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# Integrated ORC-SOEC system for green hydrogen production from incineration of solid fuels

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**Abstract.** In the present paper, green hydrogen production from biomass and municipal solid waste is evaluated, targeting circular economy and green energy transition objectives pursued through waste-to-energy valorisation. In particular, the study consists in designing and modelling a hydrogen production system that integrates a solid oxide electrolyser (SOEC) and an organic Rankine cycle (ORC) powered by the combustion of biomass or municipal solid waste (MSW). The resulting efficiency, corresponding to 100.70 kg/h of hydrogen production in the optimal scenario, varies between 15.44% and 22.70% depending on operating conditions considered. Although on the one hand, this efficiency is far from the cold gas efficiency of an alternative gasification process, on the other, incineration shows a larger processing capacity and fewer restrictions in the waste categorization. Moreover, unlike traditional gasification treatments, the proposed concept can yield three outputs, namely pure hydrogen, electricity, and heat, allowing greater flexibility according to the end users' requests and achieving a cogeneration efficiency greater than 80% in every operating mode. Furthermore, some fairly simple preliminary treatments, aiming to reduce biomass or MSW water content, can increase the green hydrogen production rate.

## 1 Introduction

Hydrogen has the potential to play a key role in the decarbonisation of energy and industry: it is a versatile, clean-burning, and effective energy vector [1] that can help renewable sources breakthrough in the market and potentially shift the use of fossil fuels in order to lead the green transition to decarbonisation of the energy sector, transportation, and industry, globally lowering GHG emissions [1,2]. Thanks to these characteristics, hydrogen is the appropriate candidate for accomplishing the restrictive European objectives regarding climate change, but regrettably nowadays most of the hydrogen production is based on fossil fuels, so much research is focused on developing sustainable hydrogen production methods.

Hydrogen from waste (HfW) is a subset of the waste to energy idea, which emphasizes the specific goal of valorising scrap materials by generating hydrogen, thus integrating environmental sustainability with the circular economy approach.

Although conventional HfW pathway includes biomass/MSW gasification [3], the present paper designs an innovative and alternative hydrogen production line which employs waste incineration in order to overcome some of the typical gasification disadvantages [4]: indeed, incineration shows larger processing capacity and speed avoiding waste categorization [5], and the global diffusion of gasification



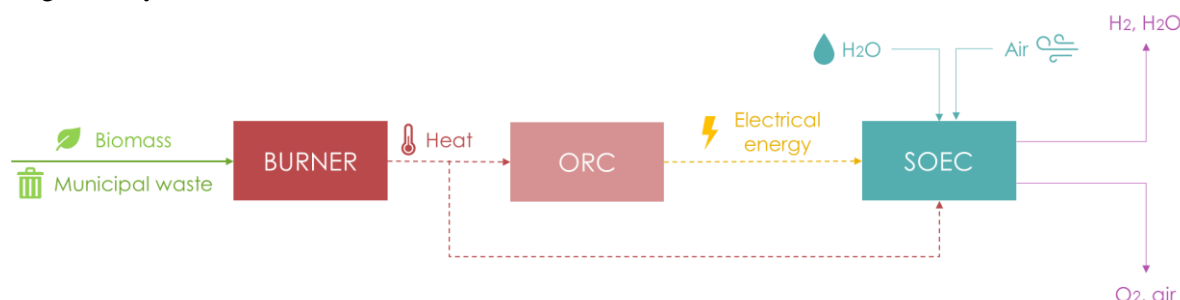
plants is certainly lower than that of waste incineration plants so the learning curve is therefore in favour of the latter, reducing their costs and the difficulties of construction and management. Moreover, gasification has only one end product - be it hydrogen, syngas or the energy derived from its use (in turbogenerators, combustion engines, ...) - while the plant conceived in this project is able to return three possible outputs simultaneously, namely hydrogen, electric energy and heat, allowing a greater flexibility according to the end users' requests.

In the proposed configuration, the heat generated by incineration of biomass or MSW is fed to an ORC, which is considered more suitable in low size (<5 MW) applications, object of study in this work, in comparison with steam power cycles that are usually employed in waste-to-energy facilities (>10 MW). The integrated SOEC is supplied by the electrical power generated by ORC and by part of the combustion heat, allowing the operation at high temperatures in order to decrease electrical needs. A sensitivity analysis is conducted, evaluating what are the effects on the system in case of the change of cell voltage and temperature. AspenTech's Aspen Plus® software is employed for the simulation of the system.

