) equal to 2 using the water produced by the redox reaction occurring in the fuel cell.

The mixture of biogas and anode recirculated flow (*biogas4)* enters the external reformer vessel (EXT-REF) which performs the steam reforming of methane. The technical choice is to reform only 50% of the stream in the external vessel and the remaining part is reformed directly in the anode of the fuel cell. External reforming is useful to avoid the internal reforming of the entire mass of biogas because the steam reforming is an endothermic reaction and it can create high-temperature gradients inside the electrode compromising the stability of the ceramic layer of the cell [10]. The external reformer and the SOFC are included in the same module, so the thermal power necessary to sustain the external reforming is given directly by the heat generated by the exothermic reaction of the fuel cell (*Qref)*.

The oxidant for the redox reaction is the oxygen present in the air *(air0)*, which is taken from the environment by the blower AIR-CMP and heated up from 25°C (outlet temperature from the blower) to 650°C before entering the cathode of the SOFC. The flow of air is in excess compared to the stoichiometric needs because the air is also used as cooling fluid for the fuel cell, to remove the heat produced by the exothermic reaction and not absorbed by the endothermic reforming reaction.

The core of the CHP plant is the Solid Oxide Fuel Cell (SOFC): the electrical power (*Wel)* is produced in this component by the reactions of oxidation of the fuel (hydrogen and carbon monoxide) in the anode and the reaction of reduction of oxygen at the cathode. The SOFC operates at 800°C and with a fuel utilization equal to 90%. The anode and cathode exhaust streams have a high thermal potential and also a residual chemical potential due to the presence of unburned fuel in the anodic flue gas (*an-ex*) and oxygen in the cathodic exhaust (*cat-ex*). Part of the anode products are recirculated, the other is sent to after-burner, in order to recover the residual chemical energy.

Inside the BURNER the remaining hydrogen is burned with the air coming from the cathode: the injection of liquid water *(water)* is necessary to keep the temperature of the flue gas (*exhaust1*) under 900°C that is the maximum admissible temperature for a standard vessel material (Ni alloy). The exhaust of the after-burner is used firstly to pre-heat inlet fuel and air streams and then to heat up a liquid water flow from 65°C to 75°C inside CHP-HX: the temperatures are compatible with a small district heating network.

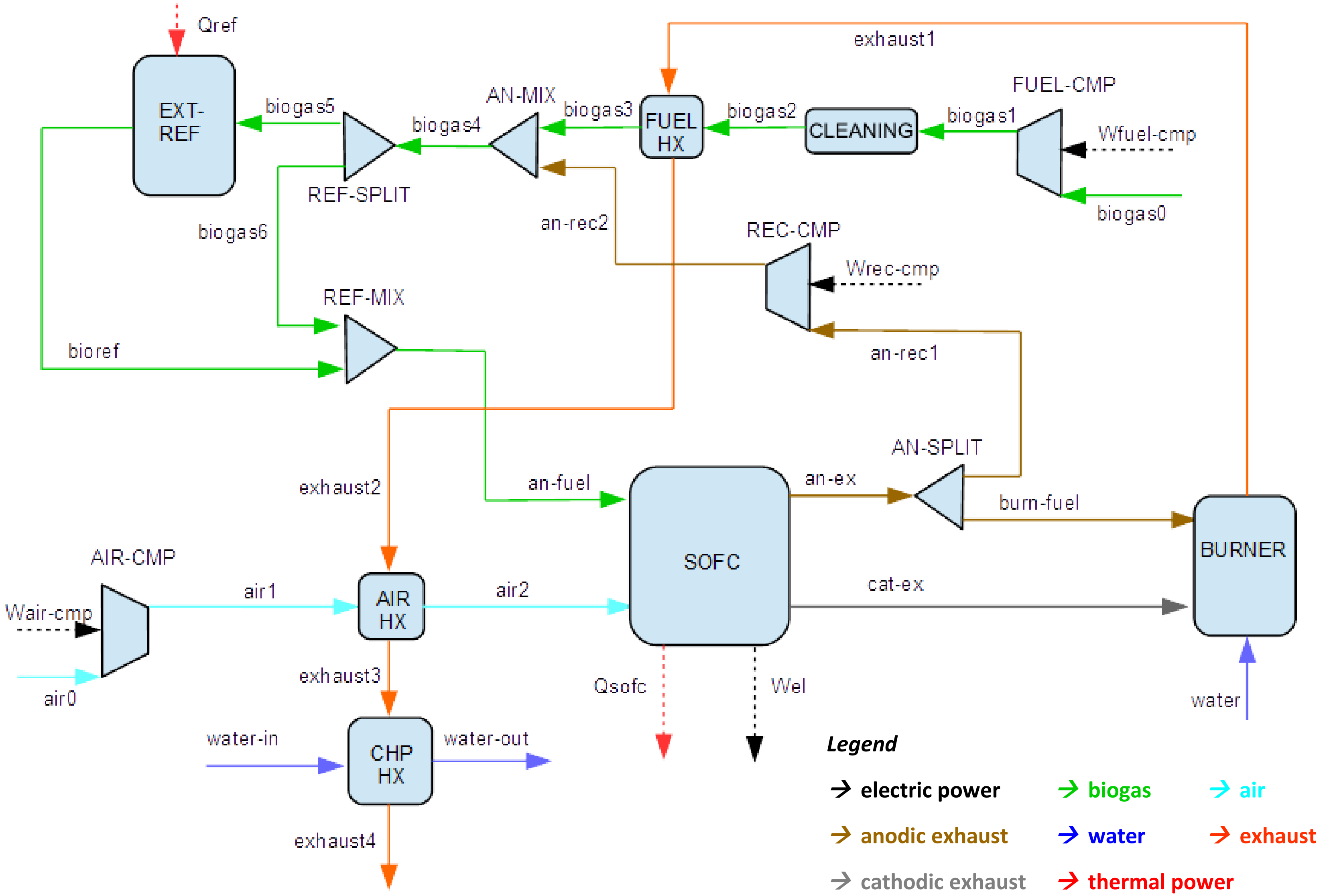


*Figure 6: SOFC plant*

The hypotheses on the efficiencies of the blower used in Aspen-Plus® simulation are the following:

- Isentropic efficiency of blowers $\eta_{id}$ = 0.82
- Electrical efficiency of blowers $\eta_{el}$ = 0.97

Combined heat and power SOFC plant has a productive structure described in *Table 6,* the thermodynamic results of the AspenPlus® simulation are reported in the appendix *Table 40.*

| **COMPONENTS** | **FUEL** | **PRODUCT** | **LOSS** |
|---|---|---|---|
| FUEL-CMP | Wfuelcmp | bioga1 - biogas0 | |
| CLEANING | biogas1 | biosas2 | |
| FUEL-HX | exhaust1 - exhaust2 | biogas3 - biogas2 | |
| AN-MIX | biogas3 + an_rec2 | biogas4 | |
| REF-SPLIT | biogas4 | biogas5 + biogas6 | |
| EXT-REF | biogas5 + Qref | bioref | |
| REF-MIX | biogas6 + bioref | an-fuel | |
| SOFC | an_fuel + air2 | Wel + Qsofc + an_ex + cat_ex | |
| AIR-CMP | Waircmp | air1 - air0 | |
| AIR-HX | exhaust2 - exhaust3 | air2 - air1 | |
| AN-SPLIT | an_ex | an-rec + burn_fuel | |
| REC-CMP | Wreccmp | an_rec2 - an_rec1 | |
| BURNER | burn_fuel + cat_ex + water | exhaust1 | |
| CHP-HX | exhaust3 | water_out - water_in | exhaust4 |
| global | biogas0 | Wel + (water_out - water_in) | exhaust4 |

*Table 6: productive structure of the SOFC-based CHP plant*

# 3 Methodology

## 3.1 Energy analysis

The first part of the analysis regards the energy characterization of the plants through the first principle efficiency, for the biogas upgrading plant the product is the stream of biomethane and its energy is the Lower Heating Value, instead, the fuels are the inlet biogas and the electric power absorbed by the two compressors and the pump, see *equation* (2).

$$\eta = \frac{\dot{m}_{biomtn} \cdot LHV_{biomtn}}{\dot{m}_{biogas} \cdot LHV_{biogas} + W_{k1} + W_{k2} + W_{pump}} \tag{2}$$

From the definition the combined heat and power plant has two different products: electrical power and recovered heat, so two different energy efficiencies are defined. Electrical efficiency considers the net power injected to the grid, *equation* (*3*), the thermal efficiency considers the thermal power recovered in the CHP-HX, *equation*(*4*), in both cases the fuel is only the biogas feeding the plant because the power consumed by auxiliary is not taken from the grid.

$$\eta_{el} = \frac{\eta_{inv} \cdot W_{el} - W_{fuel-com} - W_{air-cmp} - W_{rec-cmp}}{\dot{m}_{biogas} \cdot LHV_{biogas}} \tag{3}$$

$$\eta_{th} = \frac{\dot{m}_{water\_in} \cdot (h_{water\_out} - h_{water\_in})}{\dot{m}_{biogas} \cdot LHV_{biogas}} \tag{4}$$

## 3.2 Defining exergy

The exergy is a thermodynamic variable that measures the quality of a given type of energy and expresses the maximum theoretical work obtainable from the interaction between environment and a system evolving from the initial condition to a condition of equilibrium in terms of pressure, temperature and chemical to the environment. The environment is defined as an ideal system that surrounds the analyzed system and it is characterized by constant and uniform temperature $T_0$ and pressure $p_0$ and consist of a set of reference substances with standard concentration. When a system is in equilibrium with the environment is not possible to extract work from it, this state of the system is called dead state [14].

For mechanical and electrical power, the exergy associated is equal to the power transferred

$$E_w = W \tag{5}$$

For the thermal power, the exergy related is the heat flux multiply by the Carnot factor:

$$E_{th} = Q\left(1 - \frac{T_0}{T}\right) \tag{6}$$

There are two types of exergy for a mass stream: physical and chemical exergy.

The physical exergy measures the distance from the equilibrium with the environment, from the mechanical point of view, in terms of pressure, temperature velocity and height and it is calculated as:

$$e_{ph} = (h - h_0) - T_0(s - s_0) + \frac{V^2}{2} + gz \tag{7}$$

Where $h$ and $s$ are calculated in the system condition of temperature and pressure and $h_0$ and $s_0$ are calculated for the same stream at pressure and temperature of the dead state $p_0$ and $T_0$, $V$ is the velocity, $z$ is the elevation and $g$ is the gravity acceleration. The last two contributions are negligible in the studied case.

The chemical exergy of a mass stream is due to a difference of chemical composition between the stream itself and the environment. It is calculated for every component of the mixture as the difference between the chemical potential of the substance in the mixture and the chemical potential of the same component in the dead state:

$$e_{ch} = \mu_0 - \mu_{00} \tag{8}$$

Where $\mu_0$ is the chemical potential at pressure and temperature of the dead state but at the concentration of the stream, and $\mu_{00}$ is calculated also at the concentration of the environment. For a component that is not present in the reference ambient the chemical exergy is:

$$e_{ch} = -\Delta G - \sum \upsilon_j\left(\mu_{0j} - \mu_{00j}\right) + \sum \nu_k(\mu_{0k} - \mu_{00k}) \tag{9}$$

*Where* $\Delta G$ is the variation of the Gibbs free energy for the reaction that produces the substance starting from molecules present in the environment, $v_j$ are the stoichiometric coefficients of the reactant and $v_k$ are the stoichiometric coefficients of the products of the reaction.

For the gas, the hypothesis of ideal gas is applied and so the chemical exergy for molecules present in the atmosphere is calculated as:

$$e_{ch} = -R \cdot T_0 \cdot \ln(y) \tag{10}$$

The chemical exergy for methane is calculated as:

$$e_{ch} = -\Delta g + R \cdot T_0 \cdot ln\left(\frac{y_{O2}^2}{y_{CO2} \cdot y_{H2O}^2}\right) \tag{11}$$

Where $y$ is the concentration of the substances in the atmosphere. For gas mixtures the equation is:

$$e_{ch} = \sum y_{o,i} \cdot e_{ch,i} + R \cdot T_0 \cdot \ln\left(y_{o,i}\right) \tag{12}$$

*Where* $y_{0,i}$ is the mass fraction of the chemical component in the mixture and $e_{ch,i}$ is the chemical exergy of the pure substance.

For a liquid mixture that contains carbon dioxide and methane:

$$e_{ch} = x_{CH4} \cdot e_{CH4} + x_{CO2} \cdot e_{CO2} \tag{13}$$

Where *x* is the mass fraction of component and *e* is the specific chemical exergy of the substance.

For the exergy analysis - developed for both plants - the reference environment is defined in *Table 7* in terms of temperature, pressure and chemical composition of the two-phase considered; due to the absence of solid phase in the streams of both plants, the solid phase is not considered in the reference environment.

| **Environment** | | |
|---|---|---|
| $T_0$ = 293.15K (20°C), $p_0$ = 1atm (101325 Pa) | | |
| | **Component** | **Concentration** |
| Condensed phase | $H_2O(l)$ | 100% |
| Gas phase | $N_2$ | 75.66% |
| | $O_2$ | 20.35% |
| | $H_2O(g)$ | 3.12% |
| | $CO_2$ | 0.04% |
| | Ar | 0.83% |

*Table 7: reference environment* [14]

## 3.3 Exergy account methodology

The exergy account methodology is a type of analysis that can be developed for a productive plant this method is used to identify the flux of exergy in the system studied, the exergy lost and the process of production of the output of the plant.

The system is described by the incidence matrix $\bar{\bar{A}}$ that is a matrix whit *m* rows and *n* columns, where *m* is the number of component of the system and *n* is the number of ̀streams (mass, work and heat streams). Every position *ij* of the matrix is equal to 1 if the stream *i* is entering in the component *j*, -1 if the stream *i* is exiting from the component *j* and 0 otherwise.

### 3.3.1 Exergy lost

Energy is a conservative quantity so, in every transformation, the total amount of energy does not change. Exergy, on the other side, is not a conservative quantity and the exergy lost in a component and in the total plant is a useful index to analyze the system. For every component of the plant is possible to write a balance of exergy flow

$$\sum(E)_{in} - \sum(E)_{out} = I \tag{14}$$

where *I* is the exergy lost in the component. The balance can be written for the entire system in a matrix form:

$$\bar{\bar{A}} \cdot \bar{E} = \bar{I} \qquad (15)$$

Where $\bar{\bar{A}}$ is the incidence matrix, $\bar{E}$ is the vector of exergy that contain the exergy flow of each stream and $\bar{I}$ is the vector of the exergy lost.

The exergy destruction rate $\psi$ compares the exergy lost in a component or in a system and the exergy flow of the fuels entering the same component:

$$\psi = \frac{I}{E_{fuel}} \qquad (16)$$

### 3.3.2 Exergy efficiency and exergy factor

The exergy analysis is useful to determine the exergy efficiency ε (second principle of thermodynamic performance) for each component of the plant according to the productive structure chosen for the system. The exergy efficiency of the whole plant and of every component is defined as the ratio between the sum of the products exergies and the sum of the fuel exergies:

$$\varepsilon = \frac{\sum E_{products}}{\sum E_{fuels}} \qquad (17)$$

The exergy factor $\lambda$ is another useful index that relates the fuel exergy spent in the *k*-component with the total fuel exergy processed in the plant and it is defined as:

$$\lambda_k = \frac{E_{fuel,k}}{E_{fuel,TOT}} \qquad (18)$$

### 3.3.3 Exergy cost

The exergy cost for a mass or energy stream is defined as the exergy of the stream itself plus the exergy lost due to irreversibility in the upstream transformations needed to generate that stream and it is identified by the symbol *E**.

The exergy cost is a conservative quantity, so for every component of the system the balance of exergy costs of steams entering and exiting is:

$$\sum E_{in}^{*} - \sum E_{out}^{*} = 0 \qquad (19)$$

And in the matrix formulation for the full plant:

$$\bar{\bar{A}} \cdot \overline{E^{*}} = 0 \qquad (20)$$

The exergy cost for every stream can be found solving this system of equation: the number of streams *n* is usually higher than the number of components of the plant *m*, so to solve the problem is necessary to add *n-m* auxiliary equations. There are also four types of rules for external conditions:

- 1° rule is related to the fuel of the plant. For each stream that represents a fuel the exergy cost must be defined: if there is no information about that stream, the exergy cost is imposed equal to the exergy itself.
- 2° rule is related to the losses. The exergy cost of streams that are considered as losses for the system is imposed equal to zero; in this way, the exergy cost of losses is added to the products.
- 3° rule is applicable to components where the fuel is a difference of two streams: in this case, the specific exergy cost of both streams is considered equal.
- 4° rule is applicable to components that have more than one product: every steam of products has the same specific exergy cost.

The system of equations is then modified to include the auxiliary equation and becomes a system of *n* equation in *n* unknows that have only one solution:

$$\overline{\overline{Ac}} \cdot \overline{E^*} = \bar{Y} \tag{21}$$

Where $\overline{\overline{Ac}}$ is the cost matrix that is composed by the incidence matrix and the equations derived by the external conditions, $\bar{Y}$ is the vector of the know term.

Starting from exergy cost is possible to define the specific exergy cost $k^*$ of a stream of mass, work or heat as the ratio between exergy cost and the exergy of the stream itself:

$$k^* = \frac{E^*}{E} \tag{22}$$

From the definition, the specific exergy cost is always higher than one.

## 3.4 Exergo-economic analysis

The goal of this analysis is to integrate the exergy and economic results in the same framework, in particular, it allows to define the economic cost for each stream of mass and energy in the plant with an analytic procedure.

The balance equation for a single component is:

$$\sum C_{fuels} + Z_c = \sum C_{products} \tag{23}$$

where $C_{fuels}$ is the cost of each stream entering the component, $C_{products}$ is the cost of streams exiting component and $Z_c$ represents the cost of the component itself. All terms are expressed in €/s in this equation.

To convert the price of components from € to €/s is firstly necessary to calculate the annualized cost by using the Capital Recovery Factor (CRF) *(24)*

$$CRF = \frac{A}{P} = \frac{i \cdot (1+i)^n}{(1+i)^n - 1} \tag{24}$$

Where *P* is the total cost of the component (€) and *A* is the annualized cost (€/year), *i* is the discount rate that represents the effect of inflation and *n* is the number of years. Starting from annualized cost the term $Z_c$ is calculated as:

$$Z_c = \frac{A}{3600 \cdot 8700 \cdot f} \tag{25}$$

The operation and maintenance cost of the component can be included in the value of *A*. This is usually expressed as a percentage of the cost of component per year. In the equation *f* is the capacity factor, the fraction of time in the year during which the plant operates.

The balance equations for every component gives a set of equation that can be written in matrix form as:

$$\bar{\bar{A}} \cdot \bar{C} = \overline{Z_c} \tag{26}$$

where $\bar{\bar{A}}$ is the incidence matrix of the plant, $\bar{C}$ is the vector containing the cost of each stream and $\overline{Z_c}$ is the vector of the cost of components.

The system has *m* equations and *n* unknowns, so (*n-m*) auxiliary equations are needed to solve the problem, the additional hypotheses that have to be added can be written using the same rules defined in the exergy analysis (rule P1 P2 P3 and P4). Finally, we obtain the complete system to be solved:

$$\overline{\overline{A_c}} \cdot \bar{C} = \bar{Z} \tag{27}$$

Where $\overline{\overline{A_c}}$ is the cost matrix, which coincides with the one defined in the exergy analysis and the vector $\bar{Z}$ contains the cost of components, cost of fuels and losses.

### 3.4.1 Cost of components

Cost functions are useful tools in order to obtain the price of each component, in particular in this paper are used the cost function from "Analysis Synthesis and Design of Chemical Process" (Turton, Bailie)[15] in which the base cost of common chemical plant components is given by the *equation* (28)*:*

$$log_{10}C_p^0 = K_1 + K_2 \cdot log_{10}(A) + K_3 \cdot [log_{10}(A)]^2 \tag{28}$$

The parameter *A* is a physical quantity that represents the size of equipment, like the power for a compressor or the surface for a heat exchanger, constants $K_1$, $K_2$ and $K_3$ are a statistical fitting parameter for the specific component. $C_p^0$ is the price of the component for atmospheric operation.

If the size of equipment is outside the validity limit of the cost function for the type of component the cost is calculated starting from the cost of the component at lower or upper limiting size applying a scaling factor:

$$C_1 = C_0 \cdot \left(\frac{A_1}{A_0}\right)^n \tag{29}$$

Where $C_1$is the cost of the component that has the requested size $A_1$, $C_0$is the cost of the component that has the reference size $A_0$ and n is a scaling factor assume equal to 0.6. This methodology is used also to determine the cost of the cleaning bed. To obtain the bare erected cost two other factor are needed.

Pressure factor $F_p$ consider the pressure of operation of the component, for a vessel the formula is

$$F_p = \frac{\frac{(p+1)\cdot D}{2(850-0{,}6\cdot p)} + 0{,}00315}{0{,}0063} \tag{30}$$

$p$ is the operative relative pressure (barg) and D is the diameter of the vessel (m). For other equipment the pressure factor is given by the equation (31):

$$log_{10}F_p = C_1 + C_2 log_{10}(p) + C_3 \cdot [log_{10}(p)]^2 \tag{31}$$

Material factor $F_m$ is given in tabular form in function of the component and the material of construction that can be carbon steel, stainless steel or other alloys. Based on the type of element there are different equations to calculate the bare erected factor $F_{bm}$ using pressure and material factor. The bare erected cost is given by the equation:

$$C_{bm} = C_p^0 \cdot F_{bm} \tag{32}$$

### 3.4.2 Cost estimation methodology

The bare erected cost given by cost function comprises only the cost related to process equipment and direct and indirect labor related to installation. The NETL/DOE [16] suggests four levels of capital cost that are summarized in *Figure 7*.