Once the plant layout is defined, the analysis consists of the identification of a reference case and a comparative with other solid fuels to address how the feedstock chemical compositions and design parameters influence the system performance. In this first evaluation a direct heat exchange between flue gases and ORC working fluid is contemplated. However, the conventional ORC plant configuration involves an intermediate heat exchange loop – diathermal oil typically: for the reference case is therefore evaluated how the oil circuit engraves on the energy consumption. The second part of the analysis focuses on the influence of single system parameters: cell voltage, cell temperature and moisture content of the fuel. Then the attention moves on the hydrogen purification process and on the opportunity of cogeneration, evaluating several heat recovery possibilities. Eventually, the last part of the work regards a preliminary analysis of hydrogen production cost.

## 2 Process description and methodology

The aim of the present paper is to design an innovative hydrogen production plant exploiting the thermal energy produced by a biomass/MSW-fed incinerator and a subsequent ORC plant which, employing part of the heat generated by the combustion, provides the electrical power required by the electrolyser. The modelling takes into account the principal components of the system, as shown in figure 1: i) incinerator, ii) ORC plant and iii) solid oxide electrolyser. Among the various types of electrolysers on the market, solid oxide electrolysers (SOEC) exploit the peculiar thermodynamics for which less electricity is required at temperatures between 600-800 °C (in comparison with ambient conditions) because thermal energy accounts for a portion of the total energy needed by the reaction  $\Delta H(T)$  [6]. Therefore, SOEC are well suited for integration with cycles powered by biomass or MSW incineration. The thermal power of the flue gases is fully exploited for powering ORC cycle, for generating steam in input to the SOEC, and for maintaining the electrolyser at the set temperature. As the thermal energy source is the same, the potential hydrogen production depends on the ORC size and the thermal integration system.



**Figure 1.** Conceptual design of the hydrogen production plant

The starting point is an overall system optimisation considering a fuel flow rate equal to 1 kg/s and the technological constraints found in the literature. The generated electrical power is then completely devoted to produce hydrogen: as a result, once the ORC system is optimized, the size of the SOEC, as well as the flow rates of water and air to the electrolyser, are assigned consequently. AspenTech's Aspen Plus® software is employed for process simulations.

### 2.1 Feedstock and Incineration

Incineration can effectively address the waste issue through thermo-valorisation of the matter. In this paper, incineration of biomass, MSW and RDF (Refused Derived Fuel) is considered. The combustion must generate enough heat both to power the ORC, in order to produce electrical power for the SOEC, and also to provide the thermal power required by the electrolyser. It has, therefore, a crucial role for the hydrogen production plant.

Usually, wood chips, trees and residual wood waste are the main components of typical biomass boiler, leading to a LHV between 12 and 20 MJ/kg [7], whereas MSW LHV is around 10 MJ/kg [8], depending on the composition of the unrecycled part of the municipal waste which can contain a large presence of inert compounds as a consequence of poor recycling quality. Moreover, the different composition of the raw material affects the composition of the flue gases, thus the plant performance. MSW can undergo different preliminary process before the incineration itself: for this reason, also RDF has been analysed as a possible fuel for the plant, while both biomass and MSW are employed with the same moisture content as received.

Besides, biomass and the MSW must be considered as non-conventional solid in Aspen Plus modelling, therefore the proximate analysis, ultimate analysis and sulphur analysis are necessary [9].

Table 1 and table 2 presents the proximate and ultimate analysis of the fuels used in the simulation. It is important to claim that the real feedstock composition is complex and depend on the area where the organic and solid wastes are collected.

**Table 1** Fuels proximate analysis

	Biomass	MSW	RDF
<b>Moisture (wet basis)</b>	20	35	20
<b>Volatiles</b>	80	-	-
<b>Fixed Carbon</b>	18.84	-	-
<b>Ash</b>	1.16	44.16	16.06
<b>Reference</b>	[10]		

**Table 2** Fuels ultimate analysis and characteristics

	Biomass	MSW	RDF
<b>Carbon</b>	51.19	30.80	40.90
<b>Hydrogen</b>	6.08	0.96	6.45
<b>Nitrogen</b>	0.2	1.09	1.51
<b>Chloride</b>	0.05	0.71	0.65
<b>Sulphur</b>	0.02	0.78	0.43
<b>Oxygen</b>	41.3	21.5	34
<b>LHV, MJ/kg</b>	19.09	11.99	18.64
<b>Average Density, kg/m<sup>3</sup></b>	250	350	400
<b>Reference</b>	[10]		

To simulate the incinerator, different stages of the process are modeled separately. Gasification, led at 400 °C, is followed by combustion which takes place with an excess of air (considered as a mixture of oxygen, nitrogen and argon); this air inlet is preheated by the exhausted flue gases before going to filters

and chimney. The air flow rate is modulated in order to have 6 % of oxygen molar fraction in flue gases (on dry basis).

An additional mixer, placed after a cyclone that separates ashes, blends the fumes with a stream spilled just before the chimney, in order to cool down the gas from combustion: the flue gas temperature is set at 950 °C for biomass [7] and 850 °C [8] for MSW and RDF.

While pressure drops are neglected, thermal losses are estimated at 1% of the thermal power produced by the combustion. Several heat exchangers are used along the flue gas line to exploit the heat for each step of the hydrogen production process, making sure to obtain an outlet temperature towards the chimney around 130 °C [11] to avoid the condensation of acid substances and consequent corrosion of the metal components.

## 2.2 Organic Rankine cycle

Part of the heat generated by the incineration is then used in an ORC power plant. The layout is composed by feeding pump, regenerator, economizer, evaporator and superheater, turbine, and condenser (namely PUMP-ORC, REG, ECO, EVA, TURBINE and COND-ORC in figure 3).

Among the suitable organic working fluid [12], toluene is chosen in this paper as it can reach a turbine inlet temperature (TIT) of around 300 °C, thus allowing to obtain a good power cycle performance.

To optimize the ORC, an initial analysis is carried out to evaluate the performance at varying turbine inlet pressure, set the TIT at 315 °C (compatible with the toluene thermal stability constraint). A minimum internal temperature approach (MITA) of 25 °C set in economizer determinates the flow rate. In most ORC plants, heat is transferred from the gaseous combustion products to the working fluid via a closed circuit containing diathermic oil used as a heat-transfer medium. However, in the first layout presented in this paper, the circuit with thermal oil is not considered, since applications with a direct heat exchange recently developed [13] can increase the recovered thermal power.

## 2.3 Electrolyser

A simplified 0D model is developed in Aspen Plus®: a stoichiometric reactor is used for SOEC modelling, since all the reactions that occur in a SOEC can be defined along with the utilization factor. Concurrently, the electrical part of the stack, with the definition of voltage, current density and overpotentials, is implemented; in this way, the progress of the reaction is linked to the electrical operating parameters.

The minimum electrical power input required by a SOEC is directly proportional to Nernst potential  $V_{Nerst}$ . However, when the system is under operation and current is applied, the operating voltage deviates from the ideal one due to irreversible losses, which are called overpotentials [14]. Since the detailed modelling of all overpotential losses is complex, a simplifying approach is applied to bundle the effects of all types of resistances [15]. In a lumped model, a single area-specific resistance (ASR) which is the resistance corresponding to 1 cm<sup>2</sup> of the cell, is considered as the total overpotential loss in a SOEC. The operating voltage can be easily obtained:

$$V = V_{Nerst} + jASR \quad (1)$$

$$V_{Nernst} = \frac{\Delta G(T)}{zF} + \frac{RT \ln(K)}{zF} \quad (2)$$

$$K = \prod_{i=1}^r \left( \frac{P_i}{P_0} \right)^{\nu_i} = \left( \frac{P_{H_2}}{P_0} \right) \left( \frac{P_{O_2}}{P_0} \right)^{0.5} \left( \frac{P_{H_2O}}{P_0} \right)^{-1} \quad (3)$$

where  $j$  is the current density and  $ASR$  is the area-specific resistance,  $\Delta G(T)$  is the standard Gibbs free energy change of reactions at standard pressure,  $R$  is the universal gas constant,  $K$  is the thermodynamic equilibrium,  $z$  is the number of electrons involved in the reaction (for water electrolysis  $z = 2$ ),  $F$  is Faraday's constant,  $P_i$  the average partial pressures of the species between inlet and outlet of the SOEC.