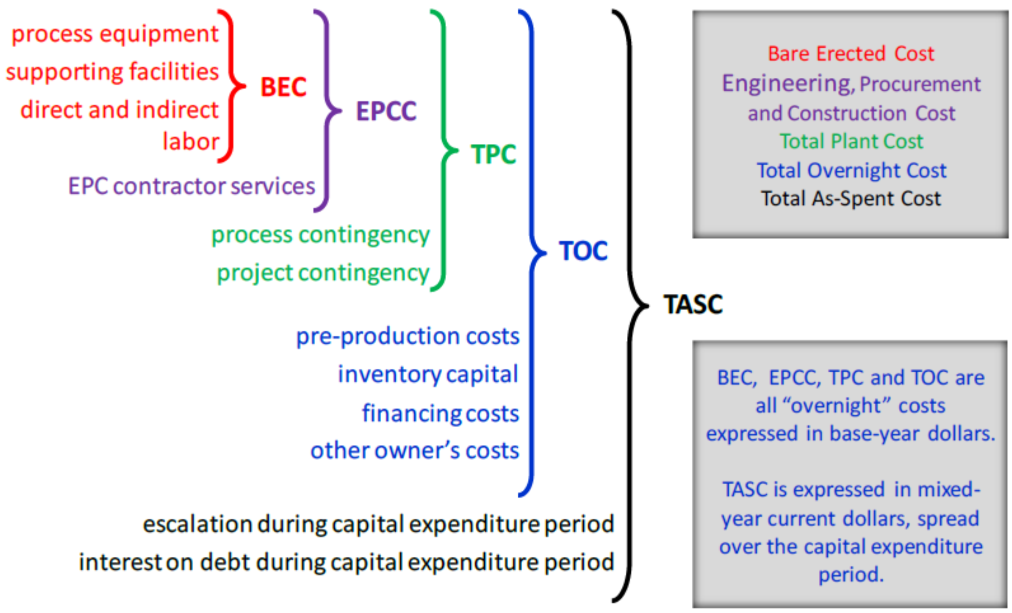


*Figure 7: NETL/DoE cost level* [16]

## 3.4.3 Exergo-economic indexes

To fully analyze a plant from the thermo-economic point of view is used to calculate some parameter for each component. The first index is the relative cost difference *r*, which indicate the increment in cost between fuels and products in a certain unit, the relative cost difference is defined as:

$$r = \frac{c_p - c_f}{c_f} \tag{33}$$

For each section fuels and products are defined according to the productive structure. If product or fuel are composed by sum or difference of streams the total cost is the weighted average of the cost of every stream.

$$c_{p,f} = \frac{c_i \cdot E_i \pm c_j \cdot E_j}{E_i \pm E_j} \tag{34}$$

The elements that have high relative cost difference are the most critical in the process of cost formation of products of the plant.

The cost difference between fuels and products is caused by two factors, the first one is the thermodynamic inefficiency or loss of exergy, the second is the economic cost of the component. The exergo-economic factor gives an analytic way to understand which of the two contributes affect more the relative cost difference calculated for the module. The exergo-economic factor is defined as:

$$f_{ex} = \frac{Z}{Z + C_D} \tag{35}$$

Where *Z* is the economic cost of a component already defined starting from the annualized cost and $C_D$is the cost of irreversibility that represent the economic value of the exergy lost and is defined as:

$$C_D = c_{fuel} \cdot I \tag{36}$$

Where *I* is the exergy lost in the component. If $f_{ex}$ is near to one the module is characterized by high efficiency, but also high investment cost so is possible to reduce the cost difference applying a cheaper and less efficient equipment. If $f_{ex}$ is near to zero, the section has low capital cost, but high thermodynamic inefficiency, in this case is possible to select a different type of component with higher efficiency and investment cost to reduce the cost difference. The described methodology is useful to achieve the thermo-economic optimization of the studied plant.

# 3.5 Economic analysis

The last analysis developed on the two plant regards the economic performance along the lifetime of systems. From the economical point of view, every productive plant is characterized by:

- an initial investment to build plant itself
- one or more products that are sold to the market
- one or more fuels necessary to the production
- operation and maintenance cost

The two main results of these analyses are the net present value (NPV) of the investment and the pay-back time (PBT) of each plant.

The net present value is the actualized value of an investment that gives revenue in the following years. For every year are considered cost and revenue that compose the net cash flow, from this value are subtracted the taxes and then is actualized. The net present value is the sum of the actualized cash flow of every year following the *equation 37*:

$$NPV = -IC + \sum CF \cdot \frac{1}{(1+i)^n} \tag{37}$$

Where *IC* is the investment cost, *CF* is the annual cash flow, *i* is the discount rate and *n* is the number of years. The investment cost is already determined by the exergo-economic analysis, the cash flow instead is determined by the difference between revenue and cost. The annual costs for both plants are represented by:

- biogas that is the principal fuel system (0.14 €/$Sm^3$ [6])
- electric power needed to compressors pumps or blowers (0.16 €/kWh [17])
- operation and maintenance costs that are considered 3% of TASC

The revenues are the result of the selling of the products of the plant at the market price considering also public subsidies to the production of renewable energy.

The pay-back time is defined as the time in which the NPV is zero or, in a simpler way, is the year in which the NPV change the sign, from negative to positive.

## 3.5.1 Upgrade to biomethane

The referenced law about public subsidies to biogas upgrading to biomethane for injection in the grid is the D.M. 5/12/2013 [18]. The ministerial decree introduces a public incentive for the plant that produces biomethane in the form of a contribution for the net renewable energy produced (€/MWh) considering the consumption of not renewable energy during the process. The methodology to calculate the public subsidies for a biogas upgrading plant is the following:

1. The base incentive is calculated as the difference between the double of the yearly average market price of natural gas in 2012 and the monthly average market price of the natural gas in the month of selling.

$$PS_{base} = 2 \cdot C_{2012} - C_{month} \tag{38}$$

2. The base subsidy is then modulated according to the size of the plant expressed by the productivity of biomethane:
   - +10% for a plant with a productivity lower than 500 $Sm^3/h$
   - 0% for a plant with a productivity between 501 $Sm^3/h$ and 1000 $Sm^3/h$
   - -10% for a plant with a productivity higher than 1000 $Sm^3/h$

$$PS_{mod} = k \cdot PS_{base} \tag{39}$$

3. The modulated incentive can be increased by another 50% if the biogas that feeds the upgrading plant comes from the anaerobic digestion of waste products reported in the "Table 1A of D.M. 6 Luglio 2012" [19].

The subsidies are recognized only for the renewable part of the biomethane produced so, from the chemical energy of the biofuel have to subtract the electric power consumed that is not considered renewable. To compare the two different form of energy, chemical and electrical is necessary to convert both in tons of oil equivalent (toe) following the equivalation suggested in the "Circolare Ministeriale n°219/F del 2 Marzo 1992 art. 21 Tabella A" [20]:

- For the biomethane 1000 $Nm^3$ = 0.82 toe
- For electric consumption 1 toe = 0.23 MWh

With these conversions is possible to calculate the renewable energy really produced in toe as the difference of the products and fuels of the system and, doing the reversal conversion, to calculate the equivalent production of biomethane in standard cubic meter to which is applied the incentive. The dates about the price of natural gas in the Italian market are available on the website of GME ("Gestore Mercato Energetico") [21].

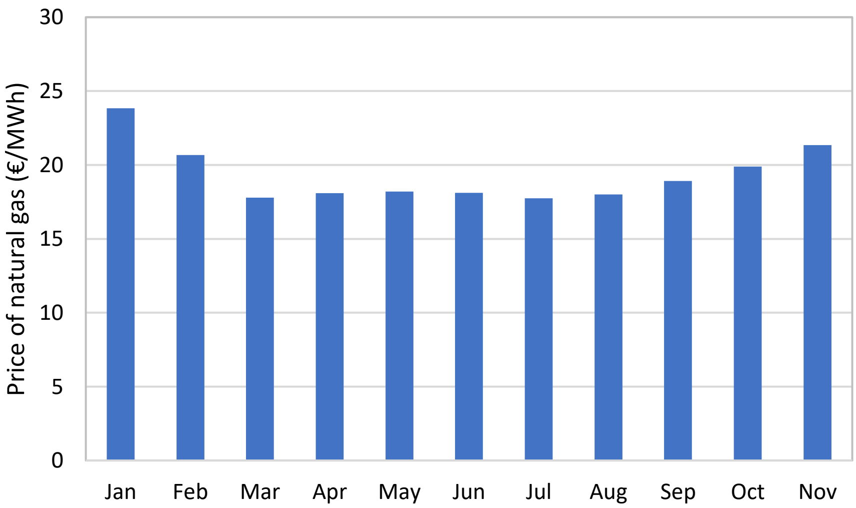


*Figure 8: Market price of natural gas in Italy 2017* [21]

In the case of biogas upgrading for transport use the incentive is based on CIC (Certificati di immissioni in consumo). One CIC is equivalent to 10 Gcal (41.86 GJ) of renewable energy produced, the value of a CIC is around 500 € [22], the value of CIC is doblet if biogas come from waste of "Table 1A of D.M. 6 Luglio 2012" [19].

## 3.5.2 SOFC-based CHP system

“D.M. 12 Luglio 2012” [19] and “D.M. 23 Giugno 2016” [23] are the reference laws for public incentive to plant that generates electric power from renewable sources different from solar photovoltaic like the studied case of the combined heat and power plant based on SOFC and fed by biogas. The decree subdivides the biomasses based on the origin into four categories with different subsidies, only two types of biomass are considered in the analysis:

- Biological origins products (a)
- Biological origins by-products (b) (“Table 1-A DM 12/07/12”)

The regulation prescribes a base incentive for each type of power plant based on the renewable sources used, the technology involved and the size of the system defined as the nominal power capacity. The incentive is expressed in euro for each megawatt hour sold to the national electric grid by the plant.

The law introduces also some type of prize incentive for a cogenerative plant with high efficiency fed by biomass, biogas or bioliquid:

- 30 €/MWh if the plant respects the limit of emission of polluting substance in the atmosphere reported in the same decree
- 40 €/MWh for CHP plant fed by biomass of the category a
- 40 €/MWh for CHP plant fed by biomass of the category b if the recovered heat is used for district heating network
- 10 €/MWh in the other cases

$$PS_{final} = PS_{base} + PR \tag{40}$$

The *equation (*40) shows the formula to calculate the public subsidies according to the ministerial decree: $PS_{final}$ is the public subsidies recognized to the plant, $PS_{base}$ is the base incentive for the type and size of the system and PR is the value of the sum of the prize. The price of electricity is available on the website of GME [21].

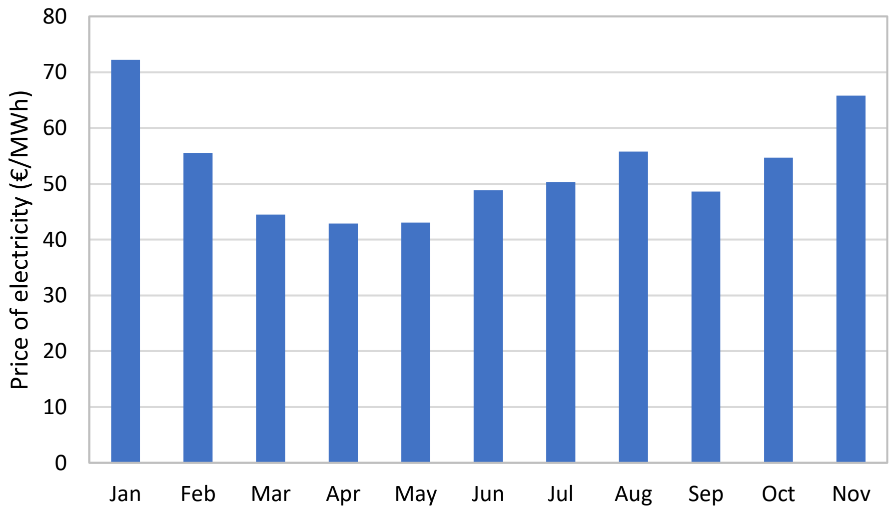


*Figure 9: Market price of electricity in Italy 2017* [21]

# 4 Energy analysis results

## 4.1 Upgrade to biomethane

The principal resource of the upgrading plant is the stream of biogas entering the clean-up bed with a flow of 120 $Sm^3/h$ corresponding to a chemical power of 680 kW based on LHV of methane (35.7 $MJ/Nm^3$). The other fuel of the system is the electrical power consumed by compressors and pump, the value of the power needed for these components are taken by AspenPlus® simulation:

- Biogas compressor K1 $W_{k1}$ = 18.32 kW
- Air compressor K2 $W_{k2}$ = 0.66 kW
- Water pump PUMP $W_{pump}$ = 13.47 kW

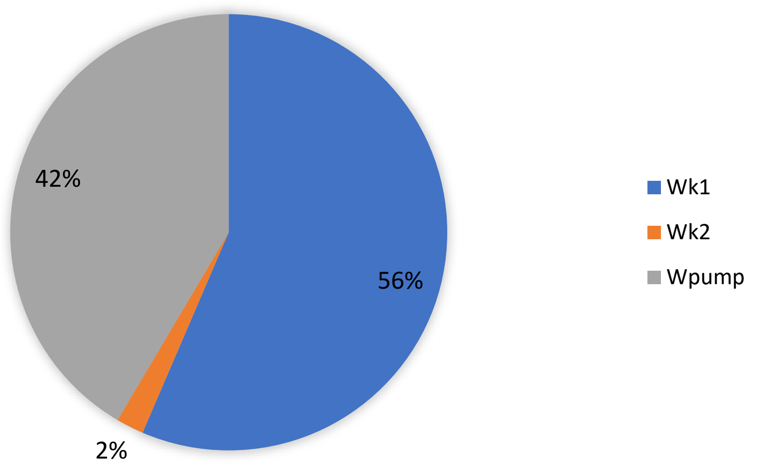


*Figure 10: Upgrade to biomethane: auxiliary power consumption*

The most consuming components are the biogas compressor K1 and the water pump PUMP (*Figure 10*) because they are responsible for the highest pressurization and they elaborate more mass flow, K2 is necessary only for circulation of stripping air and it needs less electric power to work.

The product of the studied plant is the biogas upgraded, so the stream of biomethane exiting DRY2 (*biomtn1*). The volumetric flow of stream *biomtn1* is 7.63 $m^3/h$ at 10 bar and 25°C that correspond to 73.7 $Sm^3/h$, the concentration of methane in the mixture 97.5% vol. based on LHV of biogas the power produce is 677 kW. Considering productive structure (*Table 5*) of fuel and product of the system the global energy efficiency of the plant, *equation*(2) is 95%.

Upgrading plant for biogas based on pressurized water scrubbing in choose the operative condition of pressure and mass flow realized a good concentration of $CH_4$ in the biomethane produced (97.5% vol.) and low losses (0.22%).

## 4.2 SOFC-based CHP system

The simulation of the system in AspenPlus® provides the electrical power absorbed by the blowers:

- FUEL-CMP = 0.75 kW
- AIR-CMP = 7.34 kW
- REC-CMP = 1.07 kW

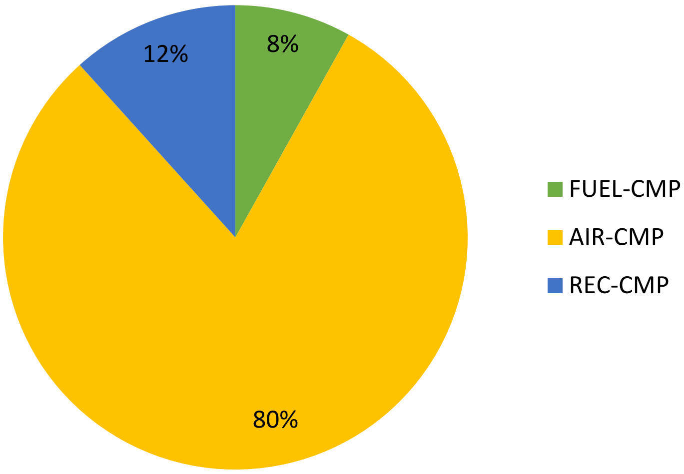


*Figure 11: SOFC-based CHP system: auxiliary power consumption*

The principal consumption of electrical power takes place in the air blower AIR-CMP (see *Figure 11*) because it treats a high flow (0.55 $m^3/s$) compared to the other blowers: FUEL-CMP (0.034 $m^3/s$) and REC-CMP (0.31 $m^3/s$).

The SOFC produces 440 kW of electric power (*Wel*), considering the efficiency of the inverter 95% and the blower consumption the net power produced by the CHP plant based on SOFC fed by biogas is equal to 408.8 kW with a net electrical efficiency, *equation* (*3*), of the plant equal to 60.7%.

Thermal power recovered by the CHP-HX from the exhaust to heat up water for the thermal user is 186.7 kW with a thermal efficiency, *equation* (*4*), of the plant equal to 27.4%. The combined heat and power global efficiency is 88.1% referred to the LHV of the biogas.

# 5 Exergy analysis results

## 5.1 Upgrade to biomethane

For each stream of mass, work and heat present in the plant the specific physical and chemical exergy has been calculated according to exergy analysis methodology and the reference environment described in *chapter 3* "Methodology", the results are summarized in *Table 8*.

| | phisical exergy (kJ/kg) | chemical exegy (kJ/kg) | mass flow (kg/s) | total exergy flow (kW) |
|---|---|---|---|---|
| air0 | 1.07 | 3.44 | 0.015 | 0.07 |
| air1 | 36.96 | 3.44 | 0.015 | 0.59 |
| biogas0 | 0.09 | 18508 | 0.038 | 712 |
| biogas1 | 0.09 | 18508 | 0.038 | 712 |
| biogas2 | 0.97 | 12369 | 0.061 | 752 |
| biogas3 | 259.09 | 12369 | 0.061 | 768 |
| biogas4 | 183.29 | 12370 | 0.061 | 763 |
| biomtn1 | 330.98 | 48173 | 0.015 | 708 |
| biomtn2 | 331.11 | 48371 | 0.015 | 708 |
| gas-out | 0.83 | 257 | 0.039 | 10.11 |
| gasrec1 | 62.99 | 1823 | 0.022 | 42.39 |
| gasrec2 | 0.66 | 1823 | 0.022 | 40.99 |
| gasrec3 | 0.73 | 1834 | 0.022 | 41.02 |
| liquid1 | 1.47 | 7.86 | 6.781 | 63.30 |
| liquid2 | 0.94 | 1.77 | 6.758 | 18.30 |
| liquid3 | 0.51 | 1.77 | 6.758 | 15.41 |
| watrec | 0.22 | 0 | 6.734 | 1.51 |
| w1 | 0.26 | 0 | 1.11E-04 | 0 |
| w2 | 1.51 | 0 | 5.91E-05 | 0 |
| water0 | 0.17 | 0 | 6.28E-04 | 0 |
| water1 | 0.17 | 0 | 6.734 | 1.14 |
| water2 | 1.61 | 0 | 6.734 | 10.85 |
| Wk1 | - | - | - | 18.32 |
| Qcool | - | - | - | 0.00 |
| Wk2 | - | - | - | 0.66 |
| Wpump | - | - | - | 13.47 |
| Qdry1 | - | - | - | 0 |
| Qdry2 | - | - | - | 0 |

*Table 8: Upgrade to biomethane: exergy*

## 5.1.1 Exergy lost

Applying the equation of balance of exergy in matrix formulation, *equation* (15), the results are the exergy lost *I* and exergy destruction rate ψ following the definition, *equation* (*16)*, for each component. The incidence matrix of the biogas upgrading plant is reported in the appendix (*Table 36*):

| | **I (kW)** | ***ψ*** |
|---|---|---|
| CLEANING | 0.00 | 0.000 |
| MIX1 | 0.76 | 0.001 |
| K1 | 2.61 | 0.143 |
| COOLER | 4.60 | 0.006 |
| PUMP | 3.76 | 0.279 |
| SCRUBBER | 2.99 | 0.004 |
| FLASH | 2.60 | 0.041 |
| V1 | 1.40 | 0.033 |
| V2 | 2.89 | 0.158 |
| K2 | 0.13 | 0.203 |
| STRIPPER | 4.38 | 0.274 |
| MIX2 | 0.37 | 0.245 |
| total | 26.51 | 0.036 |

*Table 9:Upgrade to biomethane: exergy lost*

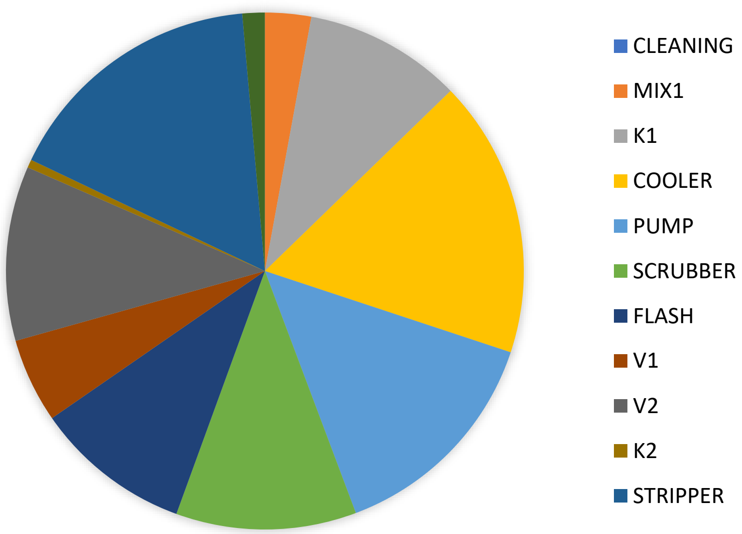


*Figure 12: Upgrade to biomethane: exergy lost*

The plant is characterized by a low exergy destruction and a negligible exergy destruction rate (only 0.036). The components with the highest irreversibility are COOLER, which is a dissipative element like valves V1 and V2, and STRIPPER. In the SCRUBBER the exergy lost is near 3 kW but with a very low destruction rate (0.004).

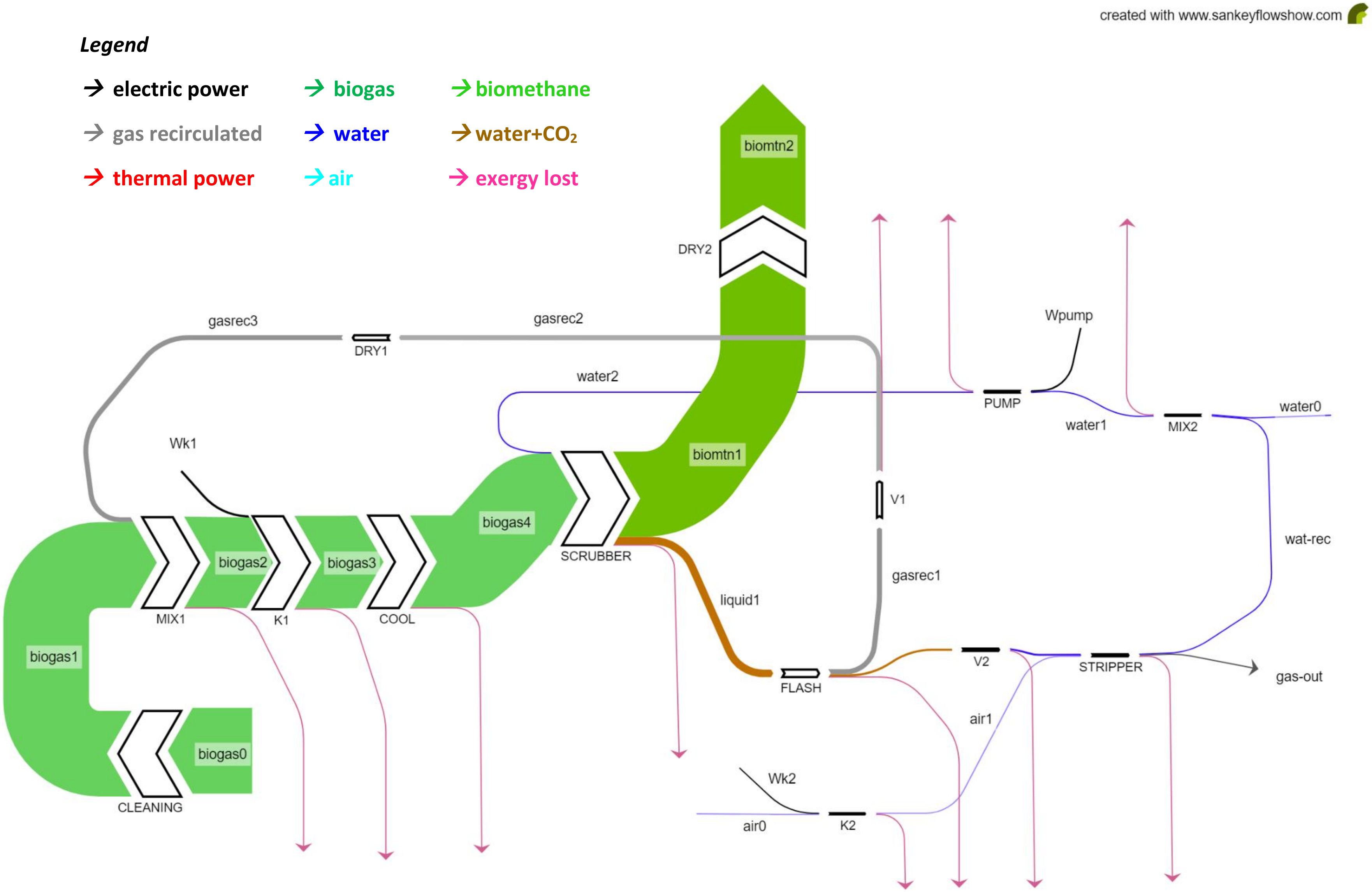


*Figure 13: Upgrade to biomethane: Sankey diagram*

*Figure 13* shows the Sankey diagram of the upgrading plant, the main exergy flow is due to the presence of methane in the mixture, so the components that elaborate the highest exergy flow are the CLEANING, the MIXER, the biogas compressor K1, the SCRUBBER and the DRY2; all these devices have low exergy destruction rate. The recirculation loop, constituted by FLASH vessel, valve V1 and the DRY1, have a lower but not negligible exergy flow due to the presence of methane in low concentration il *liquid1* and in the *gasrec.* The other elements of the system elaborate a negligible exergy compare to the main components, the streams are composed essentially by water, carbon dioxide and air which have a little exergy content, the scope of this part of the system is only to regenerate the processing water.