Experimental investigation on a commercially available solid oxide electrolyser led by Mottaghizadehab et al. [15,16] provides equation (4), on which is based the polarization curve of the present case study.

$$ASR = 0.30813 + 767370 \exp(-0.013295T) \quad (4)$$

The fundamental parameters for the SOEC's operating conditions can be computed as follows:

$$V_{TN} = \frac{\Delta H(T)}{zF} \quad (5) \quad I = 2F\dot{n}_{H_2O} \quad (6)$$

$$I_{SOEC} = U_f I \quad (7) \quad \dot{W} = V I_{SOEC} \quad (8)$$

$$|\dot{Q}| = |\dot{W} - \Delta H| = |I_{SOEC}(V - V_{TN})| \quad (9)$$

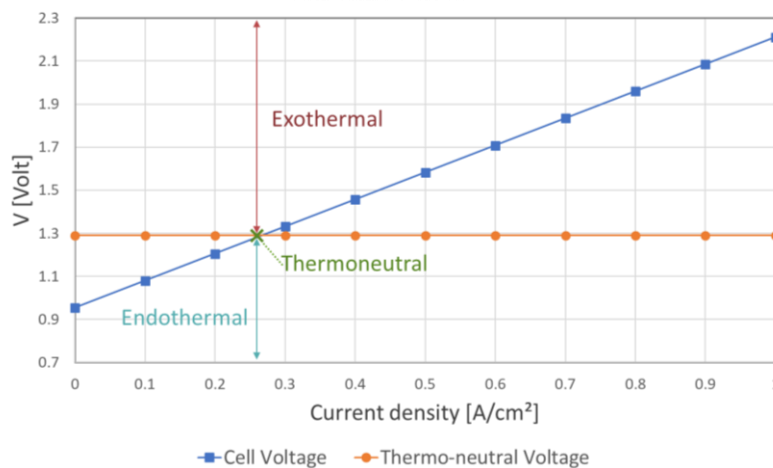
$V_{TN}$  is the precise voltage sufficient to drive the cell reaction and to provide the heat required to maintain a constant temperature, avoiding the need of additional thermal power to the cell.  $I$  and  $I_{SOEC}$  are the maximum theoretical current and the operating current of the SOEC considering the utilization factor  $U_f$ , which quantifies the fraction of reactants that undergo the electrochemical reaction;  $\dot{n}_{H_2O}$  is the molar flow in input to SOEC.  $\dot{W}$  and  $\dot{Q}$  are the electrical and thermal power required by SOEC.

The cell has an exothermic, endothermic, or isothermal behaviour depending on the current density chosen, and thus influences the thermal and electrical demands from the boiler and power plant, along with the performance of the electrolyser. Following the studies of Cai et al [17], the following parameters shall be considered:

- To avoid the material oxidation, part of hydrogen generated must be recirculated to the cathode side; 10% molar fraction of  $H_2$  in the cathode is an acceptable value.
- On the anode side, an air flow is requested to maintain a constant and homogeneous temperature in the cell and to sweep the produced oxygen. Air enters the SOEC at operating temperature and the flow is defined by the air ratio  $AR$  defined in (10), based on the molar flow of oxygen in the air  $\dot{n}_{O_2,in}$  and produced oxygen  $\dot{n}_{O_2,an}$ . An acceptable  $AR$  value is between 0.4 and 14 to keep the energy cost low and ensure a sufficient nitrogen flow to avoid the corrosion of metallic components.

$$AR = \frac{\dot{n}_{O_2,in}}{\dot{n}_{O_2,an}} \quad (10)$$

Starting from the eq. (1) and (4), Figure 2 shows the resulting polarisation curve at 750 °C and 1.01325 bar, using  $U_f$  to 0.85 and  $AR$  to 10.



**Figure 2.** SOEC Polarization curve at 750°C and 1.01325 bar,  $U_f=85\%$ ,  $AR=10$  and hydrogen molar fraction at inlet stack is set to 10%



In the proposed configuration, the useful heat  $\dot{Q}_{abs}$  comprehends the heat transferred to the SOEC feeding water and air, besides the heat required by the ORC  $\dot{Q}_{in}$ . In this first analysis, the plant is investigated with the SOEC operating in endothermic conditions.