## 5.1.2 Exergy efficiency

According to the productive structure described in *Table 5* is possible to calculate the efficiency of second principle ε, see *equation (17)*, for the main components of the plant. The devices which elaborate high flow of exergy are the most important in the system and they have high exergy factor *λ,* see *equation (18).* The results are shown in *Table 10* and *Figure 14.*

| | $E_{fuel}$ (kW) | $E_{product}$ (kW) | ε | λ |
|---|---|---|---|---|
| CLEANING | 712.51 | 712.51 | 100.0% | 0.956 |
| MIX1 | 753.53 | 752.77 | 99.9% | 1.011 |
| K1 | 18.32 | 15.71 | 85.7% | 0.025 |
| COOL | 768.47 | 763.87 | 99.4% | 1.031 |
| PUMP | 13.47 | 9.71 | 72.1% | 0.018 |
| SCRUB | 774.72 | 771.73 | 99.6% | 1.040 |
| FLASH | 63.30 | 60.70 | 95.9% | 0.085 |
| V1 | 42.39 | 40.99 | 96.7% | 0.057 |
| V2 | 18.30 | 15.41 | 84.2% | 0.025 |
| K2 | 0.66 | 0.53 | 79.7% | 0.001 |
| STRIP | 16.00 | 1.51 | 9.5% | 0.021 |
| MIX2 | 1.51 | 1.14 | 75.5% | 0.002 |
| global | 745.02 | 708.45 | 95.1% | 1.000 |

*Table 10: Upgrade to biomethane: exergy efficiency*

The most important components of the plant, based on exergy flow elaborated, are SCRUBBER and COOLER that have an exergy factor higher than one due to the recirculation of gas from FLASH vessel; exergy efficiency of SCRUBBER is very high (99,6 %) because in this component there are not energy transformation: inlet energy is chemical energy of methane present in biogas and exiting energy is the same chemical energy. Also the COOLER has high efficiency unless it is a dissipative component, the thermal energy of the stream after the compression stage must be rejected in the environment. The other equipment has a lower efficiency, but they elaborate a low quantity of exergy so their contribution to global efficiency is small, for this reason, the exergy efficiency of the whole plant is 95.1%. The worst component is STRIPPER (9.5%), the low efficiency is the result of the loss of exergy due to release in the atmosphere of *gas-out* with chemical exergy flow of the $CO_2$ contained.

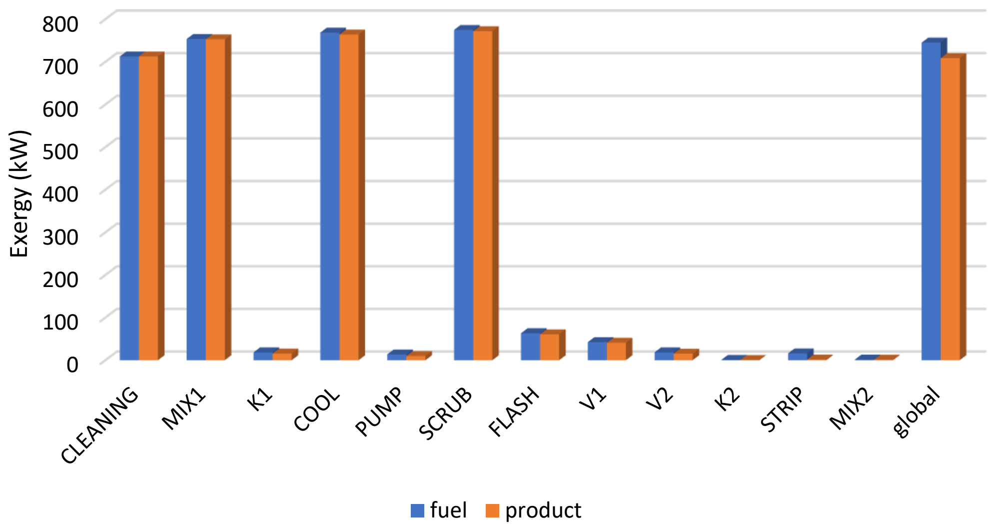


*Figure 14: Upgrade to biomethane: exergy efficiency*

### 5.1.3 Exergy cost

The upgrading plant is constituted by 14 components and 28 streams (mass, work and heat), so the incidence matrix $\bar{\bar{A}}$ (*Table 36* of the appendix) has dimensions (14x28), for this reason to solve the system of exergy cost, *equation* (20), is necessary to add 14 auxiliary equations following the rules described in chapter 3.3.3:

- rule 1 is applied to the stream *air0, biogas0, Wk1, Wk2, Wpump* and *water0*. Each stream has a specific exergy cost k*=1 because they are the external fuel of the plant:
  a. $E^*_{air0} = E_{air0}$
  b. $E^*_{biogas0} = E_{biogas0}$
  c. $E^*_{Wk1} = E_{Wk1}$
  d. $E^*_{Wk2} = E_{Wk2}$
  e. $E^*_{Wpump} = E_{Wpump}$
  f. $E^*_{water0} = E_{water0}$
- rule 2 is applied to streams that represent a loss for the system, so the streams *Qcool, Qdry1, w1, Qdry2, w2* and *gas-out*, for these streams the exergy cost is zero:
  g. $E^*_{Qcool} = 0$
  h. $E^*_{Qdry1} = 0$
  i. $E^*_{w1} = 0$
  j. $E^*_{Qdry2} = 0$
  k. $E^*_{w2} = 0$
  l. $E^*_{gas-out} = 0$
- rule 4 is applied to SCRUBBER where *liquid1*and biomtn1 have the same specific exergy cost and to FLASH tank to streams *gasrec1* and *liquid2.*
  m. $\left({E_{liquid1}}/{E_{biomtn1}}\right) \cdot E^*_{biomtn1} - E^*_{liquid1} = 0$
  n. $\left({E_{liquid2}}/{E_{gasrec1}}\right) \cdot E^*_{gasrec1} - E^*_{liquid2} = 0$

With these external hypotheses is possible to build the cost matrix $\overline{\overline{Ac}}$ for the upgrading plant show in *Table 38* of the appendix and to resolve the system of *equation* (21), the auxiliary equation in the cost matrix are reported in the row with the corresponding letter. After the definition of cost matrix, it is possible to calculate the exergy cost E* and the specific exergy cost k* for each stream, the results are reported in *Table 11*.

| | **E* (kW)** | **E (kW)** | **k*** |
|---|---|---|---|
| air0 | 0.07 | 0.07 | 1.00 |
| air1 | 0.73 | 0,.59 | 1.23 |
| biogas0 | 712 | 712 | 1.00 |
| biogas1 | 712 | 712 | 1.00 |
| biogas2 | 759 | 752 | 1.01 |
| biogas3 | 777 | 768 | 1.01 |
| biogas4 | 777 | 763 | 1.02 |
| biomtn1 | 745 | 708 | 1.05 |
| biomtn2 | 745 | 708 | 1.05 |
| gas-out | 0 | 10.11 | 0 |
| gasrec1 | 46.49 | 42.39 | 1.10 |
| gasrec2 | 46.49 | 40.99 | 1.13 |
| gasrec3 | 46.49 | 41.02 | 1.13 |
| liquid1 | 66.57 | 63.30 | 1.05 |
| liquid2 | 20.07 | 18.30 | 1.10 |
| liquid3 | 20.07 | 15.41 | 1.30 |
| watrec | 20.80 | 1.51 | 13.74 |
| w1 | 0 | 0 | 0 |
| w2 | 0 | 0 | 0 |
| water0 | 0 | 0 | 1.00 |
| water1 | 20.80 | 1.14 | 18.19 |
| water2 | 34.27 | 10.85 | 3.16 |
| Wk1 | 18.32 | 18.32 | 1.00 |
| Qcool | 0 | 0 | 0 |
| Wk2 | 0.66 | 0.66 | 1.00 |
| Wpump | 13.47 | 13.47 | 1.00 |
| Qdry1 | 0 | 0 | 0 |
| Qdry2 | 0 | 0 | 0 |

*Table 11: Upgrade to biomethane: exergy cost*

## 5.2 SOFC-based CHP system

The reference environment for the exergy analysis is the one defined in *chapter 3.2* composed of atmospheric air and liquid water. Gases are considered ideal gas, so physical and chemical exergy are calculated following this hypothesis, for the chemical exergy the presence of $NO_x$ in the flue gas is neglected because its concentration is very low and do not influence the global value.The value of physical and chemical exergy for each stream are reported in *Table 12* with the total exergy flow.

| | **phisycal exergy (kJ/kg)** | **chemical exergy (kJ/kg)** | **mass flow (kg/s)** | **total exergy flow (kW)** |
|---|---|---|---|---|
| biogas0 | -1.10 | 18591 | 0.038 | 707 |
| biogas1 | 12.97 | 18591 | 0.038 | 707 |
| biogas2 | 8.28 | 18591 | 0.038 | 707 |
| biogas3 | 574.97 | 18591 | 0.038 | 729 |
| biogas4 | 609.24 | 6390 | 0.139 | 975 |
| biogas5 | 609.24 | 6390 | 0.070 | 491 |
| biogas6 | 609.24 | 6390 | 0.069 | 483 |
| bioref | 644.82 | 7337 | 0.070 | 560 |
| an-fuel | 629.48 | 6851 | 0.139 | 1042 |
| air0 | -1.06 | 1.65 | 0.661 | 0.39 |
| air1 | 6.95 | 1.65 | 0.661 | 5.68 |
| air2 | 320.07 | 1.65 | 0.661 | 212 |
| cat-ex | 441.09 | 2.42 | 0.615 | 272 |
| an-ex | 615.19 | 1853 | 0.185 | 456 |
| burn-fuel | 615.19 | 1853 | 0.084 | 206 |
| an-rec1 | 615.19 | 1853 | 0.101 | 250 |
| an-rec2 | 624.18 | 1853 | 0.101 | 251 |
| water | -0.17 | 0 | 0.010 | 0 |
| exhaust1 | 556.05 | 26.64 | 0.709 | 413 |
| exhaust2 | 509.13 | 26.64 | 0.709 | 380 |
| exhaust3 | 106.60 | 26.64 | 0.709 | 94.5 |
| exhaust4 | 6.48 | 26.64 | 0.709 | 23.5 |
| wat-in | 13.49 | 0 | 4.059 | 54.8 |
| wat-out | 20.66 | 0 | 4.059 | 83.9 |
| Wfuelcmp | - | - | - | 0.74 |
| Waircmp | - | - | - | 7.33 |
| Wreccmp | - | - | - | 1.07 |
| Wel | - | - | - | 440 |
| Qsofc | - | - | - | 76.8 |
| Qref | - | - | - | 78.1 |

*Table 12: SOFC-based CHP system: exergy*

## 5.2.1 Exergy lost

Applying the equation of exergy lost, *equation (15)*, the results are the exergy lost *I* and exergy destruction rate *ψ, see equation (16)*, for each component, the results are reported in *Table 13* and *Figure 15*:

| | ***I* (kW)** | ***ψ*** |
|---|---|---|
| FUEL-CMP | 0.20 | 0.277 |
| CLEANING | 0.18 | 0.000 |
| FUEL-HX | 11.73 | 0.352 |
| AN-MIX | 4.65 | 0.005 |
| REF-SPLIT | 0 | 0.000 |
| EXT-REF | 9.10 | 0.016 |
| REF-MIX | 1.94 | 0.002 |
| SOFC | 8.90 | 0.007 |
| AIR-CMP | 2.04 | 0.278 |
| AIR-HX | 78.59 | 0.275 |
| AN-SPLIT | 0 | 0.000 |
| REC-CMP | 0.16 | 0.148 |
| BURNER | 66.07 | 0.138 |
| CHP-HX | 41.95 | 0.454 |
| total | 225.50 | 0.316 |

*Table 13: SOFC-based CHP system: exergy lost*

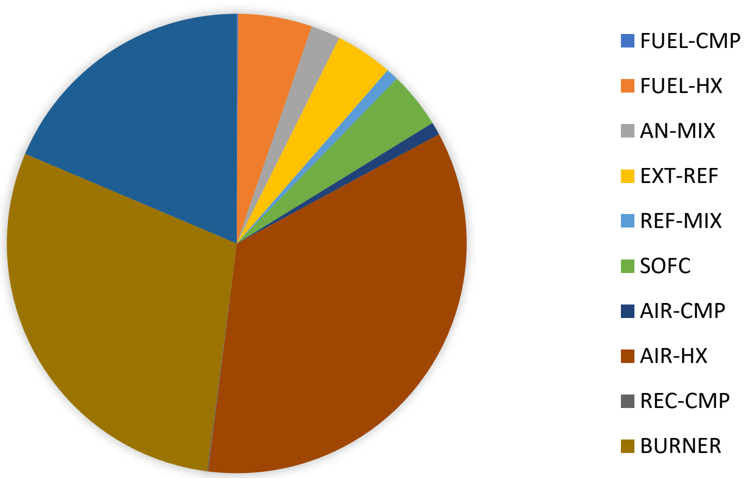


*Figure 15: SOFC-based CHP system: exergy lost*

The components responsible for the highest irreversibility generation are the heat exchanger especially the AIR-HX and CHP-HX which are characterized by a large amount of thermal power exchange. Another component with high exergy destruction is the BURNER because the chemical exergy of the fuels is converted in thermal exergy. The SOFC has a low exergy loss because electric power is produced directly from the electrochemical reaction occurring at electrodes, the source of irreversibility in this component is the internal over-voltage.

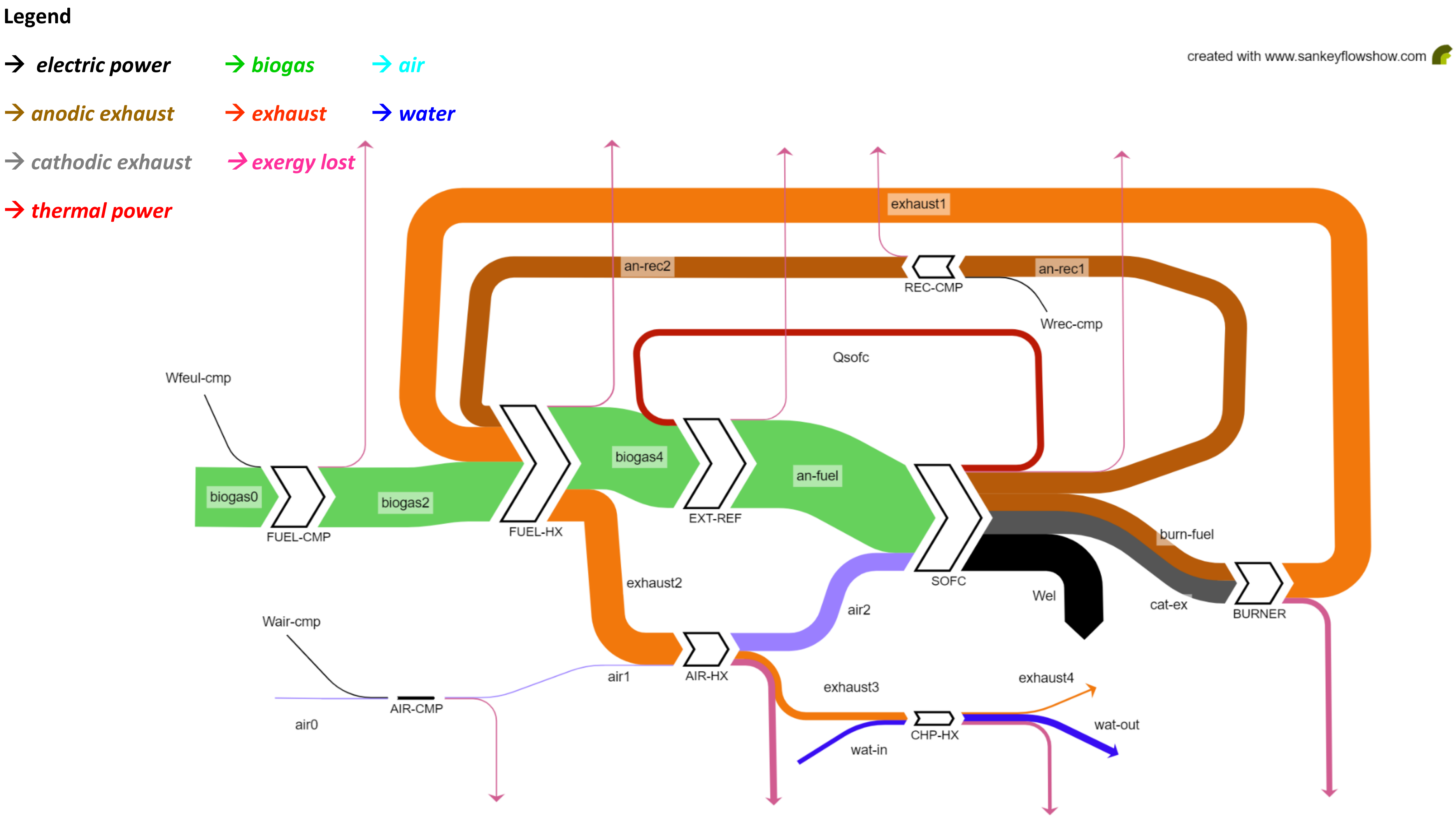


*Figure 16: SOFC-based CHP system: Sankey diagram*

*Figure 16* reports the Sankey diagram for the SOFC-based CHP system, the comparison with the same diagram of the upgrading plant underlines how the SOFC system is more complex from the point of view of exergy flow due to the recirculation of the anodic exhaust (*an-rec1*) and the use of burner exhaust to pre-heat fuel and air streams. The components which elaborate the highest flux of exergy are the FUEL-HX and the SOFC, the amount of exergy is due to the chemical contribution of the biogas and the thermal exergy of hot gas. The only component with a negligible exergy flow is the AIR-CMP because it elaborates a stream of cold air causing a small pressure increase. REC-CMP and BURNER have a quite high exergy flow due to the high temperature of the SOFC products and the presence in the anodic exhaust of $H_2$ and CO that are the products of reforming of biogas not used in the cell (fuel utilization 90%). The exergy flux in the heat exchangers AIR-HX and CHP-HX is due only to the high temperature of the burner exhaust.

### 5.2.2 Exergy efficiency

According to the productive structure of the power plant, reported in *Table 6,* is possible to calculate the exergy efficiency ε *(17)* and exergy factor *λ (18)* of every component.

| | $E_{fuel}$ (kW) | $E_{product}$ (kW) | ε | λ |
|---|---|---|---|---|
| FUEL-CMP | 0.74 | 0.54 | 72.3% | 0.001 |
| CLEANING | 705 | 705 | 100.0% | 0.988 |
| FUEL-HX | 33.3 | 21.6 | 64.8% | 0.047 |
| AN-MIX | 976 | 971 | 99.5% | 1.366 |
| REF-SPLIT | 971 | 971 | 100.0% | 1.360 |
| EXT-REF | 567 | 559 | 98.6% | 0.795 |
| REF-MIX | 1041 | 1039 | 99.8% | 1.458 |
| SOFC | 1253 | 1243 | 99.2% | 1.754 |
| AIR-CMP | 7.33 | 5.30 | 72.2% | 0.010 |
| AIR-HX | 285 | 207 | 72.5% | 0.400 |
| AN-SPLIT | 452 | 452 | 100.0% | 0.634 |
| REC-CMP | 1.07 | 0.91 | 85.2% | 0.001 |
| BURNER | 478 | 411 | 86.0% | 0.670 |
| CHP-HX | 92.5 | 29.1 | 31.5% | 0.129 |
| global | 714 | 437 | 61.2% | 1.000 |

*Table 14: SOFC-based CHP system: exergy efficiency*

The most important device of the plant is SOFC that is characterized by very high exergy efficiency, close to 100%, the exergy factor is higher than one due to the partial anodic recirculation. The worst component is CHP-HX because the exhaust is released into the atmosphere at 90°C, so with a not negligible thermal exergy content, and it is considered a loss for the plant. The efficiency of the heat exchangers depends on the temperature difference between cold and hot fluids, FUEL-HX has the largest logarithmic mean temperature difference so it has the lowest efficiency. The importance of a component of the plant is determined by the amount of exergy elaborated, this parameter is shown in *Figure 17.*

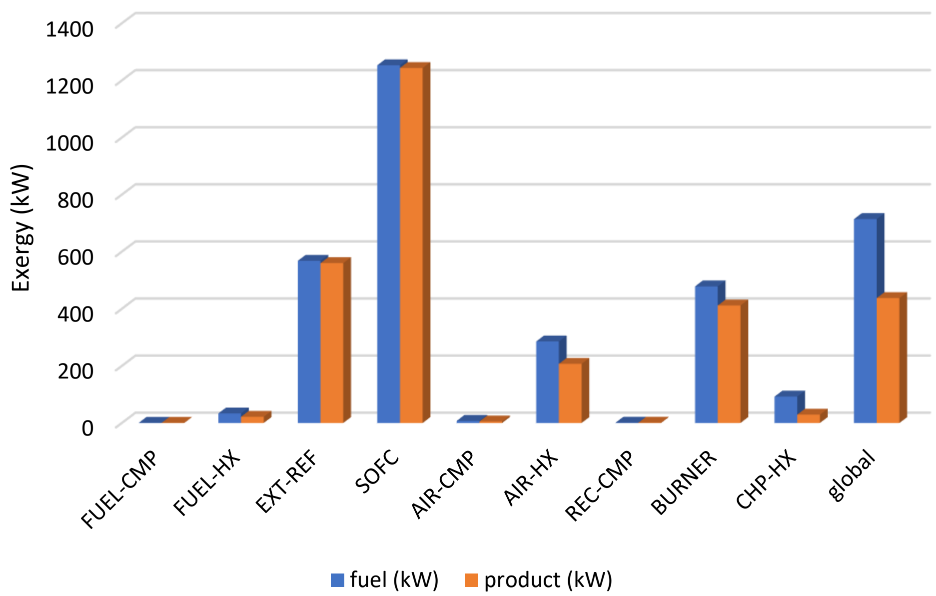


*Figure 17: SOFC-based CHP system: exergy efficiency*

## 5.2.3 Exergy cost

The analyzed SOFC plant is defined by 30 streams of mass, work and heat and by 14 components. To calculate the exergy cost E* and specific exergy cost k* of each stream, 16 auxiliary equation are needed, the four type of external hypotheses are described in chapter *3.3.3*:

- Rule 1 is applied to the fuels of the plant the specific exergy cost k* is imposed equal to 1; k* for the electric power absorbed by blowers (*Wfuel-cmp, Wair-cmp* and *Wrec-cmp*) is considered equal to k* of the power produced by SOFC (*Wel*) because they are internal consume.
    a. $E^*_{biogas0} = E_{biogas0}$
    b. $E^*_{air0} = E_{air0}$
    c. $E^*_{water} = E_{water}$
    d. $E^*_{wat-in} = E_{wat-in}$
    e. $\left({E_{Wel}}/{E_{Wfuelcmp}}\right) \cdot E^*_{Wfuelcmp} - E^*_{Wel} = 0$
    f. $\left({E_{Wel}}/{E_{Waircmp}}\right) \cdot E^*_{Waircmp} - E^*_{Wel} = 0$
    g. $\left({E_{Wel}}/{E_{Wreccmp}}\right) \cdot E^*_{Wreccmp} - E^*_{Wel} = 0$

- Rule 2 is applied to the only loss in the system that is *exhaust4*, for this stream the exergy cost is zero.
    h. $E^*_{exhaust4} = 0$

- Ruel 3 is applied to FUEL-HX: k* of *exhaust2* equal to k* of *exhaust1*; AIR-HX: k* of *exhaust3* equal to k* of *exhaust2.*
    i. $\left({E_{exhaust2}}/{E_{exhaust1}}\right) \cdot E^*_{exhaust1} - E^*_{exhaust2} = 0$
    j. $\left({E_{exhaust3}}/{E_{biogas5}}\right) \cdot E^*_{biogas5} - E^*_{biogas6} = 0$

- Rule 4 is applied to REF-SPLIT: *biogas5* and *biogas6* have the same specific exergy cost; SOFC: *Wel*, *Qsofc*, *an-ex* and *cat-ex* have the same k*; AN-SPLIT: k* of *an-rec1* is equal to k* of *burn-fuel.*
    k. $\left({E_{Wel}}/{E_{cat-ex}}\right) \cdot E^*_{cat-ex} - E^*_{Wel} = 0$
    l. $\left({E_{Qsofc}}/{E_{Wel}}\right) \cdot E^*_{Wel} - E^*_{Qsofc} = 0$
    m. $\left({E_{an-rec}}/{E_{burn-fuel}}\right) \cdot E^*_{burn-fuel} - E^*_{an-rec} = 0$
    n. $\left({E_{biogas6}}/{E_{biogas5}}\right) \cdot E^*_{biogas5} - E^*_{biogas6} = 0$
    o. $\left({E_{Wel}}/{E_{an-ex}}\right) \cdot E^*_{an-ex} - E^*_{Wel} = 0$

- The last external hypothesis is related to the heat flux between SOFC and EXT-REF: *Qsofc* and *Qref* have the same exergy cost E* because they are the same stream.
    p. $E^*_{Qsofc} = E^*_{Qref}$

With these external hypotheses is possible to write the cost matrix shown in *Table 41* of the appendix for the system and calculate the exergy cost *E** and the specific exergy cost *k** for every stream. The results of the computation are reported in *Table 15.*

| | ***E**** (kW) | ***E*** (kW) | ***k**** |
|---|---|---|---|
| biogas0 | 707 | 705 | 1.00 |
| biogas1 | 708 | 705 | 1.00 |
| biogas2 | 708 | 705 | 1.00 |
| biogas3 | 758 | 727 | 1.04 |
| biogas4 | 1087 | 971 | 1.12 |
| biogas5 | 548 | 489 | 1.12 |
| biogas6 | 539 | 481 | 1.12 |
| bioref | 648 | 559 | 1.16 |
| an-fuel | 1188 | 1039 | 1.14 |
| air0 | 0.39 | 1.09 | 0.36 |
| air1 | 9.99 | 6.39 | 1.56 |
| air2 | 443 | 213 | 2.08 |
| cat-ex | 357 | 273 | 1.31 |
| an-ex | 597 | 452 | 1.32 |
| burn-fuel | 270.35 | 204 | 1.32 |
| an-rec1 | 327 | 248 | 1.32 |
| an-rec2 | 328 | 248 | 1.32 |
| water | 0 | 0 | 1.00 |
| exhaust1 | 627 | 411 | 1.53 |
| exhaust2 | 577 | 378 | 1.53 |
| exhaust3 | 143 | 92 | 1.55 |
| exhaust4 | 0 | 21.43 | 0 |
| wat-in | 54 | 54 | 1.00 |
| wat-out | 198 | 83 | 2.36 |
| Wfuelcmp | 0.97 | 0.74 | 1.31 |
| Waircmp | 9.60 | 7.33 | 1.31 |
| Wreccmp | 1.40 | 1.07 | 1.31 |
| Wel | 576 | 440 | 1.31 |
| Qsofc | 100 | 77 | 1.31 |
| Qref | 100 | 78 | 1.29 |

*Table 15: SOFC-based CHP system: exergy cost*

# 6 Exergo-economic analysis results

## 6.1 Upgrade to biomethane

The exergo-economic analysis starts from the sizing of all components of the plant studied, for pumps and compressor the power needed is determined by AspenPlus® simulation:

- Biogas compressor K1 = 18.32 kW
- Air compressor K2 = 0.66 kW
- Water pump PUMP = 13.47 kW

For scrubber and stripper vessels the dimensions are taken from the master thesis of Paolo Rotunno "*Analisi Tecno-Economica Della Produzione e Distribuzione di Biometano*"[6]:

- Scrubber diameter 0.8 m height 10 m and volume 5 $m^3$
- Stripper diameter 0.5 m height 5 m and volume 1 $m^3$

The flash vessel is designed based on the technical paper of "*Sprirax Sarco*" [24] in function of the pressure operation and mass flow: diameter 0.457 m height 1.521 m and volume 0.25 $m^3$.

The size of cooler is 10 $m^2$ that correspond to the lower limit of validity for the relative cost function, the dryers are based on silica gel, for these components the price is taken directly from the technical paper of "*Deltaadsorber*" [25] in function of the volumetric flow. For clean-up system the investment cost is referred to [13] applying a scaling factor equal to 0.6:

$$cost(Q_r) = cost(Q_0) \cdot \left(\frac{Q_r}{Q_0}\right)^{0.6} \tag{41}$$

where $Q_r$ is the real volumetric flow of biogas and $Q_0$ is the reference volumetric flow.

In this analysis components like mixer and valve are considered whit a price equal to zero. The constants and the sizing parameter necessary to calculate the $C_p^0$ for each part with cost function *(28)* are shown in *Table 16*.

| | **k1** | **k2** | **k3** | **A** | **$C_p^0$ ($)** |
|---|---|---|---|---|---|
| K1 | 2.2897 | 1.3604 | -0.1027 | 18.32 kW | 6,981 |
| COOL | 4.0336 | 0.2341 | 0.0497 | 10 $m^2$ | 20,768 |
| PUMP | 3.3892 | 0.0536 | 0.1538 | 13.47 kw | 4,424 |
| SCRUBBER | 3.4974 | 0.4485 | 0.1074 | 5 $m^3$ | 7,300 |
| SCRUB. FILL. | | | | 5 $m^3$ | 1,192 |
| FLASH | 3.4974 | 0.4485 | 0.1074 | 0.25 $m^3$ | 1,846 |
| K2 | 2.2897 | 1.3604 | -0.1027 | 0.66 kw | 110 |
| STRIPPER | 3.4974 | 0.4485 | 0.1074 | 1 $m^2$ | 3,143 |
| STRIP. FILL | | | | 1 $m^2$ | 241 |

*Table 16: Upgrade to biomethane: cost functions*

Considering pressure factor for scrubber equal to 1.25, see *equation (30)*, and 1 for all the other components. The material factor is different from one only for SCRUBBER, FLASH vessel and STRIPPER, for this equipment the material choose is carbon steel whit stainless steel cladding to protect the internal surface from acid solution due to the presence of water and $CO_2$, for this configuration the material factor is equal to 2.8.

The equation to calculate bare module factor for heat exchangers vessels and pumps is:

$$F_{bm} = B_1 + B_2 \cdot F_p \cdot F_m \tag{42}$$

| | B1 | B2 | $F_p$ | $F_m$ | $F_{bm}$ |
|---|---|---|---|---|---|
| COOL | 0.96 | 1.21 | 1 | 1 | 2.17 |
| SCRUBBER | 2.25 | 1.82 | 1.25 | 2.8 | 8.63 |
| FLASH | 2.25 | 1.82 | 1 | 2.8 | 7.35 |
| STRIPPER | 2.25 | 1.82 | 1 | 2.8 | 7.35 |
| PUMP | 1.89 | 1.35 | 1 | 1 | 3.24 |

*Table 17: Upgrade to biomethane: bare module factor*

*Table 17* shows constants and result of the calculation of the bare module factor for components of the plant, for compressors the bare module factor is equal to 2.7.

The cost of elements calculated with cost function is valid for 2001 so it is necessary to scale this price in 2017 value, the ratio between 2001 and 2017 money can be expressed by the ratio between the CEPCI (Chemical Engineering Plant Index) indexes in the two years.

$$C_{2017} = C_{2001} \cdot \frac{CEPCI_{2017}}{CEPCI_{2001}} \tag{43}$$

Reference [15] report the CEPCI for 2001 equal to 397, the value of CEPCI for equipment in 2017 is 672 [26]. With these hypotheses the bare erected cost for each component are:

| | CLEAN-UP | K1 | COOL | PUMP1 | SCRUBBER | DRY2 | FLASH | DRY1 | K2 | STRIPPER |
|---|---|---|---|---|---|---|---|---|---|---|
| **$C_{bec}$ ($)** | 17,1540 | 18,851 | 4,5067 | 14,336 | 64,209 | 720 | 13,563 | 687 | 298 | 23,333 |

*Table 18: Upgrade to biomethane: bare erected cost*

Following the NETL/DoE cost estimation methodology [16] [27] TASC (total as-spent capital) can be calculated as the sum of various cost:

- EPCC = 10% of BEC
- process contingency = 30% of BEC (small pilot plant)
- project contingency = 15% of BEC + EPCC + Process contingency
- start up = 2% of TPC
- inventory = 0.5% of TPC
- financing cost = 2.7% of TPC
- other owners cost = 15% of TPC
- financial cost =10% of TOC

applying the euro/dollar change equal to 1.15 €/$ [28] the result is the TASC form each component in euro reported in *Table 19*.

| | CLEAN-UP | K1 | COOL | PUMP1 | SCRUBBER | DRY2 | FLASH | DRY1 | K2 | STRIPPER |
|---|---|---|---|---|---|---|---|---|---|---|
| **TASC (€)** | 200,702 | 79,475 | 189,997 | 60,439 | 270,696 | 1,793 | 57,180 | 1,711 | 1,256 | 98,368 |

*Table 19: Upgrade to biomethane: TASC*

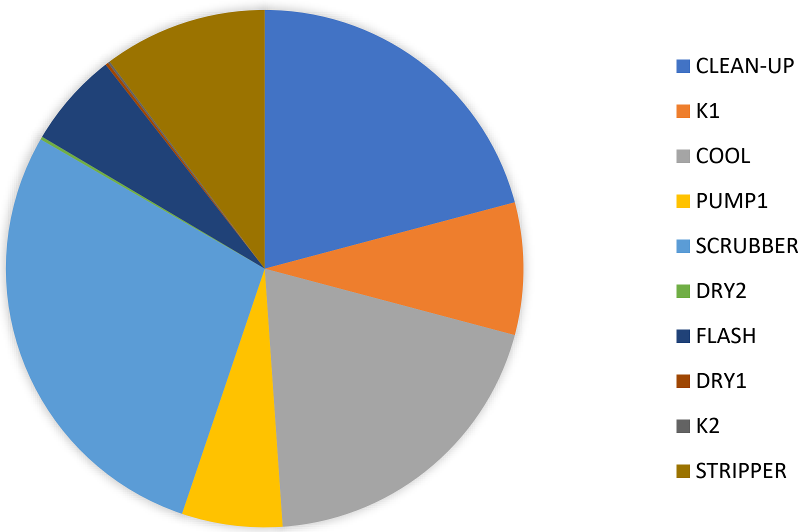


*Figure 18: Upgrade to biomethane: cost of components*

The total cost of biogas upgrading plant based on pressurized water scrubbing that can elaborate 120 $Sm^2/h$ of biogas and produce 73.7 $Sm^3/h$ of biomethane is less than one million euro, around 962,000 €. The most expensive component is the SCRUBBER vessel that is the principal stage of the plant and it is the bigger tank in the system, also CLEANING bed and COOLER are very expensive. The two dryers have a cost negligible with respect to the other components.

The technical assumption for calculation of the value of $Z_c$, *equation (25)*, for exergo-economic analysis are [22]:

- number of years of operation $n$ = 20
- utilization factor $f$ = 0.9
- discount rate $i$ = 5%
- operation and maintenance cost = 3% of TASC

another external cost in the plant is electric power consumed by compressors and pump, from the EUROSTAT database for 2017 the price of electricity for the industrial user in Italy is 0.16 €/kWh [17], the simulation performed considers the price of biogas of 0.14 €/Sm$^3$ [22].

The result of the analysis is reassumed in *Table 20:*

| | **C (€/s)** | **E (kW)** | **c (€/kJ)** |
|---|---|---|---|
| air0 | 0 | 0.07 | 0 |
| air1 | 3.43E-05 | 0.59 | 5.78E-05 |
| biogas0 | 4.67E-03 | 712 | 6.55E-06 |
| biogas1 | 5.45E-03 | 712 | 7.65E-06 |
| biogas2 | 6.23E-03 | 752 | 8.28E-06 |
| biogas3 | 7.35E-03 | 768 | 9.57E-06 |
| biogas4 | 8.10E-03 | 763 | 1.06E-05 |
| biomtn1 | 9.86E-03 | 708 | 1.39E-05 |
| biomtn2 | 9.87E-03 | 708 | 1.39E-05 |
| gas-out | 0 | 10.11 | 0 |
| gasrec1 | 7.72E-04 | 42.4 | 1.82E-05 |
| gasrec2 | 7.72E-04 | 41.0 | 1.88E-05 |
| gasrec3 | 7.78E-04 | 41.0 | 1.90E-05 |
| liquid1 | 8.81E-04 | 63.3 | 1.39E-05 |
| liquid2 | 3.33E-04 | 18.3 | 1.82E-05 |
| liquid3 | 3.33E-04 | 15.4 | 2.16E-05 |
| watrec | 7.52E-04 | 1.51 | 4.97E-04 |
| w1 | 0 | 0 | 0 |
| w2 | 0 | 0 | 0 |
| water0 | 0 | 0 | 0 |
| water1 | 7.2E-04 | 1.14 | 6.58E-04 |
| water2 | 1.59E-03 | 10.8 | 1.46E-04 |
| Wk1 | 8.14E-04 | 18.3 | 4.44E-05 |
| Qcool | 0 | 0 | 0 |
| Wk2 | 2.4E-05 | 0.66 | 4.44E-05 |
| Wpump | 5.99E-04 | 13.4 | 4.44E-05 |
| Qdry1 | 0 | 0 | 0 |
| Qdry2 | 0 | 0 | 0 |

*Table 20: Upgrade to biomethane: exergo-economic cost*

The only product of the studied plant is the biomethane exiting from DRY2, the stream named *biomtn2*, in this case, the calculated cost of biomethane is 0.00987 €/s that corresponds to 0.0139 €/MJ or 0.48 €/$Sm^3$.

### 6.1.1 Exergo-economic index

The following tables summarize the exergo-economic indexes for the principal components of the plant, the first is the relative cost difference, *equation* (*33*), that shows how much change the cost of products of a component respect to the fuels (*Table 21* and *Figure 19*).

| | $C_{fuel}$ (€/kJ) | $C_{product}$ (€/kJ) | r |
|---|---|---|---|
| CLEAN UP | 6.55E-06 | 7.65E-06 | 0.168 |
| K1 | 4.44E-05 | 7.16E-05 | 0.611 |
| COOL | 9.57E-06 | 1.06E-05 | 0.108 |
| PUMP1 | 4.44E-05 | 8.60E-05 | 0.935 |
| SCRUB | 1.25E-05 | 1.39E-05 | 0.114 |
| DRY2 | 1.39E-05 | 1.39E-05 | 0.001 |
| FLASH | 1.39E-05 | 1.82E-05 | 0.307 |
| DRY | 1.88E-05 | 1.90E-05 | 0.008 |
| K2 | 4.44E-05 | 6.51E-05 | 0.465 |
| STRIP | 2.30E-05 | 4.97E-04 | 20.630 |

*Table 21: Upgrade to biomethane: relative cost difference*

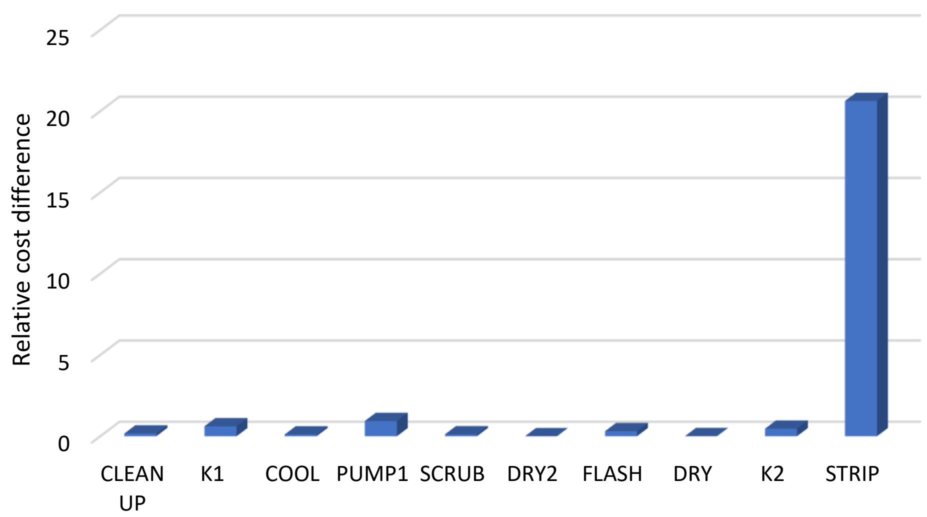


*Figure 19: Upgrade to biomethane: relative cost difference*

The relative cost difference is low for every stage of the plant except for STRIPPER, in this case, index *r* is very high because, considering the gas-out a lost for the system, all increase of cost is charged on the stream *wat-rec* that has a very poor exergy content, so the specific cost is very high.

The reason for the low relative cost difference in a component like SCRUBBER, COOLER or CLEANING is the absence of energy transformation and the low exergy losses. Compressors and pump are characterized by a bit higher index because they convert electrical power in physical exergy of pressure.

The *Table 22* and *Figure 20* show the exergo-economic factor $f_e$, *equation* (*35*).

| | Z (€/s) | I (kW) | $C_d$ | $f_{ex}$ |
|---|---|---|---|---|
| CLEAN UP | 7.85E-04 | 0 | 0 | 1.00 |
| K1 | 3.11E-04 | 2.49 | 1.11E-04 | 0.74 |
| COOL | 7.43E-04 | 4.60 | 4.41E-05 | 0.94 |
| PUMP1 | 2.36E-04 | 1.47 | 6.55E-05 | 0.78 |
| SCRUB | 1.06E-03 | 2.98 | 3.73E-05 | 0.97 |
| DRY2 | 7.01E-06 | 0 | 0.00E+00 | 1.00 |
| FLASH | 2.24E-04 | 2.59 | 3.61E-05 | 0.86 |
| DRY1 | 6.69E-06 | 0 | 0 | 1.00 |
| K2 | 4.91E-06 | 0.10 | 4.58E-06 | 0.52 |
| STRIP | 3.85E-04 | 4.40 | 1.01E-04 | 0.79 |

*Table 22: Upgrade to biomethane: exergo-economic factor*

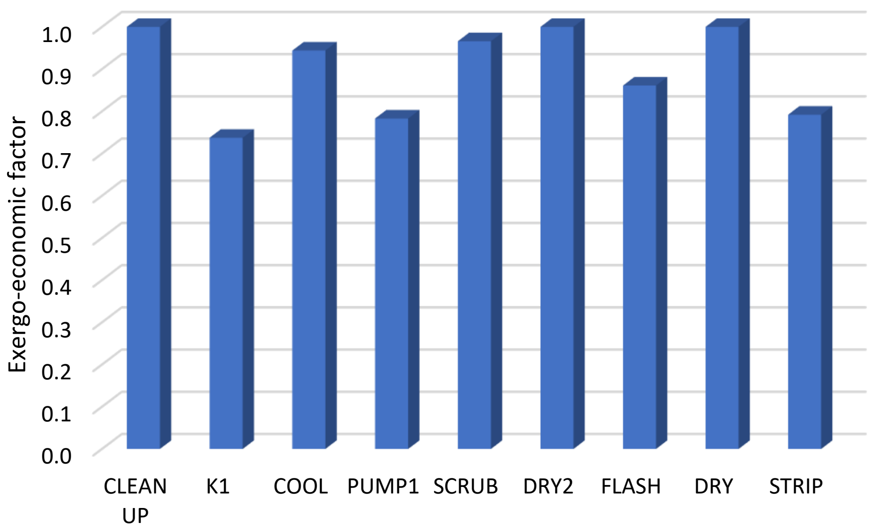


*Figure 20: Upgrade to biomethane: exergo-economic factor*

Due to low exergy losses for every component the exergo-economic factors are near to one for most of the equipment, in particular for CLEAN-UP, COOLER, SCRUBBER and FLASH vessel, so the main contribution to increase of cost between fuels and products is economic and not thermodynamic. For compressors and pump this index is lower but higher than 0.5. For components like mixers and valves, the index is equal to zero because in the analysis these elements are considered without economic cost.