The produced hydrogen  $\dot{m}_{H_2,prod}$  refers to the mass flow rate in stream “H2-W4” (figure 3) and the lower heating value  $LHV_{H_2}$  is equal to 120 MJ/kg.

All the necessary data are consistent with typical parameter for small-medium ORC power plant [11,18] and SOEC system [15]; the following table (table 3) sums up the key parameters for the reference case.

**Table 3.** Reference case key parameters

SOEC temperature	750 °C	Fuel mass flow rate	1 kg/s
SOEC pressure	1.01325 bar	Inlet Turbine Temperature	315 °C
SOEC current density	0.2 A/cm <sup>2</sup>	O <sub>2</sub> in flue gases (dry basis) [11]	6 % <sub>mol</sub>
SOEC Air ratio	10	Biomass boiler efficiency [19]	0.88-0.90
SOEC Hydrogen at cathode inlet	10 % <sub>mol</sub>	MSW boiler efficiency [19]	0.75-0.78
SOEC thermal behaviour	Endothermal	ORC working fluid	Toluene
Inlet streams conditions	Std conditions	ORC Condenser conditions	50°C sat.liq.
		Combustion Thermal Loss	1%

The sensitivity analysis is carried out by investigating three different parameters: the cell voltage at constant temperature and pressure, the temperature at constant pressure, the influence of the moisture content in the feedstocks.

### 3 Results and discussion

#### 3.1 System performance

For the reference case, the biomass is considered, and the analysis is made by maximizing the efficiency of the power plant. The simulation shows (table 4) that the thermal power generated by the combustion  $\dot{Q}_{comb}$  is 14.78 MW: 12.28 MW are absorbed by the power plant and 1.1 MW by the air and water exchangers leading to a boiler efficiency equal to 92.26%. In the ORC system the flow rate of fluid is equal to 20.97 kg/s, obtaining a net power of 3.22 MW and 26.26% ORC efficiency.

**Table 4.** Results for different fuels

	Biomass	MSW	RDF	Unit
$\eta_{system}$	22.70	15.44	19.94	
$\eta_{SOEC}$	72.63	72.63	72.63	
$\eta_{SOEC,no\ steam}$	97.21	97.21	97.21	%
$\eta_{ORC}$	26.26	26.04	26.01	
$\eta_{boiler}$	92.26	63.24	81.74	
$\dot{Q}_{comb}$	14.78	6.94	14.28	MW
$\dot{m}_{H_2}$	100.70	32.18	85.44	kg/h
$A_{SOEC}$	1338.69	427.81	1135.96	m <sup>2</sup>

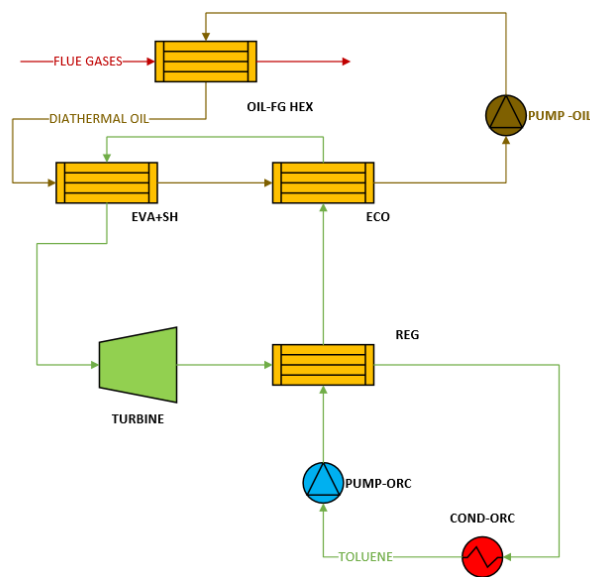
The values set for the SOEC allow the electrolyser to work with a voltage of 1.20 V and a current of 2.67 MA with a cell area  $A_{SOEC}$  of 1338 m<sup>2</sup>. It follows that the mass flow rates are 0.29 kg/s for water inlet  $\dot{m}_{H_2O}$  and 9.57 kg/s for air inlet  $\dot{m}_{AIR}$ . By setting a current density equal to  $0.2 \frac{A}{cm^2}$ , the conditions are endothermic and the necessary thermal power is -0.227 MW. Thus, the SOEC efficiency is equal to 72.63% and the efficiency system is 22.70%, corresponding to a 100.70 kg/h hydrogen production.