## 6.2 SOFC-based CHP system

The cost of every component is determined from the cost functions contained in "Analysis Synthesis and Design of Chemical Process" [15] based on the type of equipment and its size, *equation* (28):

For blowers the size parameter A is the volumetric flow ($m^3/s$), the value is taken from the AspenPlus® simulation, for the heat exchangers the exchange area is the design variable and it is calculated using the equation:

$$A_{hx} = \frac{\dot{Q}}{U \cdot \Delta T_{ml}} \tag{44}$$

Where $\dot{Q}$ is thermal power exchanged, U is global thermal exchange coefficient that is assumed equal to 50 $W/m^2K$ for the gas-gas exchange and 100 $W/m^2K$ for gas-liquid exchange. $\Delta T_{ml}$ is the mean logarithmic temperature difference:

$$\Delta T_{ml} = \frac{\Delta T_1 - \Delta T_2}{ln \frac{\Delta T_1}{\Delta T_2}} \tag{45}$$

The size parameter of the reaction vessel is the volume ($m^3$) that ensures a sufficient residence time in the tank:

$$V = \frac{\dot{V}}{\tau} \tag{46}$$

Where $\dot{V}$ is the volumetric flow and $\tau$ is residence time, the hypothesis made for residence time are: for external reformer τ is 3 seconds instead for the after burner τ is 2 seconds (assumptions).

The solid oxide fuel cell is a component that is not reported in the cost function, for this reason, the cost of SOFC module is calculated from the NETL data [11] applying the scaling factor, *equation (29)*, to a reference base cost as reported in *Table 23*

| | **Reference power $W_{el\_0}$ (MW)** | **Reference cost $C_0$ (k$)** | **Scaling factor *n*** | **Request power $W_{el}$ (kW)** | **Effective cost $C_p^0$ ($)** |
|---|---|---|---|---|---|
| SOFC stack | 1 | 657 | 1 | 440 | 289,080 |

*Table 23: SOFC-based CHP system: SOFC stack cost*

| | k1 | k2 | k3 | A | $C_p^0$ ($) |
|---|---|---|---|---|---|
| FUEL-CMP | 3.5391 | -0.3533 | 0.4477 | 0.034 $m^3/s$ | 454 |
| FUEL-HX | 3.3444 | 0.2745 | -0.0472 | 2 $m^2$ | 2,647 |
| EXT-REF | 3.5565 | 0.3776 | 0.0905 | 1 $m^3$ | 3,601 |
| AIR-CMP | 3.5391 | -0.3533 | 0.4477 | 0.55 $m^3/s$ | 2,417 |
| AIR-HX | 3.3444 | 0.2745 | -0.0472 | 35 $m^2$ | 4,526 |
| REC-CMP | 3.5391 | -0.3533 | 0.4477 | 0.31 $m^3/s$ | 1,713 |
| BURNER | 3.5565 | 0.3776 | 0.0905 | 5 $m^3$ | 7,322 |
| CHP-HX | 3.3444 | 0.2745 | -0.0472 | 17 $m^2$ | 4,080 |

*Table 24: SOFC-based CHP system: cost functions*

*Table 24* report the coefficients for the cost function of the principal component of the plant and relative base cost. Cost of mixer and splitter is considered negligible. To calculate the bare erected cost is necessary to know the bare module factor that considers the material and pressure factor, the plant operates at atmospheric pressure so the pressure factor for every component is 1. Due to high temperatures reach in the plant vessels and heat exchangers are made of stainless steel and not by carbon steel for this reason the material factor for vessels is 3 and for heat exchangers is 2.7. The bare module factor for blowers is equal to 2.7.

The bare module factor for heat exchangers and vessel is calculated following *equation (42)*

| | B1 | B2 | $F_p$ | $F_m$ | $F_{bm}$ |
|---|---|---|---|---|---|
| FUEL-HX | 1.74 | 1.55 | 1 | 2.7 | 5.925 |
| EXT-REF | 1.49 | 1.52 | 1 | 3 | 6.05 |
| AIR-HX | 1.74 | 1.55 | 1 | 2.7 | 5.925 |
| BURNER | 1.49 | 1.52 | 1 | 3 | 6.05 |
| CHP-HX | 1.74 | 1.55 | 1 | 2.7 | 5.925 |

*Table 25: SOFC-based CHP system: bare module factor*

The cost of component calculated with cost function is valid for 2001 so it is necessary to scale this price in 2017 value, the ratio between 2001 and 2017 money can be expressed by the ratio between the CEPCI (Chemical Engineering Plant Index) indexes in the two years.

As already used in the section dedicated to upgrading the CEPCI for 2001 equal to 397 [15], the value of CEPCI for equipment in 2017 is 672 [26]. With these hypotheses the bare erected cost for each component are shown in *Table 26:*

| | FUEL-CMP | FUEL-HX | EXT-REF | SOFC | AIR-CMP | AIR-HX | REC-CMP | BURNER | CHP-HX |
|---|---|---|---|---|---|---|---|---|---|
| $C_{bm}$ ($) | 2,079 | 26,547 | 36,884 | 28,080 | 11,047 | 45,392 | 7,832 | 74,986 | 40,923 |

*Table 26: SOFC-based CHP system: bare erected cost*

Following the NETL/DoE cost estimation methodology [16] [27] TASC (total as-spent capital) can be calculated as the sum of various cost:

- EPCC = 10% of BEC
- process contingency = 30% of BEC (small pilot plant)
- project contingency = 15% of BEC + EPCC + Process contingency
- start-up = 2% of TPC
- inventory = 0.5% of TPC
- financing cost = 2.7% of TPC
- other owners cost = 15% of TPC
- financial cost =10% of TOC

Applying the euro/dollar change equal to 1.15 [28] the result is the TASC form each component in euro:

| | FUEL-CMP | CLEANING | FUEL-HX | EXT-REF | SOFC | AIR-CMP | AIR-HX | REC-CMP | BURNER | CHP-HX |
|---|---|---|---|---|---|---|---|---|---|---|
| **TASC (€)** | 5,179 | 171,540 | 66,120 | 91,864 | 719,991 | 27,515 | 113,055 | 19,506 | 186,762 | 101,923 |

*Table 27: SOFC-based CHP system: total as spent capital*

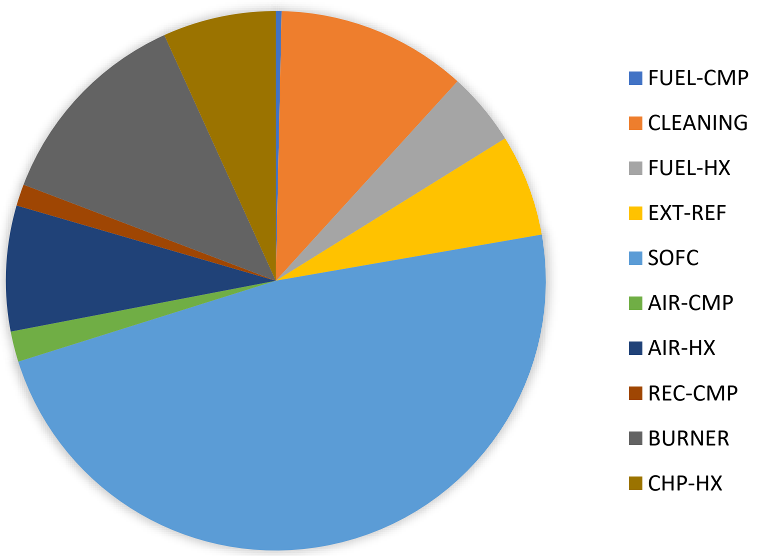


*Figure 21: SOFC-based CHP system: cost of components*

The total cost of the CHP plant is around 1,500,000 € and the main contribution to total investment cost is attributable to the solid oxide fuel cell that represents 50% of TASC. The other devices belong to the BOP and have a lower impact on the investment, the CLEANING bed and BURNER are the most expensive elements after the SOFC module, also the heat exchangers net has a not negligible economic cost. The blowers, instead, have a very marginal weight in the economic balance of the system.

The assumptions to calculate the value of the known term $Z_c$ for the exergo-economic equation are [10], [29]:

- lifetime of the plant 20 years
- inflation rate 5%
- utilization factor for the system 85% [11]
- replacement time of SOFC module 5 years
- operation and maintenance cost 3% of TASC every year

The cost of biogas is 0.14 €/$Sm^3$, electric power consumed by blowers is taken directly from the power produce by SOFC, so is not buy from the grid. Others input of the plant like air and water have a no cost, also the loss of plant *exhaust4* have a cost equal to zero.

The exergo-economic system described by the matrix formulation in *equation* (27) can be solved using the cost matrix in *Table 41* to determine the cost of each stream of the plant (€/s) and, based on the exergy flow, the specific cost (€/kJ) reported in *Table 28*.

The principal products of a CHP plant are electric power (stream *Wel*) and thermal power (difference of streams *wat-out* and *wat-in*). Using the exergo-economic methodology we can estimate the cost of products of the system:

- net electric power (408.4 kW) = 0.103 €/$kWh_e$
- recovered thermal power (186.7 kW) = 0.067 €/$kWh_t$

| | C (€/s) | Exergy (kW) | c (€/kJ) |
|---|---|---|---|
| biogas0 | 4.67E-03 | 707 | 6.60E-06 |
| biogas1 | 4.71E-03 | 707 | 6.65E-06 |
| biogas2 | 5.42E-03 | 707 | 7.66E-06 |
| biogas3 | 6.77E-03 | 729 | 9.29E-06 |
| biogas4 | 1.35E-02 | 975 | 1.38E-05 |
| biogas5 | 6.80E-03 | 491 | 1.38E-05 |
| biogas6 | 6.69E-03 | 483 | 1.38E-05 |
| bioref | 9.21E-03 | 560 | 1.64E-05 |
| an-fuel | 1.59E-02 | 1042 | 1.52E-05 |
| air0 | 0 | 0.39 | 0 |
| air1 | 3.13E-04 | 5.68 | 5.51E-05 |
| air2 | 9.97E-03 | 212 | 4.69E-05 |
| cat-ex | 7.21E-03 | 272 | 2.64E-05 |
| an-ex | 1.21E-02 | 456 | 2.64E-05 |
| burn-fuel | 5.45E-03 | 206 | 2.64E-05 |
| an-rec1 | 6.61E-03 | 250 | 2.64E-05 |
| an-rec2 | 6.72E-03 | 251 | 2.68E-05 |
| water | 0 | 0 | 0 |
| exhaust1 | 1.34E-02 | 413 | 3.25E-05 |
| exhaust2 | 1.24E-02 | 380 | 3.25E-05 |
| exhaust3 | 3.07E-03 | 94.5 | 3.25E-05 |
| exhaust4 | 0 | 23.5 | 0 |
| wat-in | 0 | 54.8 | 0 |
| wat-out | 3.36E-03 | 83.9 | 4.01E-05 |
| Wfuelcmp | 1.96E-05 | 0.74 | 2.65E-05 |
| Waircmp | 1.99E-04 | 7.33 | 2.72E-05 |
| Wreccmp | 2.83E-05 | 1.07 | 2.65E-05 |
| Wel | 1.16E-02 | 440 | 2.64E-05 |
| Qsofc | 2.03E-03 | 77 | 2.64E-05 |
| Qref | 2.03E-03 | 78 | 2.60E-05 |

*Table 28: SOFC-based CHP system: exergo-economic cost*

## 6.2.1 Exergo-economic indexes

Starting from the result of the previous analysis is possible to calculate some exergo-economic indices useful to identify the most impacting components of the plant on the final price of products from economic or exergy point of view.

The first index is the relative cost difference that can identify how much change the cost between fuel and product of a specific component, fuels and products are defined following the productive structure choose in *Table 6*.

| | $C_{fuel}$ (€/kJ) | $C_{product}$ (€/kJ) | r |
|---|---|---|---|
| FUEL-CMP | 2.65E-05 | 7.67E-05 | 1.894 |
| CLEANING | 6.65E-06 | 7.66E-06 | 0.151 |
| FUEL-HX | 3.25E-05 | 6.29E-05 | 0.935 |
| EXT-REF | 1.55E-05 | 1.64E-05 | 0.060 |
| REF-MIX | 1.55E-05 | 1.64E-05 | 0.060 |
| SOFC | 2.06E-05 | 2.64E-05 | 0.282 |
| AIR-CMP | 2.72E-05 | 5.91E-05 | 1.176 |
| AIR-HX | 3.25E-05 | 4.66E-05 | 0.435 |
| REC-CMP | 2.65E-05 | 1.20E-04 | 3.519 |
| BURNER | 2.64E-05 | 3.25E-05 | 0.231 |
| CHP-HX | 3.25E-05 | 1.16E-04 | 2.555 |

*Table 29: SOFC-based CHP system: relative cost difference*

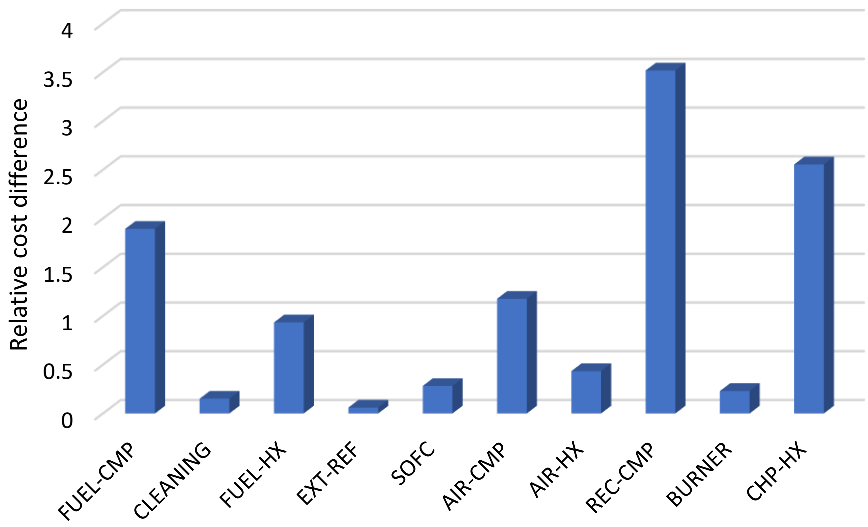


*Figure 22: SOFC-based CHP systme: relative cost difference*

The components with highest relative cost difference are the recirculation blower (REC-CMP) and the heat exchanger for hot water (CHP-HX), others equipment characterized by high relative cost difference are the other blowers.

Another important index is the exergo-economic index $f_{ex}$ that shows the contribution of economic cost and exergy lost in the process of cost formation.

| | Z (€/s) | I (kW) | $C_d$ (€/s) | $f_{ex}$ |
|---|---|---|---|---|
| FUEL-CMP | 2.14E-05 | 0.20 | 5.43E-06 | 0.80 |
| CLEANING | 7.10E-04 | 0.18 | 1.19E-06 | 1.00 |
| FUEL-HX | 2.74E-04 | 11.73 | 3.81E-04 | 0.42 |
| EXT-REF | 3.80E-04 | 9.10 | 1.41E-04 | 0.73 |
| REF-MIX | 3.80E-04 | 1.94 | 1.41E-04 | 0.73 |
| SOFC | 7.06E-03 | 8.90 | 1.83E-04 | 0.97 |
| AIR-CMP | 1.14E-04 | 2.04 | 5.53E-05 | 0.67 |
| AIR-HX | 4.68E-04 | 78.59 | 2.55E-03 | 0.15 |
| REC-CMP | 8.08E-05 | 0.16 | 4.18E-06 | 0.95 |
| BURNER | 7.73E-04 | 66.07 | 1.75E-03 | 0.31 |
| CHP-HX | 4.22E-04 | 41.95 | 1.36E-03 | 0.24 |

*Table 30: SOFC-based CHP system: exergo-economic factor*

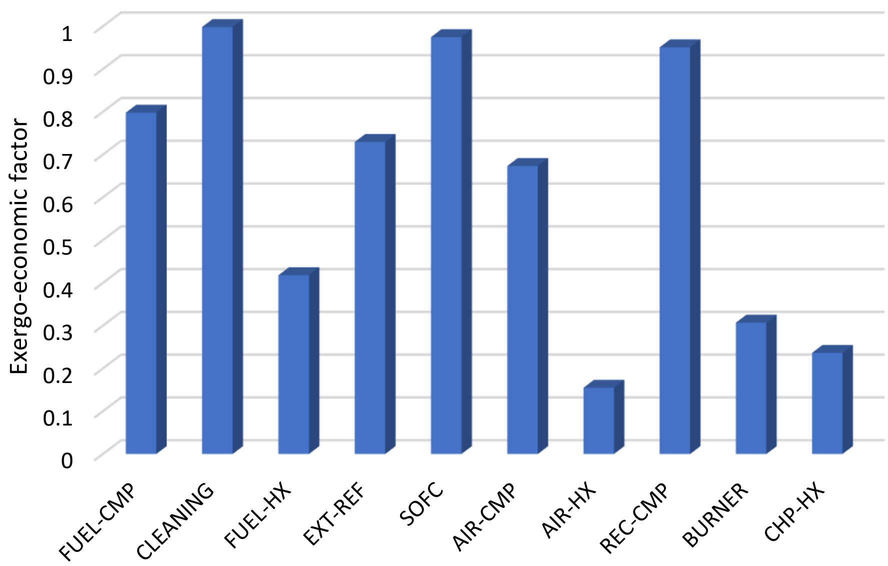


*Figure 23: SOFC-based CHP system: exergo-economic factor*

The SOFC module has an exergo-economic factor close to one (0.97) because it is characterized by a very high exergy efficiency, so a few exergies lost, and also a high cost, for this reason, the contribution to the increase of cost is mostly economic. The same concept is valid also for the cleaning. Components like burner and heat exchangers have a low factor because are simple and cheap and the exergy destroy in these devices is quite high. Blowers represent a good compromise between economic cost and thermodynamic efficiency and are characterized by exergo-economic factors near to 0.5.

# 7 Optimization of the biogas upgrading plant

The exergy and exergo-economic analysis developed on the biogas upgrading plant shows that the most critical device, concerning the process of cost formation, is the STRIPPER that is characterized by a low efficiency and high relative cost difference. Both negative properties are due to the release in the atmosphere of the stream *gas-out* which contains a not negligible exergy due to the high concentration of carbon dioxide, around 50% vol.

A possible change to improve the exergy and economic performance of the plant is to consider the gas stream exiting the STRIPPER (*gas-out*) as a secondary product of the system and not a loss. The stream *gas-out* is a mixture of 50% $CO_2$ and 50% air, due to the low concentration of carbon dioxide the mixture is not suitable for industrial application like enhanced oil recovery (EOR) or bauxite residue carbonation [2], a possible application of the $CO_2$ recovered in the upgrading plant is the use for the production of algal biomass [2], in this case a low concentration stream of carbon dioxide is used as nutrient to augment the grow of algae that can be used as feedstock for anaerobic digester. This improvement does not need a change of physical structure of the plant but only a change in the productive structure. *Table 31* summarizes the new productive matrix of the plant.

| COMPONENTS | FUEL | PRODUCT | LOST |
|---|---|---|---|
| CLEAN-UP | biogas0 | biogas1 | |
| MIX1 | biogas1 + gasrec2 | biogas2 | |
| K1 | Wk1 | biogas3 - biogas2 | |
| COOL | biogas3 | biogas4 | Qcool |
| PUMP1 | Wpump1 | water2 - water1 | |
| SCRUBBER | biogas4 + water2 | biomtn1 + liquid1 | |
| DRY2 | biomtn1 | biomtn2 | w2 + Qdry2 |
| FLASH | liquid1 | liquid2 + gasrec1 | |
| V1 | gasrec1 | gasrec2 | |
| DRY1 | gasrec2 | gasrec3 | w1 + Qdry |
| V2 | liquid2 | liquid3 | |
| DRY1 | gasrec2 | gasrec3 | Qdry1 + w1 |
| K2 | Wk2 | air1 - air | |
| STRIPPER | liquid2 + air1 | wat-rec + gas_out | |
| MIX2 | wat_rec + water0 | water1 | |
| global | biogas0 + Wk1 + Wk2 + Wpump1 | biomtn1 + gas_out | Qcool + w1 + Qdry1 + w2 + Qdry2 |

*Table 31: Productive matrix of improved biogas upgrade to biomethane*

Because of the new system configuration of fuels products and losses the exergy efficiency of the STRIPPER rises from 9.5% to 72.6%, also the global efficiency increases to 96.4%.

The new productive structure of biogas upgrading plant have the same incidence matrix of the previous configuration shown in *Table 36*, instead, the cost matrix changes to consider the other product, the new cost matrix is reported in the appendix *Table 42.* The change of productive structure has positive effects also on the economic performance of the plant because the Total As-Spent Capital does not increase but, adding a product to the system, the specific cost of biogas decreases. The results of the exergo-economic analysis developed on the new configuration are reported in *Table 32.*

| | C (€/s) | E (kW) | c (€/kJ) |
|---|---|---|---|
| air0 | 0 | 0.07 | 0 |
| air1 | 3.43E-05 | 0.59 | 5.78E-05 |
| biogas0 | 4.67E-03 | 712.51 | 6.55E-06 |
| biogas1 | 5.45E-03 | 712.51 | 7.65E-06 |
| biogas2 | 6.19E-03 | 752.77 | 8.22E-06 |
| biogas3 | 7.32E-03 | 768.47 | 9.52E-06 |
| biogas4 | 8.06E-03 | 763.87 | 1.05E-05 |
| biomtn1 | 9.22E-03 | 708.44 | 1.30E-05 |
| biomtn2 | 9.23E-03 | 708.45 | 1.30E-05 |
| gas-out | 6.39E-04 | 10.11 | 6.32E-05 |
| gasrec1 | 7.32E-04 | 42.39 | 1.73E-05 |
| gasrec2 | 7.32E-04 | 40.99 | 1.79E-05 |
| gasrec3 | 7.38E-04 | 41.02 | 1.80E-05 |
| liquid1 | 8.24E-04 | 63.30 | 1.30E-05 |
| liquid2 | 3.16E-04 | 18.30 | 1.73E-05 |
| liquid3 | 3.16E-04 | 15.41 | 2.05E-05 |
| watrec | 9.58E-05 | 1.51 | 6.32E-05 |
| w1 | 0 | 0 | 0 |
| w2 | 0 | 0 | 0 |
| water0 | 0 | 0 | 0 |
| water1 | 9.58E-05 | 1.14 | 8.37E-05 |
| water2 | 9.31E-04 | 10.85 | 8.58E-05 |
| Wk1 | 8.14E-04 | 18.32 | 4.44E-05 |
| Qcool | 0 | 0 | 0 |
| Wk2 | 2.94E-05 | 0.66 | 4.44E-05 |
| Wpump | 5.99E-04 | 13.47 | 4.44E-05 |
| Qdry1 | 0 | 0 | 0 |
| Qdry2 | 0 | 0 | 0 |

*Table 32: Improved upgrade to biomethane: exergo-economic cost*

Considering the stream of air rich of $CO_2$ exiting the STRIPPER, *gas-out*, as a product the cost of the principal product of the plant, *biomtn2,* decreases from 0.00987 €/s to 0.00923 €/s that correspond to a specific cost of 0.013 €/MJ of exergy or 0.45 €/Sm$^3$ instead of 0.48 €/Sm$^3$ calculated with the previous configuration. The stream *gas-out* which is now a product has a cost of 0.00064 €/s that correspond to 0.025 €/Sm$^3$ of a mixture of air and carbon dioxide at 50% vol.

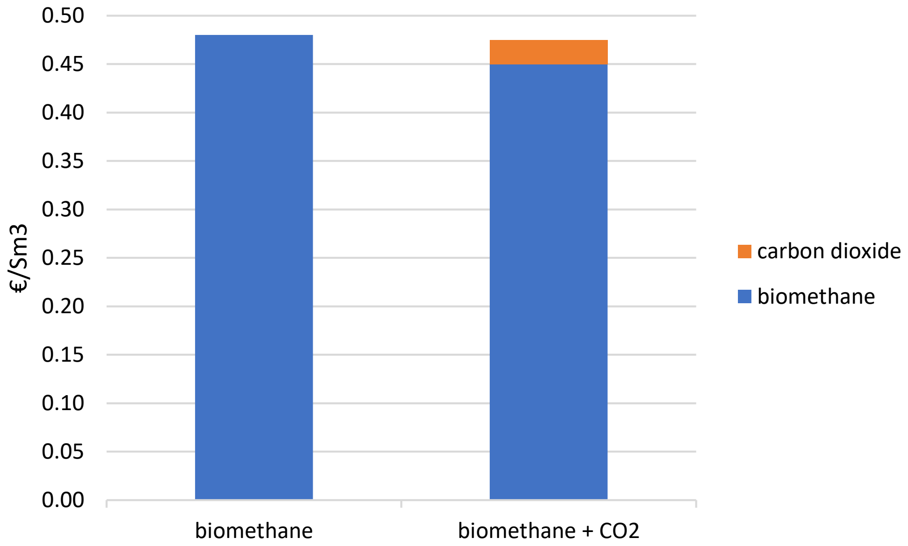


*Figure 24: Upgrade to biomethane: cost of products*

The improve of the economic performance of the biogas upgrading plant is underline by the relative cost difference index, the index for STRIPPER decreases a for 20 to less than 2. The relative cost differences indexes for every device in the new productive structure are shown in *Figure 25.*

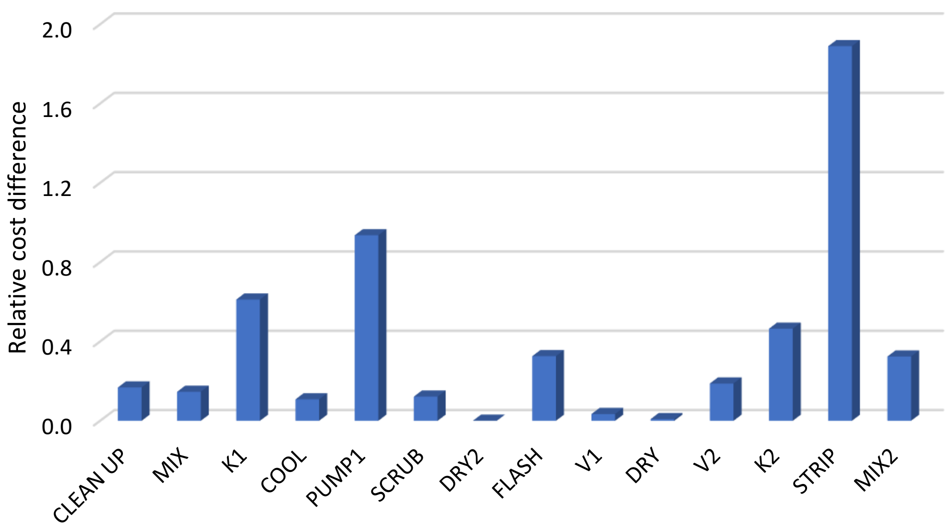


*Figure 25: Improved upgrade to biomethane: relative cost difference*

# 8 Economic results

## 8.1 Upgrade to biomethane

The result of the exergo-economic analysis is a specific cost of biomethane of 0.48 €/$Sm^3$ in the case of production of only biomethane or 0.45 €/$Sm^3$ in the case of polygeneration of methane and $CO_2$. The average price of the natural gas in the Italian market in 2017 is 0.19 €/$Sm^3$ [21] so the price of selling of the product of the upgrading plant is lower than the cost of production, without public incentive the upgrading of the biogas is a process economically disadvantageous.

The technical and economic assumptions needed for the economic analysis of the biogas upgrading plant are the one already used for the exergo-economic analysis (*chapter 6.1*):

- number of years of operation $n$ = 20
- utilization factor $f$ = 0.9
- discount rate $i$ = 5%
- operation and maintenance cost = 3% of TASC

cost of electricity is 0.16 €/MWh [17] and 0.14 €/$Sm^3$ [6] for the raw biogas entering the system, for the selling price of methane is not used a different value for each month but the yearly average price in 2017.

The first value of the economic analysis is the investment cost that corresponds to the TASC already compute in the exergo-economic analysis, this value, for the upgrading, is equal to 961,566 €, for hypothesis the investment is made in the year zero. In the following years, during the lifetime of the plant, the cost are the raw biogas, electric power and operation and maintenance of the components.

With a consumption of 120 $Sm^3$/h of biogas and a utilization factor of 0.9, the studied system needs 946,080 $Sm^3$ of biogas every year that means a cost of 132,451 €/y. The power request to drive compressors and pump is 32.4 kW with a yearly consumption of 255.5 MWh/y and an economic cost of 40,921 €/y. The last costs of the plant are the operation and maintenance with a global value of 28,847 €/y and the cost of labor of 50,000 €/y.

The earnings of the biogas upgrading plant are the selling of biomethane to the market, the public subsidies recognized to the production of renewable energy and eventually the selling of the mixture of air and carbon dioxide contained in the stream *gas-out*. The reference value for the selling price of biomethane is the yearly mean price of 2017 equal to 0.19 €/$Sm^3$ [21], the annual production is 581,050 $Sm^3$/y that produces an earning of 111,760 €/y. The $CO_2$ does not have a national market with a defined price and also the stream considered is not pure carbon dioxide so is not simple to define a price for this product, for these reasons the hypothesis is to consider the selling price equal to the production cost (0.025 €/$Sm^3$). The yearly volume of *gas-out* produced is 737,154 $Sm^3$/y which is around 50% of $CO_2$, in the case of selling this can increases the earnings of 18,140 €/y.

The calculation of the incentive starts from the value of the mean price of natural gas in 2012 that is 0.26 €/$Sm^3$ [22] and the monthly mean price which is considered equal to the yearly mean price of 2017. With this date and the application of *equation* (38), the base subsidies result in 0.33 €/$Sm^3$, considering the size of the studied plant smaller than 500 $Sm^3/h$ the base incentive is modulated with an increase of 10% so the modulated subsidies become 0.36 €/$Sm^3$. In the case, the biogas is produced from the waste listed in Table 1A [19] the modulated incentive increases of 50% so the final subsidy becomes 0.54 €/$Sm^3$. Considering the energy produced and the energy consumed converted in following the equivalations shown in *chapter 3.5.1* the equivalent renewable energy produced by the upgrading plant is around 393 toe/y that correspond to 505,375 $Sm^3$/y. Applying the final subsidies to the amount of equivalent biomethane produced the yearly public incentive is 275,773 €/y, the analysis is performed also without taking account the final increase of 50% of subsidies to consider the hypothesis of biogas not produced from waste or a possible decrease of public incentive in the future. If the biomethane produced in the biogas upgrading plant is used in transport application the public incentives are based on CIC, the production of 581,050 $Sm^3$/y of biomethane correspond to 19,174 GJ equivalent to 458 CIC, with a value of 500 € for each CIC [22] the incentive recognized to the plant is 229,000 €/y, the value is doubled if the biogas comes from substrates of Table 1A [19].

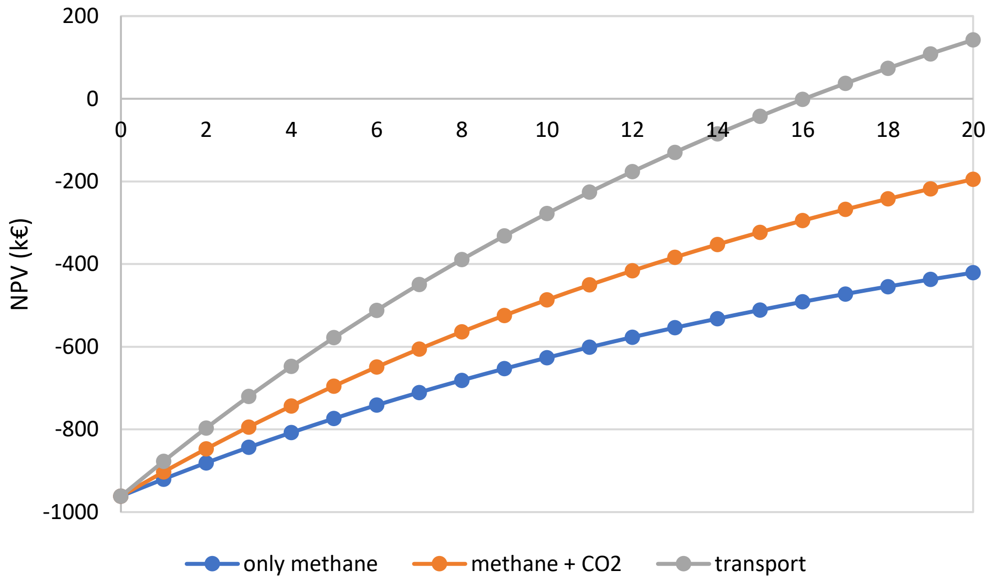


*Figure 26: Upgrade to biomethane: NPV lower incentives*

*Figure 26* shows the behavior of NPV, *equation* (*37*), of the biogas upgrading plant along the 20 years of the lifetime with the lowest subsidies in the two case of selling only the biomethane or selling both biomethane and mixture containing carbon dioxide. The results of the analysis are:

- only biomethane: NPV (20 years) = -420,838 €, there is not a PBT.
- biomethane and carbon dioxide: NPV (20 years) = 194,771 € and no PBT.
- transportation use: NPV (20 years) = 142,259 € and PBT = 17 years.

In the case of lowest incentive, the selling of the biomethane for grid injection and $CO_2$ is not sufficient to cover the cost of production so the investment is not economically convenient, this condition is confirmed by the negative NPV after 20 years. In the case of utilization of the biomethane produced for transport use, the process became convenient with a positive NPV but a long PBT.

*Figure 27* reports the result of the economic analysis for the upgrading in the case of the highest incentive considered, the increase of state contribution improves the economic performance of the plant increasing the NPV and decreasing the PBT due to the higher annual positive cash flow.

- Only biomethane: NPV (20 years) = 724,745 € and PBT = 9 years.
- Biomethane and carbon dioxide: NPV (20 years) = 950,811 € and PBT = 8 years.
- Transport use: NPV (20 years) = 2,996,521 € and PBT = 4 years.

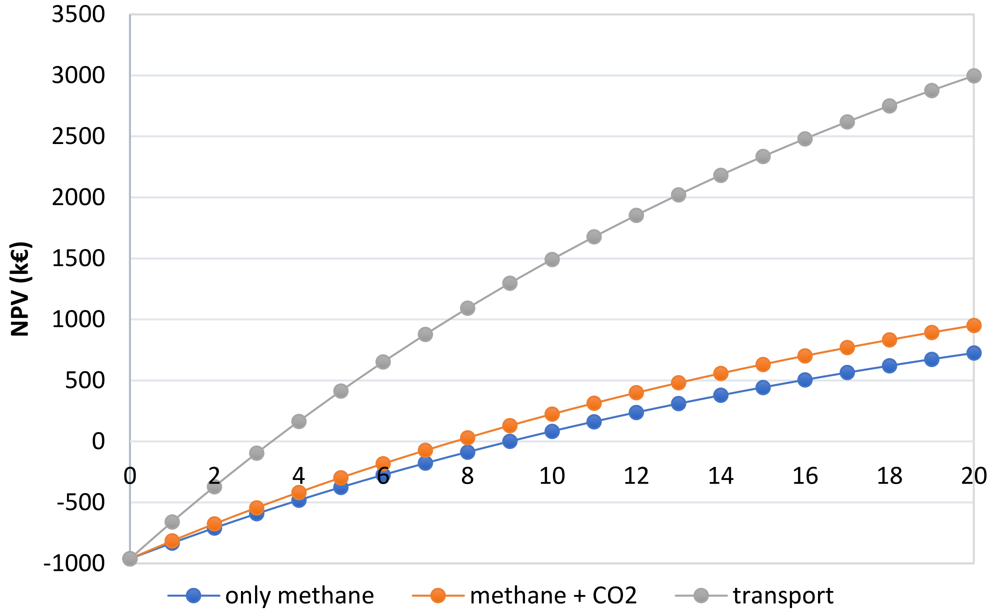


*Figure 27: Upgrade to biomethane: NPV higher incentives*

## 8.2 SOFC-based CHP system

The economic analysis starts from the results of the exergo-economic analysis of the SOFC plant which gives a production cost of 103 €/$MWh_e$ for electric power and 65 €/$MWh_t$ for thermal power. The market price of electricity sold to nation Italian grid is 65.77 €/MWh (November 2017 [21]) so, like the case of biogas, without a national regulation of public subsidies, the investment in SOFC power plant fed by biogas is economically disadvantageous.

The technical and economic assumptions needed for the economic analysis of the CHP SOFC plant are the one already used for the exergo-economic analysis (*chapter 6.2*):

- number of years of operation $n$ = 20
- utilization factor $f$ = 0.85
- discount rate $i$ = 5%
- operation and maintenance cost = 3% of TASC

The starting investment cost for the combined heat and power plant studied is already calculated in the exergo-economic analysis corresponding to the global TASC that has a value of 1'503'454 €, as already hypothesized the investment is made in the year zero. When the plant is in function the costs are represented by the biogas and operating and maintenance, the power consumed by blowers is not a cost for the plant because the neede electricity is not bought from the grid but is an internal consumption directly taken from the power produced by SOFC.

The specific biogas consumption is 120 $Sm^3/h$, considering the utilization factor and the cost of biogas the yearly consumption is 893,520 $Sm^3/y$ with a cost of 125,093 €/y. O&M defined by a

fraction of TASC has a cost of 45,104 €/y. The power plant has a periodical extra cost due to the replacement of the SOFC module every 5 years with a cost of 719,991 €.

The earings of the plant are the selling of power produced and the thermal power recovered from the exhaust. Considering the internal consumption and the inverter efficiency the net power produced is 408.8 kW with a yearly production of 3,044 MWh/y, considering a selling price of 65.77 €/MWh the annual earning from electricity is 200,199 €/MWh. Like the case of carbon dioxide for the upgrading plant, there is not a national market for the thermal power so is not possible to define a selling price for this product of the system, so the selling price is considered equal to the product cost itself. The thermal power recovered has a value of 186.7 kW so a yearly amount of 1,390 MWh/y that has an economic value of 90,361 €/y.

The public incentives are fundamentals for the economic balance of the SOFC power plant because without them the selling price of electricity is lower than the production cost. The economic analysis performed considered different subsidies level:

a) the biogas is produced from biomass of type (a), in this case, the base incentive for a plant with a nominal power between 300 kW and 600 kW is equal to 140 €/MWh [23]. The plant has also a prize incentive of 40 €/MWh due to the cogeneration [19], applying the *equation (40)* the final public subsidies is equal to 180 €/MWh.
b) The biogas is produced from biomass type (b), following the regulation the base incentive for the plant with a nominal power between 300 kW and 600 kW is 180 €/MWh. In this case, the thermal power recovered is used to feed a district heating network so prize incentive is 40 €/MWh. The final subsidie is equal to 220 €/MWh.
c) The conditions are the same of the point b) but the recovered heat is not used in a district heating network so the prize incentive is only 10 €/MWh [19], for this reason, the final public subsidies is 190 €/MWh.

The studied plant does not have a right to have the prize subsidies recognized for atmospheric emission because the decree prescribes a limiting concentration of $NO_x$ equal to 200 $mg/Nm^3$ of flue gas with a concentration of oxygen of 11%, instead the emissions of the plant are higher than this value and equal to 320 $mg/Nm^3$ (11% $O_2$).

*Figure* 28 shows the net present value, see *equation* (*37*), for every year of the lifetime of the SOFC power plant with the three different case of public subsidies:

a) Is the case of lower incentive and the result of the analysis is the NPV (20 years) = 2,450,655 € and PBT = 5 years,
b) NPV (20 years) = 2’829’996 € and PBT = 4 years
c) In the case of highest public subsidies NPV (20 years) = 3’968’017 € and PBT = 4 years

The plant based on SOFC has a great economical performance with every subsidies condition considered the NPV after 20 years always highest than 2 million euro and a short pay-back time.

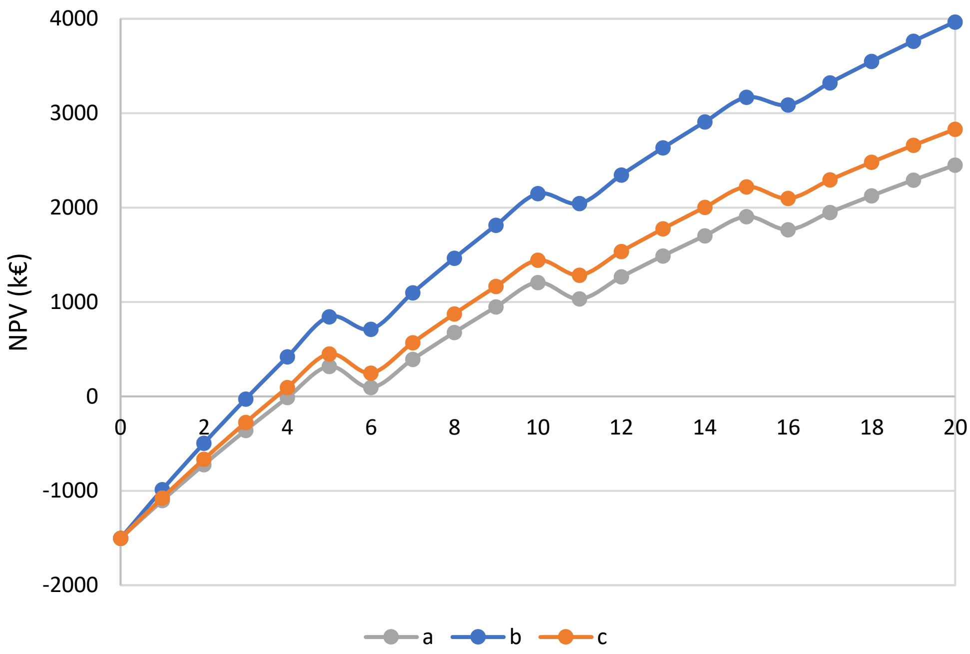


*Figure 28: SOFC-based CHP system: NPV; case a) public incentive=180 €/mWh; case b) public incentive=220 €/MWh; case c) public incentive=190 €/MWh*

# 9 Sensitivity analysis

## 9.1 Variation of the biogas cost

After the determination of the exergo-economic cost of production of the products of both the studied plants, a sensitivity analysis is necessary to evaluate the variation of the production cost at the variation of some input parameter of the two plants. The first input variable is the cost of feeding biogas, the base price is 0.14 €/Sm$^3$ and the sensitivity analysis considers the increase of 10% and a specular decrease, in these conditions for both plants is reported the variation of OPEX and of the cost of the principal product. OPEX is the operational cost of the plant that comprehends the maintenance cost, the cost of labor and the cost of fuels. The result of the analysis on the biogas cost is reported in *Figure 29* for upgrading and in *Figure 30* for SOFC.

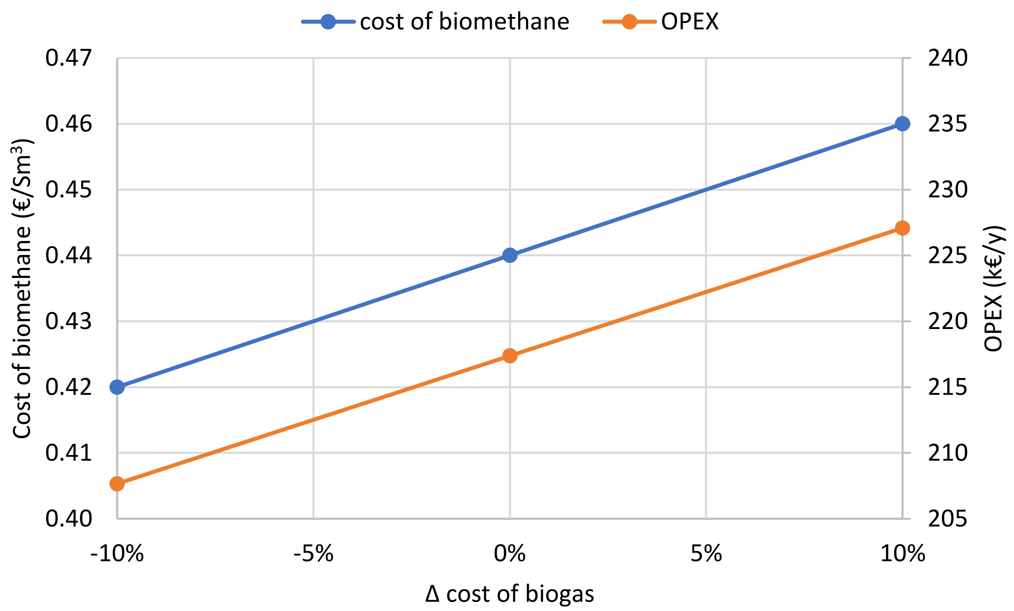


*Figure 29: Upgrade to biomethane: cost of biomethane vs. cost of biogas*

The increase or the decrease of the cost of biogas causes the consequential increase or decrease of the CAPEX, the variation is lower in the case of upgrading (±4%) than in the case of SOFC (±6%) because the biogas is not the only fuel of the upgrading plant, there is also the electric power to drive the compressors. For both plants the relative variation of the product cost is lower than the relative variation of the fuel price, ±5% for upgrading and ±3% for SOFC plant, because the production cost of the output of the system is the product of the capital cost (CAPEX) which are constant and the operative cost (OPEX) which varies with the cost of biogas feeding the plant.

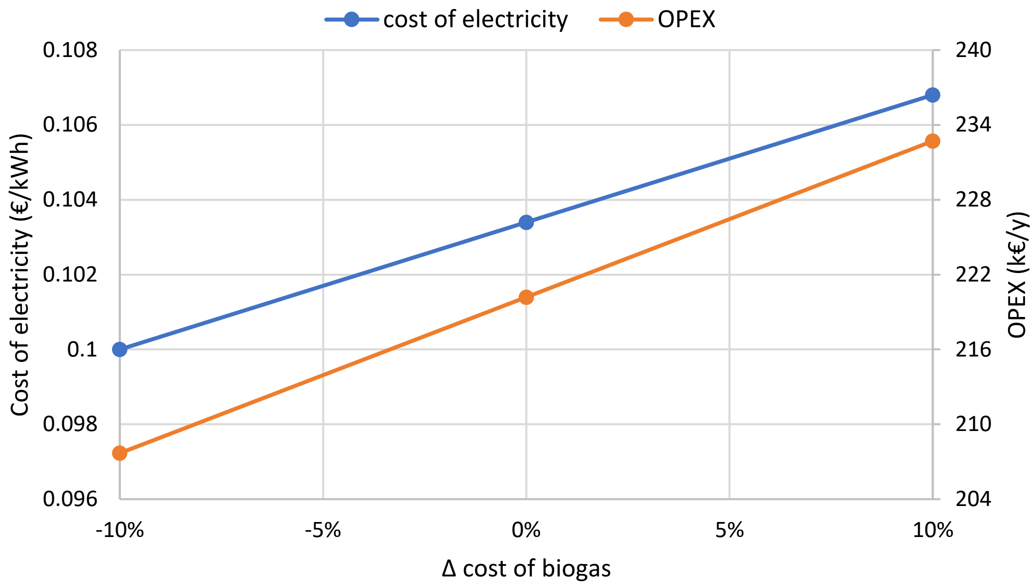


*Figure 30: SOFC-based CHP system: cost of electricity vs. cost of biogas*

## 9.2 Variation of the SOFC investment cost

The most important contribution to the investment cost in the cogenerative power plant based on solid oxide fuel cell is the SOFC module (around 50%), the development of the fuel cell technology can reduce the specific cost of this component in the next years. *Figure 31* shows the results of the analysis in the case of decreasing the price of the SOFC.

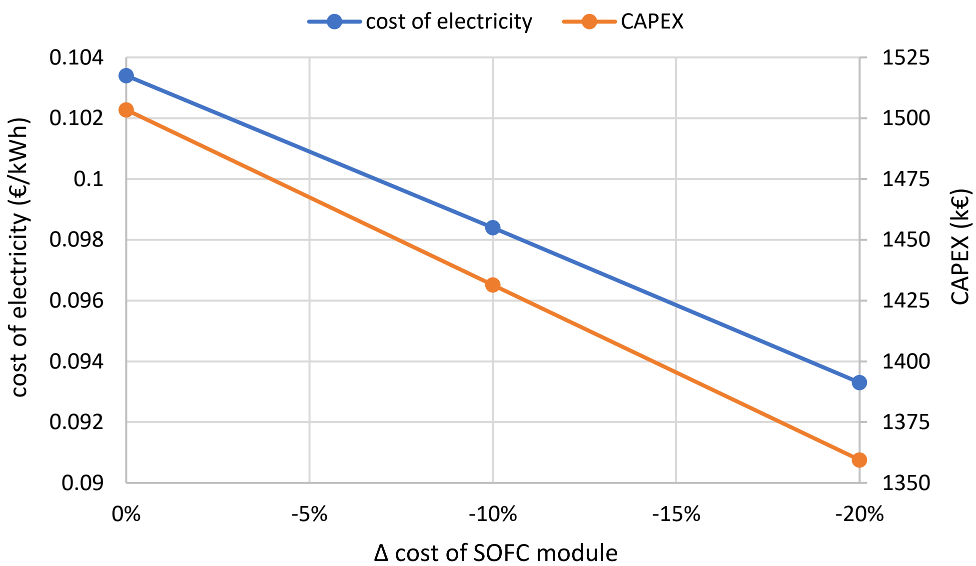


*Figure 31: SOFC-based CHP system: cost of electricity vs. investment cost*

In this case the CAPEX decrease with the decrease of SOFC module cost, a reduction of 10% of the price of this component produces a reduction of 5% of the CAPEX, instead, the OPEX has a constant contribution due to the biogas and the labor and a variable contribution due to maintenance cost (calculated as 3% of TASC). The cost of electricity produced is depending on the capital cost, but the decrease is lower due to the constant contribution to the OPEX, a reduction of 10% in the cost of SOFC stack produces a reduction of less than 5% of the production cost of electricity.

## 9.3 Variation of plant size

The economic performances of plants are strongly related to the size of the plants itself because it affects the specific cost of investment and it can influence the value of the public subsidies. The sensitivity analysis on the size of both systems is performed considering other three volumetric flow of biogas: 88 $Sm^3/h$, 178 $Sm^3/h$ and 298 $Sm^3/h$. The three sizes are chosen to investigate other levels of incentive for the electric power production [23]:

- 88 $Sm^3/h$ (-27% respect to the nominal power) correspond to a net power produced by SOFC of 300 kW, so, based on "D.M 6 Giugno 2016" the incentive to power plant fed by biogas with a power lower or equal to 300 kW is 233 €/MWh for type (b) substrates and 170 €/MWh for type (a) substrates.
- 178 $Sm^3/h$ (+48% respect to the nominal power) correspond to 605 kW of electric power produced by CHP-SOFC plant, for this size (600<P≤1000 kW) the ministerial decree gives a subsidy of 160 €/MWh if biogas comes from type (b) products or 120 €/MWh if biogas comes from type (a) products.
- 298 $Sm^3/h$ (+148% respect to the nominal power) correspond to 1015 kW of power produced, for a plant with power higher than 1 MW the public subsidy is 112 €/MWh for type (b) substrates and 97 €/MWh for type (a) substrates.

For the biogas upgrading to biomethane plant, the public subsidies do not change in this analysis because the productivity of biomethane is always lower than 500 $Sm^3/h$. The data reported in this analysis refers to the best economic condition, so to the case with highest possible public incentives.

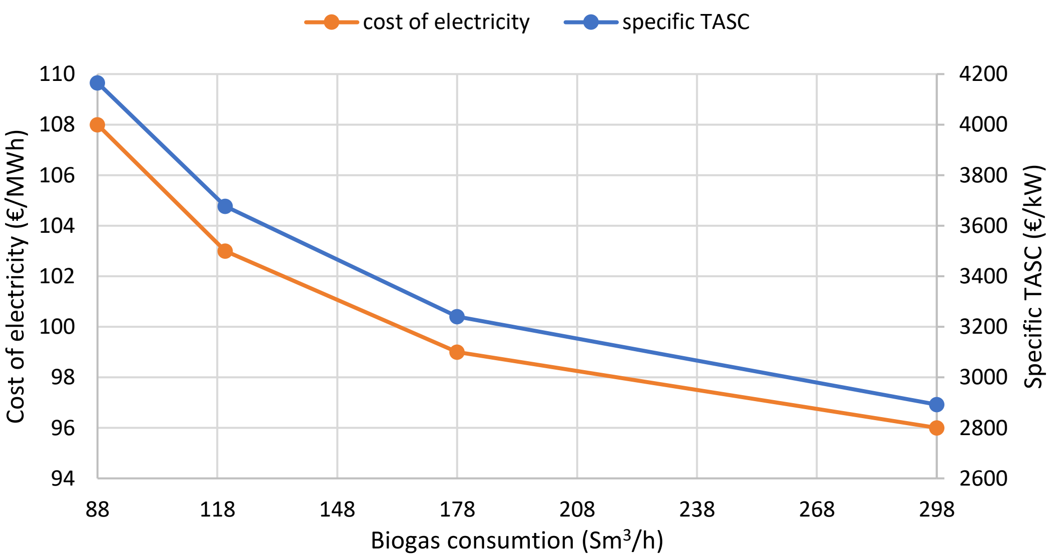


*Figure 32: SOFC-based CHP system: cost of electricity and specific investment cost vs. size*

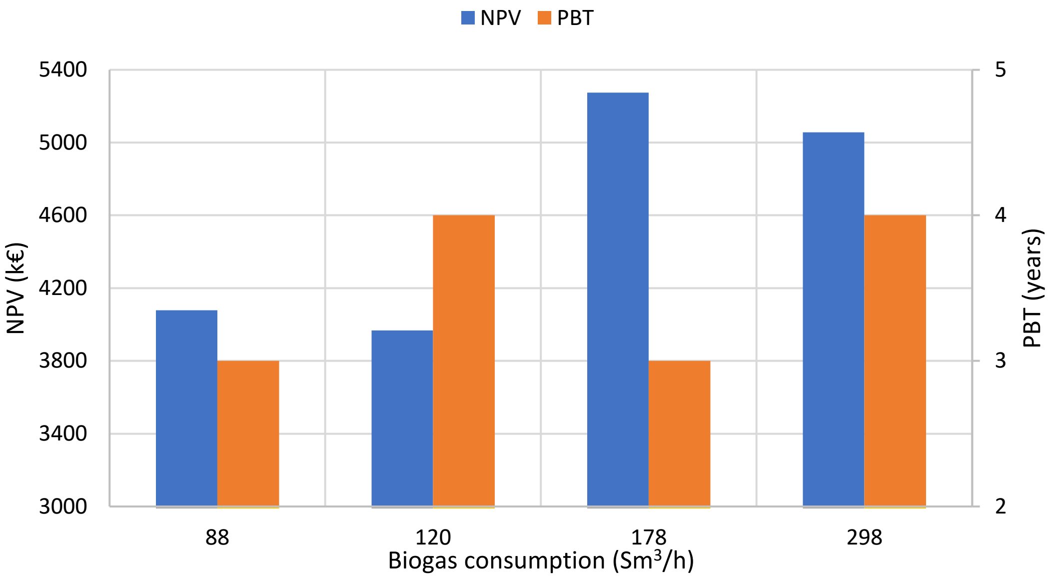


*Figure 33: SOFC-based CHP system: net present value and pay-back time vs. size*

*Figure 32* and *Figure 33* report the variation of the principal economic indicators of the SOFC plant with the change of biogas flowrate elaborated. The reduction of the specific investment cost is due to the decrease of specific price of the components of the plant, because the cost functions used to calculate the cost of components are not linear with the size of the component itself, see *equation* (28), except the SOFC module which has a constant specific cost, in consequence, the specific production cost of electricity decreases. The nominal power of 408.8 kW (120 $Sm^3/h$ of biogas) is not the best choice from the economic point of view because it is characterized by lower NPV and longer PBT then the other three options. In the case of 88 $Sm^3/h$ the specific investement is 13% higher (4166 €/kW) than nominal case, so the production cost is 5% higher (108 €/MWh) but the higher incentive (233 €/MWh) and lower capital cost (1,250,00 €) produce a higher NPV of 4,079,000 € and shorter PBT, in the case of 178 $Sm^3/h$ the lower incentive, 160 €/MWh is compensated by a 4% lower production cost (99 €/MWh) due to a reduction of 12% of specific investment cost (3240 €/kW), the resulting NPV after 20 years is 5,274,00 € and the PBT is equal to 3 years. The case of 298 $Sm^3/h$ have the same PBT of the nominal case (4 years) and higher NPV (5,057,000 €) but lower than the case of 178 $Sm^3/h$, the reduction of 21% in specific TASC, respect to nominal case, can not compensate the reduction of public subsidies (112 €/MWH) that has negative effects on economic performance of the plant.

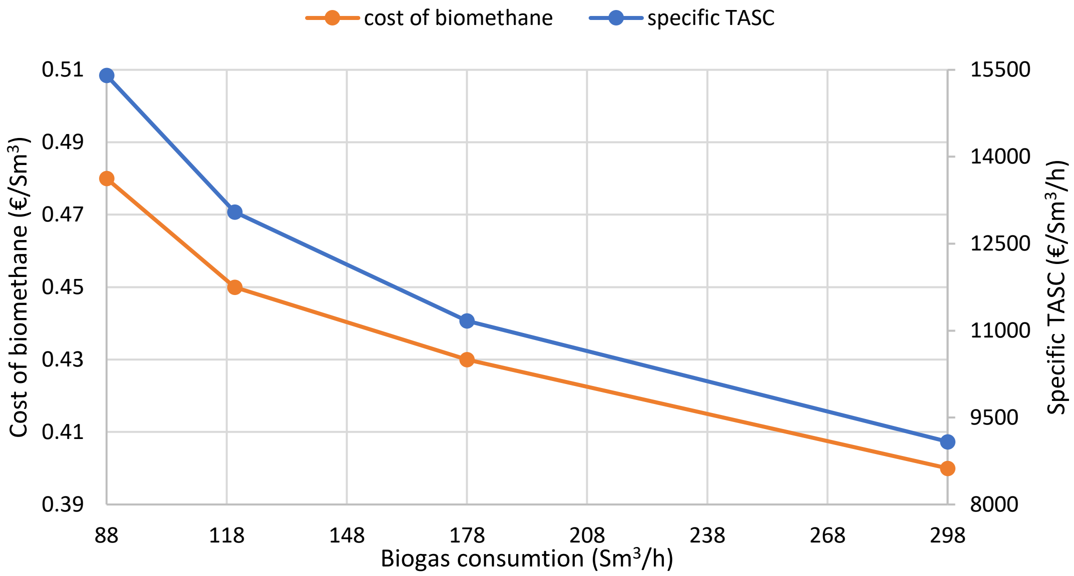


*Figure 34: Upgrade to biomethane: cost of biomethane and specific investment cost vs. size*

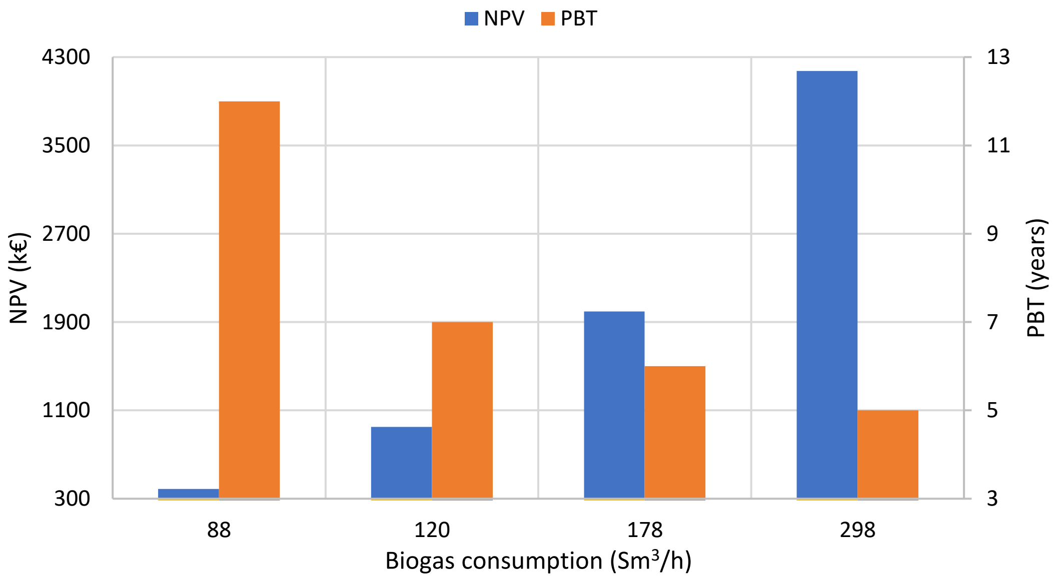


*Figure 35: Upgrade to biomethane: net present value and pay-back time vs. size*

*Figure 34* and *Figure 35* show the results of the sensitivity analysis on the upgrading plant in which the variable is the size of the system determined by the flow rate of raw biogas entering the plant. As expected the increase of dimension of the facility causes a decreasing of the specific investment cost, due to the non-linearity of cost function, and a consequential reduction of the specific cost of production of the biomethane, unless, the cost of biomethane is always higher than the market price of natural gas. The results of the lower specific TASC and the constant value of public subsidy in this range of productivity are the higher NPV and the shorter PBT of the plant with the increase of size.

# 10 Discussion and conclusion

The aim of the analysis is to develop a comparison between two types of plant fed by biogas to determine which technology is the best to valorize this important renewable and sustainable source of energy. The two productive plants have different principal products: biomethane for grid injection and electric power, both plants have also the possibility to produce and sell a secondary product: gas stream rich of carbon dioxide and thermal power recover. The exergo-economic analysis is useful to investigate the exergy and economic performance of the biogas upgrading plant and the CHP plant based on SOFC. The first comparison is based on the energy efficiency, see *equation* (2)*,* (*3*) and (*4*)*,* and exergy performances, see *equation*(*16*) and (*17*) reported in *Table 33:*

| | **Upgrade to biomethane** | **SOFC-based CHP system** |
|---|---|---|
| Energy efficiency | 95.0% | 88.1% |
| Exergy efficiency | 95.1% | 61.2% |
| Exergy destroy (kW) | 26.51 | 225.5 |
| Exergy destruction rate | 0.036 | 0.316 |

*Table 33: Exergy comparison*

The comparison shows how the upgrading plant has better exergy performance then SOFC power plant: the biogas upgrading is characterized by low exergy destruction and high exergy efficiency due to the absence of chemical reaction. The energy efficiency of the combined heat and power system is relatively high thanks to the thermal recovery of the exhaust gas, instead, the exergy efficiency is lower because the low-temperature heat has low exergy content.

The results of exergo-economic analysis of the two plants can be compared to understand the behavior of the technology in terms of cost of fuel and products and index that describes the global system:

| | **Upgrade to biomethane** | **SOFC-based CHP system** |
|---|---|---|
| Cost of fuel (€/kJ) | 8.20E-06 | 6.60E-06 |
| Cost of product (€/kJ) | 1.37E-05 | 2.86E-05 |
| Exergo-economic factor | 0.945 | 0.874 |
| Relative cost difference | 0.675 | 3.335 |

*Table 34: Exergo-economic comparison*

The fuel cost of biogas upgrading is bigger because the electric power needed to auxiliaries increase this cost, instead, the only fuel of the SOFC plant is the biogas with a lower specific cost. the cost of products of the upgrading is lower than the cost of products of the SOFC thanks to lower investment capital and lower exergy lost, for this reason, the relative cost difference of SOFC is larger than the one of upgrading. The exergo-economic factors of both systems are bigger than 0.5 so for both plants the economic cost of investment is more relevant than the economic cost due to exergy lost.

The last comparison regards the economic aspects of the plants and the elaboration of some indexes that can help to evaluate the profitability of investment and the economic maturity of the technology in respect to the energy market of Italy. The first index measures the relative distance between the cost of production of the principal product of the system and the market selling price without public incentive, *equation* (*47*):

$$DBE = \frac{cost\ of\ production - price\ of\ selling}{price\ of\ selling} \cdot 100 \tag{47}$$

The second index is the ratio between the net present value of the investment in the best economic situation determined from the public subsidies after the lifetime of the plant (20 years) and the initial investment cost to build the plant, *equation (*48*)*, higher is this index more advantageous is the investment from the economic point of view.

$$IR = \frac{NPV(20\ years)}{TASC} \tag{48}$$

| | Upgrade to biomethane | SOFC-based CHP system |
|---|---|---|
| Total as-spent cost (€) | 961,618 | 1,503,454 |
| Market price of product | 0.19 | 0.065 |
| Product cost | 0.44 | 0.103 |
| Distance from break-even | 131.6% | 56.6% |
| NPV (20 years) (€) | 950,818 | 3,968,017 |
| Investment return | 0.99 | 2.64 |
| PBT (years) | 8 | 4 |

*Table 35: Economic comparison*

*Table 35* shows the results of the economic comparison of two different plants, it is clear that the CHP plant base on solid oxide fuel cell technology has better economic performance than the biogas upgrading plant. Although the power plant has a larger starting investment cost respect to the upgrading and it is necessary to replace the SOFC module every five years the net present value of the SOFC after the lifetime of the system is more than three times bigger than the net present value of the upgrading after the same time. The investment rate index and the pay-back time underline the economic advantages of the SOFC plant with respect to the upgrading: IR of the solid oxide fuel cell is higher than two, so the initial investment is more than doubled in twenty years, instead, the IR of upgrading is lower than one. The shorter PBT means that the economic balance of the plant becomes positive in less time and this is important for investors. The distance from the economical break-even shows that, unless SOFC is a developing technology, the power plant base on fuel cell is already more competitive on the market then the PWS upgrading that is a commercial technology. The improving of the SOFC, especially the reduction of the production cost and the increment of the lifetime, can reduce the difference between the cost of production of power with this plant and the market price of electricity, so the SOFC CHP system can become economically competitive also without public incentives. The pressurized water scrubbing is a developed technique which uses standard components like scrubber and compressors so a significant reduction of the cost of investment is not predictable, for this reason, the technology is not a valid solution in a market without subsidies.

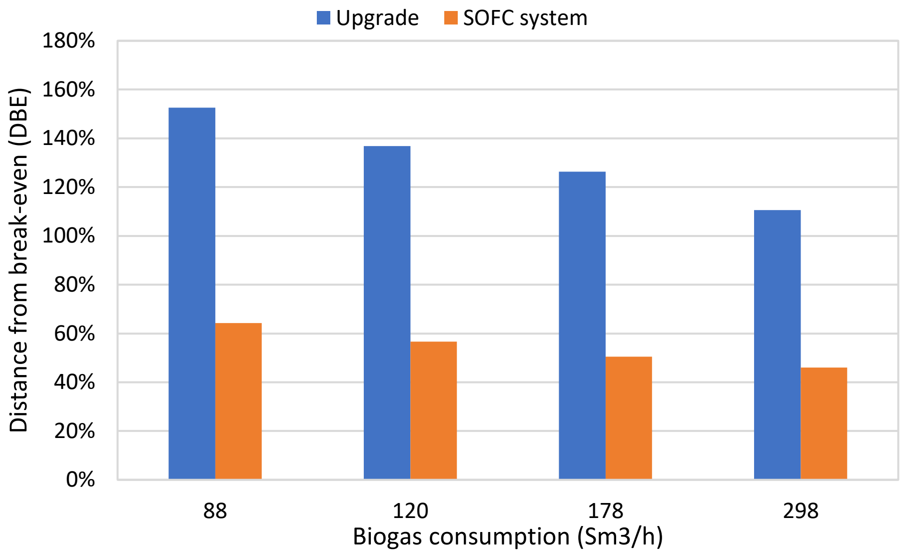


*Figure 36: Comparative distance from break-even (DBE)*

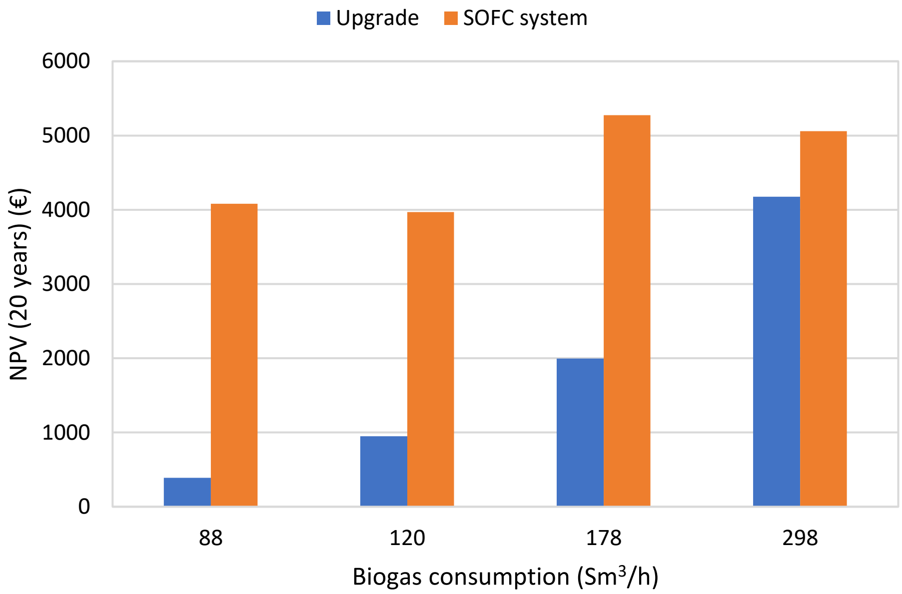


*Figure 37: Comparative net present value NPV*

*Figure 36* shows the variation of the DBE for the principal product of both plants, although the DBE of the biomethane produced from the upgrading decreases with higher gradient than the DBE of electric power produced by SOFC plant, the SOFC system has always a lower value of distance from break-even so, in a market without incentives, it is economically more competitive than upgrading also with the increasing of size. The economic performance of both plants are strongly influenced by the subsidies legislation, due to the decrease of incentive with the size of the plant in the case of power generation and the constant subsidy recognized to the biomethane in this range of production the net present value of the upgrading plant has a steep increase with the size of system, instead, the NPV of the SOFC is quite constant as shown in *Figure 37.* The investment return considers not only the NPV but also the initial investment of the plant, the values of IR for both plant varying the size are reported in *Figure 38*, for the upgrading the index has an increasing behavior because, with the increase of the size, the Net Present Value for this plant icreases more than the initial investment cost, instead for SOFC the IR is decreasing because the increase of the plant size causes an increase of investment cost but not a corresponding increase of the NPV as shown in *Figure 37* so, for a flow of biomethane elaborated higher than 298 $Sm^3/h$ the biogas upgrading becomes a more profitable solution to use the biogas.

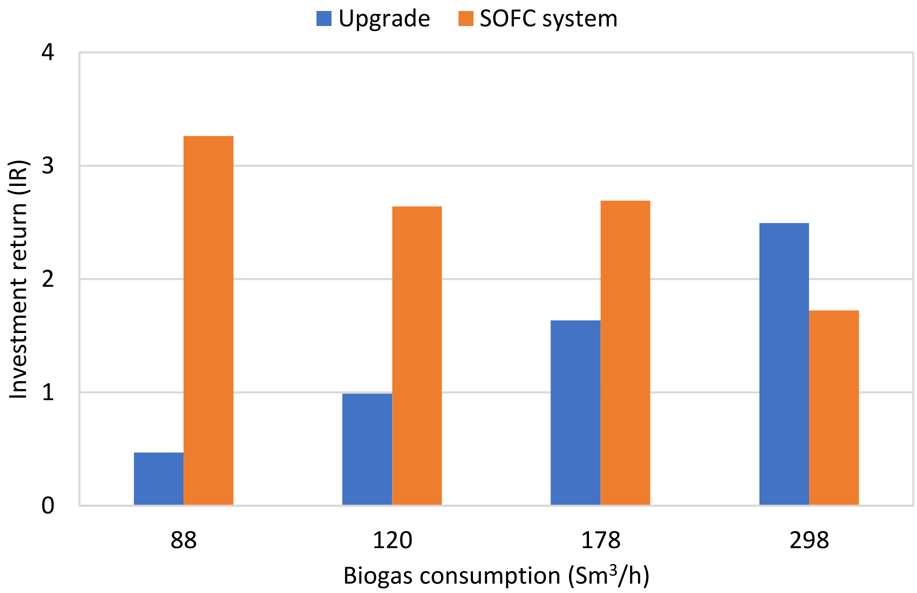


*Figure 38: Comparative investment return (IR)*

The results of the comparative analysis show that the pressurized water scrubbing upgrading has better exergy and exergo-economic performances with less exergy destroy and lower relative cost difference, due to absence of energy conversion in the plant, but the solid oxide fuel cell based power plant has better economic performance, NPV, PBT and investment return, and the future development in this sector can also decrease the investment cost and the productive cost of electric power as shown in the sensitivity analysis in *chapter 9.2*. The economic performance depends on the selling price of the products that is strongly dependent from the public subsidies. The Italian legislation recognize a high public subsidy to the power generation from biogas especially for small plant. In the technical and economic condition described in the paper the SOFC-based combined heat and power system results the better option to valorize biogas produced from municipal waste, agricultural waste or wastewater in the case of small facilities. The biogas upgrade to biomethane becomes a more interesting solution with the increase of the plant size because the decrese of specific investment cost and productive cost of biomethane and the constant public incentive until 500 $Sm^3/h$ of biomethane produced the economic performances of this type of plant improves.

# 11 Appendix

| | air0 | air1 | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biomtn1 | biomtn2 | gas-out | gasrec1 | gasrec2 | gasrec3 | liquid1 | liquid2 | liquid3 | watrec | w1 | w2 | water0 | water1 | water2 | Wk1 | Qcool | Wk2 | Wpump | Qdry1 | Qdry2 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| CLEANING | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MIX1 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| K1 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| COOLER | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 |
| PUMP1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 1 | 0 | 0 |
| SCRUBBER | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| DRY2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 |
| FLASH | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| V1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| DRY1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 |
| V2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| K2 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| STRIPPER | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MIX2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

*Table 36: incidence matrix of upgrading plant*

| | air0 | air1 | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biomtn1 | biomtn2 | gas-out | gasrec1 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Molar Flow (mol/s)** | 5.09E-01 | 5.09E-01 | 1.41E+00 | 1.41E+00 | 1.95E+00 | 1.95E+00 | 1.95E+00 | 8.72E-01 | 8.68E-01 | 1.09E+00 | 5.39E-01 |
| **Mass Flow (kg/s)** | 1.47E-02 | 1.47E-02 | 3.85E-02 | 3.85E-02 | 6.09E-02 | 6.09E-02 | 6.08E-02 | 1.46E-02 | 1.45E-02 | 3.91E-02 | 2.25E-02 |
| **Temperature (°C)** | 25.0 | 69.5 | 15.0 | 15.0 | 17.4 | 242.2 | 60.0 | 25.1 | 25.1 | 25.5 | 25.7 |
| **Pressure (bar)** | 1.0 | 1.5 | 1.0 | 1.0 | 1.0 | 10.5 | 10.5 | 10.0 | 10.0 | 1.0 | 3.0 |
| **Enthalpy (kJ/kg)** | -2.16E-01 | 4.48E+01 | -7.44E+03 | -7.44E+03 | -7.95E+03 | -7.65E+03 | -7.90E+03 | -4.98E+03 | -4.94E+03 | -5.68E+03 | -8.85E+03 |
| **Entropy (J/kg*K)** | 1.51E+02 | 1.75E+02 | -1.58E+03 | -1.58E+03 | -9.70E+02 | -8.51E+02 | -1.45E+03 | -5.77E+03 | -5.79E+03 | 2.44E+02 | -2.44E+02 |
| **Density (kg/cum)** | 1.16 | 1.52 | 1.2 | 1.15 | 1.30 | 7.65 | 12.08 | 6.86 | 6.86 | 1.45 | 5.11 |
| **Mole Fraction** | | | | | | | | | | | |
| **CO2** | 0.00E+00 | 0.00E+00 | 4.00E-01 | 4.00E-01 | 5.44E-01 | 5.44E-01 | 5.44E-01 | 2.52E-02 | 2.53E-02 | 4.99E-01 | 9.17E-01 |
| **CH4** | 0.00E+00 | 0.00E+00 | 6.00E-01 | 6.00E-01 | 4.56E-01 | 4.56E-01 | 4.56E-01 | 9.71E-01 | 9.75E-01 | 1.75E-03 | 7.15E-02 |
| **O2** | 2.10E-01 | 2.10E-01 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 9.72E-02 | 0.00E+00 |
| **N2** | 7.90E-01 | 7.90E-01 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 3.69E-01 | 0.00E+00 |
| **H2O** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 3.76E-03 | 0.00E+00 | 3.31E-02 | 1.14E-02 |

| | gasrec2 | gasrec3 | liquid1 | liquid2 | liquid3 | watrec | w1 | w2 | water0 | water1 | water2 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| **Molar Flow (mol/h)** | 5.39E-01 | 5.32E-01 | 3.75E+02 | 3.74E+02 | 3.74E+02 | 3.74E+02 | 6.15E-03 | 3.28E-03 | 3.97E-02 | 3.74E+02 | 3.74E+02 |
| **Mass Flow (kg/s)** | 2.25E-02 | 2.24E-02 | 6.78E+00 | 6.76E+00 | 6.76E+00 | 6.73E+00 | 1.11E-04 | 5.91E-05 | 6.28E-04 | 6.73E+00 | 6.73E+00 |
| **Temperature (°C)** | 23.6 | 23.6 | 25.9 | 25.7 | 25.6 | 25.4 | 23.6 | 25.1 | 25.0 | 25.0 | 25.1 |
| **Pressure (bar)** | 1.0 | 1.0 | 10.0 | 3.0 | 1.5 | 1.0 | 1.0 | 10.0 | 1.0 | 1.0 | 10.5 |
| **Enthalpy (kJ/kg)** | -8.85E+03 | -8.82E+03 | -1.58E+04 | -1.58E+04 | -1.58E+04 | -1.59E+04 | -1.59E+04 | -1.59E+04 | -1.59E+04 | -1.59E+04 | -1.59E+04 |
| **Entropy (J/kg*K)** | -2.71E+01 | -2.74E+01 | -9.02E+03 | -9.04E+03 | -9.04E+03 | -9.08E+03 | -9.10E+03 | -9.08E+03 | -9.08E+03 | -9.08E+03 | -9.08E+03 |
| **Density (kg/$m^3$)** | 1.70 | 1.71 | 754.48 | 753.65 | 510.05 | 752.83 | 753.76 | 753.06 | 753.0 | 753.02 | 753.08 |
| **Mole Fraction** | | | | | | | | | | | |
| **CO2** | 9.17E-01 | 9.28E-01 | 2.77E-03 | 1.45E-03 | 1.45E-03 | 4.41E-06 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 |
| **CH4** | 7.15E-02 | 7.23E-02 | 1.08E-04 | 5.07E-06 | 5.07E-06 | 1.17E-10 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 |
| **O2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 4.15E-06 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 |
| **N2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 6.47E-06 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 |
| **H2O** | 1.14E-02 | 0.00E+00 | 9.97E-01 | 9.99E-01 | 9.99E-01 | 1.00E+00 | 1.00E+00 | 1.00E+00 | 1.00E+00 | 1.00E+00 | 1.00E+00 |

*Table 37: Upgrading termodynamic results*

| | air0 | air1 | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biomtn1 | biomtn2 | gas-out | gasrec1 | gasrec2 | gasrec3 | liquid1 | liquid2 | liquid3 | watrec | w1 | w2 | water0 | water1 | water2 | Wk1 | Qcool | Wk2 | Wpump | Qdry1 | Qdry2 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| CLEANING | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MIX1 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| K1 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| COOLER | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 |
| PUMP1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 1 | 0 | 0 |
| SCRUBBER | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| DRY2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 |
| FLASH | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| V1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| DRY1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 |
| V2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| K2 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| STRIPPER | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MIX2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| a. | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| b. | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| c. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| d. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| e. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| f. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 |
| g. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| h | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| i. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| j. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 |
| k. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 |
| l. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 |
| m. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.09 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| n. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.43 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

*Table 38: Incidence matrix of upgrading plant*

| | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biogas5 | biogas6 | bioref | an-fuel | air0 | air1 | air2 | cat-ex | an-ex | burn-fuel | an-rec1 | an-rec2 | water | exhaust1 | exhaust2 | exhaust3 | exhaust4 | wat-in | wat-out | Wfuelcmp | Waircmp | Wreccmp | Wel | Qsofc | Qref |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| FUEL-CMP | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| CLEANING | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| FUEL-HX | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AN-MIX | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| REF-SPLIT | 0 | 0 | 0 | 0 | 1 | -1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| EXT-REF | 0 | 0 | 0 | 0 | 0 | 1 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 |
| REF-MIX | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOFC | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 1 | -1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | -1 | 0 |
| AIR-CMP | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 |
| AIR-HX | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AN-SPLIT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| REC-CMP | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| BURNER | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHP-HX | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 |

*Table 39: Incidence matrix of SOFC plant*

| | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biogas5 | biogas6 | bioref | an-fuel | air0 |
|---|---|---|---|---|---|---|---|---|---|---|
| **Molar flow (mol/s)** | 1.40E+00 | 1.40E+00 | 1.40E+00 | 1.40E+00 | 5.12E+00 | 2.58E+00 | 2.54E+00 | 3.42E+00 | 5.96E+00 | 2.28E+01 |
| **Mass flow (kg/s)** | 3.80E-02 | 3.80E-02 | 3.80E-02 | 3.80E-02 | 1.39E-01 | 7.03E-02 | 6.91E-02 | 7.03E-02 | 1.39E-01 | 6.61E-01 |
| **Temperature (°C)** | 14.9 | 28.2 | 28.1 | 645.9 | 750.0 | 750.0 | 750.0 | 750.0 | 750.0 | 15.0 |
| **Pressure (bar)** | 1.0 | 1.2 | 1.1 | 1.1 | 1.1 | 1.1 | 1.1 | 1.1 | 1.1 | 1.0 |
| **Enthalpy (kJ/kg)** | -7.44E+03 | -7.42E+03 | -7.42E+03 | -6.28E+03 | -8.04E+03 | -8.04E+03 | -8.04E+03 | -6.55E+03 | -7.29E+03 | -1.03E+01 |
| **Entropy (J/kg*K)** | -1.57E+03 | -1.56E+03 | -1.54E+03 | 4.24E+02 | 1.47E+03 | 1.47E+03 | 1.47E+03 | 3.25E+03 | 2.42E+03 | 1.32E+02 |
| **Density (kg/cum)** | 1.14 | 1.28 | 1.21 | 0.39 | 0.35 | 0.35 | 0.35 | 0.26 | 0.30 | 1.21 |
| **Mole fraction** | | | | | | | | | | |
| **H2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 6.92E-02 | 6.92E-02 | 6.92E-02 | 3.33E-01 | 2.20E-01 | 0.00E+00 |
| **CO2** | 4.00E-01 | 4.00E-01 | 4.00E-01 | 4.00E-01 | 3.90E-01 | 3.90E-01 | 3.90E-01 | 2.07E-01 | 2.85E-01 | 0.00E+00 |
| **O2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 2.10E-01 |
| **H2O** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 3.27E-01 | 3.27E-01 | 3.27E-01 | 2.12E-01 | 2.61E-01 | 0.00E+00 |
| **CO** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 4.98E-02 | 4.98E-02 | 4.98E-02 | 2.48E-01 | 1.63E-01 | 0.00E+00 |
| **CH4** | 6.00E-01 | 6.00E-01 | 6.00E-01 | 6.00E-01 | 1.64E-01 | 1.64E-01 | 1.64E-01 | 1.03E-03 | 7.03E-02 | 0.00E+00 |
| **N2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 5.65E-09 | 3.41E-09 | 3.41E-09 | 3.41E-09 | 2.58E-09 | 2.93E-09 | 7.80E-01 |
| **NO** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 8.23E-18 | 8.23E-18 | 8.23E-18 | 5.87E-19 | 3.84E-18 | 0.00E+00 |
| **NO2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 1.10E-27 | 1.10E-27 | 1.10E-27 | 6.21E-30 | 4.73E-28 | 0.00E+00 |
| | **air1** | **air2** | **cat-ex** | **an-ex** | **burn-fuel** | **an-rec1** | **an-rec2** | **water** | **exhaust1** | **exhaust2** |
| **Molar flow (mol/s)** | 2.28E+01 | 2.28E+01 | 2.14E+01 | 6.80E+00 | 3.07E+00 | 3.72E+00 | 3.72E+00 | 5.74E-01 | 2.48E+01 | 2.48E+01 |
| **Mass flow (kg/s)** | 6.61E-01 | 6.61E-01 | 6.15E-01 | 1.85E-01 | 8.36E-02 | 1.01E-01 | 1.01E-01 | 1.03E-02 | 7.09E-01 | 7.09E-01 |
| **Temperature (°C)** | 25.0 | 650.0 | 800.0 | 800.0 | 800.0 | 800.0 | 805.7 | 15.0 | 900.0 | 850.9 |
| **Pressure (bar)** | 1.1 | 1.1 | 1.1 | 1.1 | 1.1 | 1.1 | 1.1 | 1.0 | 1.0 | 1.0 |
| **Enthalpy (kJ/kg)** | -3.55E-01 | 6.60E+02 | 8.34E+02 | -8.71E+03 | -8.71E+03 | -8.71E+03 | -8.70E+03 | -1.60E+04 | -5.36E+02 | -5.98E+02 |
| **Entropy (J/kg*K)** | 1.38E+02 | 1.32E+03 | 1.48E+03 | 1.71E+03 | 1.71E+03 | 1.71E+03 | 1.71E+03 | -9.47E+03 | 1.67E+03 | 1.62E+03 |
| **Density (kg/cum)** | 1.29 | 0.41 | 0.34 | 0.33 | 0.33 | 0.33 | 0.33 | 1003.55 | 0.31 | 0.31 |
| **Mole fraction** | | | | | | | | | | |
| **H2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 9.51E-02 | 9.51E-02 | 9.51E-02 | 9.51E-02 | 0.00E+00 | 1.78E-09 | 1.78E-09 |
| **CO2** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 3.86E-01 | 3.86E-01 | 3.86E-01 | 3.86E-01 | 0.00E+00 | 5.64E-02 | 5.64E-02 |
| **O2** | 2.10E-01 | 2.10E-01 | 1.57E-01 | 2.46E-18 | 2.46E-18 | 2.46E-18 | 2.46E-18 | 0.00E+00 | 1.26E-01 | 1.26E-01 |
| **H2O** | 0.00E+00 | 0.00E+00 | 0.00E+00 | 4.50E-01 | 4.50E-01 | 4.50E-01 | 4.50E-01 | 1.00E+00 | 9.08E-02 | 9.08E-02 |

| | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| CO | 0.00E+00 | 0.00E+00 | 0.00E+00 | 6.85E-02 | 6.85E-02 | 6.85E-02 | 6.85E-02 | 0.00E+00 | 1.41E-09 | 1.41E-09 |
| CH4 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 1.63E-06 | 1.63E-06 | 1.63E-06 | 1.63E-06 | 0.00E+00 | 6.69E-38 | 6.69E-38 |
| N2 | 7.80E-01 | 7.80E-01 | 8.32E-01 | 2.57E-09 | 2.57E-09 | 2.57E-09 | 2.57E-09 | 0.00E+00 | 7.18E-01 | 7.18E-01 |
| NO | 0.00E+00 | 0.00E+00 | 0.00E+00 | 1.13E-17 | 1.13E-17 | 1.13E-17 | 1.13E-17 | 0.00E+00 | 1.29E-04 | 1.29E-04 |
| NO2 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 1.51E-27 | 1.51E-27 | 1.51E-27 | 1.51E-27 | 0.00E+00 | 1.90E-06 | 1.90E-06 |

| | **exhaust3** | **exhaust4** | **wat-in** | **wat-out** |
|---|---|---|---|---|
| **Molar flow (mol/s)** | 2.48E+01 | 2.48E+01 | 2.25E+02 | 2.25E+02 |
| **Mass flow (kg/s)** | 7.09E-01 | 7.09E-01 | 4.06E+00 | 4.06E+00 |
| **Temperature (°C)** | 330.1 | 90.0 | 65.0 | 75.0 |
| **Pressure (bar)** | 1.0 | 1.0 | 1.0 | 1.0 |
| **Enthalpy (kJ/kg)** | -1.21E+03 | -1.47E+03 | -1.58E+04 | -1.57E+04 |
| **Entropy (J/kg*K)** | 8.89E+02 | 3.43E+02 | -8.75E+03 | -8.62E+03 |
| **Density (kg/cum)** | 0.58 | 0.95 | 954.49 | 944.32 |
| **Mole fraction** | | | | |
| **H2** | 1.78E-09 | 1.78E-09 | 0.00E+00 | 0.00E+00 |
| **CO2** | 5.64E-02 | 5.64E-02 | 0.00E+00 | 0.00E+00 |
| **O2** | 1.26E-01 | 1.26E-01 | 0.00E+00 | 0.00E+00 |
| **H2O** | 9.08E-02 | 9.08E-02 | 1.00E+00 | 1.00E+00 |
| **CO** | 1.41E-09 | 1.41E-09 | 0.00E+00 | 0.00E+00 |
| **CH4** | 6.69E-38 | 6.69E-38 | 0.00E+00 | 0.00E+00 |
| **N2** | 7.18E-01 | 7.18E-01 | 0.00E+00 | 0.00E+00 |
| **NO** | 1.29E-04 | 1.29E-04 | 0.00E+00 | 0.00E+00 |
| **NO2** | 1.90E-06 | 1.90E-06 | 0.00E+00 | 0.00E+00 |

*Table 40: SOFC thermodynamic results*

| | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biogas5 | biogas6 | bioref | an-fuel | air0 | air1 | air2 | cat-ex | an-ex | burn-fuel | an-rec1 | an-rec2 | water | exhaust1 | exhaust2 | exhaust3 | exhaust4 | wat-in | wat-out | Wfuelcmp | Waircmp | Wreccmp | Wel | Qsofc | Qref |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| FUEL-CMP | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| CLEANING | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| FUEL-HX | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AN-MIX | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| REF-SPLIT | 0 | 0 | 0 | 0 | 1 | -1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| EXT-REF | 0 | 0 | 0 | 0 | 0 | 1 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 |
| REF-MIX | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| SOFC | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 1 | -1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | -1 | 0 |
| AIR-CMP | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 |
| AIR-HX | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| AN-SPLIT | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| REC-CMP | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| BURNER | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CHP-HX | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 |
| a. | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| b. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| c. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| d. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| e. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 594.2 | 0 | 0 | -1 | 0 | 0 |
| f. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 60.01 | 0 | -1 | 0 | 0 |
| g. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 412 | -1 | 0 | 0 |
| h. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| i. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.919 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| j. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.245 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| k. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.608 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 |
| l. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.175 | -1 | 0 |
| m. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1.211 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| n. | 0 | 0 | 0 | 0 | 0 | 0.983 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| o. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.972 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 |
| p. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 |

*Table 41: cost matrix of SOFC plant*

| | air0 | air1 | biogas0 | biogas1 | biogas2 | biogas3 | biogas4 | biomtn1 | biomtn2 | gas-out | gasrec1 | gasrec2 | gasrec3 | liquid1 | liquid2 | liquid3 | watrec | w1 | w2 | water0 | water1 | water2 | Wk1 | Qcool | Wk2 | Wpump | Qdry1 | Qdry2 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| CLEANING | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MIX1 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| K1 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| COOLER | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 |
| PUMP1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 1 | 0 | 0 |
| SCRUBBER | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 |
| DRY2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 |
| FLASH | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| V1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| DRY1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 |
| V2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| K2 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| STRIPPER | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| MIX2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 1 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| a. | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| b. | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| c. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| d. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 0 |
| e. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 |
| f. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 |
| h. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.15 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| i. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| j. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| k. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 |
| l. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 | 0 |
| m. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | -1 |
| n. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.089 | 0 | 0 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| o. | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0.432 | 0 | 0 | 0 | -1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |

*Table 42: upgrading improve cost matrix*

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