The conducted simulations with MSW and RDF show a  $\eta_{system}$  and  $\dot{m}_{H_2}$  lower than the biomass case (table 4). This is likely due to the greater LHV of biomass compared to MSW and because of the different selected maximum temperature of the boiler. However, using fluidized bed for waste incineration, it may be possible to achieve higher temperatures [8].

The efficiency of the electrolyser is the same for all the cases analysed, because once the enthalpy of the reaction is fixed, the quantity of hydrogen theoretically produced is fixed too. The only parameters that can affect the performance are the utilization factor and the fraction of hydrogen that is recirculated. If the amount of heat needed by the water  $\dot{Q}_{steam}$  is not considered in the SOEC efficiency, the yield  $\eta_{SOEC, no\ steam}$  increases up to 97.21%.

As previously mentioned, the well-know and consolidated technology for biomass CHP power plant involves an intermediate heat exchanger using diathermal oil interposed between the hot flue gases and the organic fluid [13] (figure 4). To compare the traditional layout with the one proposed, a simulation is carried out. The diathermal oil circuit is usually composed of a circulating pump and two heat exchangers (oil – flue gases and oil – working fluid): the oil is pumped at 5/6 bar at temperature around 280 °C and heated up to around 320-340 °C [13], then the accumulated thermal power is transferred to the organic fluid. For this evaluation, considering the optimization of the power plant and self-sustaining system, the diathermal oil is fed to the pump at 280 °C and 5 bar and heated up at 340 °C. The pump (PUMP-OIL) has an efficiency of 85% and a pressure ratio of 1.1, the MITA for the oil flue gases heat exchangers (OIL-FG HEX) is set to 15 °C, meanwhile the parameters for the other components do not vary. For the simulation, the studied oil is the THERMIOL VP-1, an eutectic mixture of diphenyl oxide (73.5%) and biphenyl (26.5%) [20].



**Figure 4.** Diathermal oil circuit design

The simulation points out that the net electric power is 3 MW, corresponding to a hydrogen production rate of 93.71 kg/h. The flow of circulated oil is 84.36 kg/s with a consumption of 6.34 kW power in the pump circuit. The system efficiency is 21.09%, less than the direct heat exchange case, since the bottom temperature of the circuit oil is set at 280 °C: maintaining the MITA for the oil – flue gases exchanger at 15°C and reducing the bottom temperature it would be possible to exploit more thermal power available from the combustion.

Comparing the system against other production processes that implement biomass as raw material, the efficiency is lower than gasification (65%) [21]. A case study for waste gasification proposed in [22] led to a hydrogen production of 300 kg/h with a yield of 167.5 kg of hydrogen per dry tonne of waste, against a 29.97 kg of hydrogen per dry tonne of biomass yield in the present work.

### 3.2 Sensitivity analyses

#### 3.2.1 Voltage

Recalling eq. (6) (8) and (9), it is possible to highlight that, at a fixed electrical energy input, a decrease in the cell voltage increases the electric current and, therefore, the water flow rate that can be converted. However, selecting a low voltage can be counterproductive for the thermal behaviour of the cell itself (eq. (9)), leading to a severe stress and a high demand of cooling air. Consequently, the ideal operating condition is close to the thermoneutral voltage, in endothermic conditions. This allows obtaining a better efficiency, an easier cell temperature control and a reduced air demand.

The voltage variation not only affects the efficiency of the system but also the system components, especially the products-water heat exchanger called "HEW1". In endothermic behaviour, the parameters assigned to the components do not affect the thermodynamics of the system, since the products are at a lower temperature than the reactants and the products enter the hot side of the "HEW1" exchanger through the "O2-A1" and "H2-W1" streams. As the MITA is fixed, it follows that the temperature of the stream "W3" is lower than that of the cell and consequently the water in the stream "W3" is brought to the operating temperature of the SOEC with the "HEW2" heat exchanger.

In exothermic conditions, particular attention must be given to the air ratio and the MITA of the "HEW1" exchanger. Two simulations are made with a voltage higher than 6.58% compared to the thermoneutral one (table 4), showing these two considerations: i) if the air ratio remains low, the MITA value of the heat exchanger must increase ii) if the MITA value is to be kept low, the air ratio must increase. In exothermic conditions, the cell can sustain itself and a further alternative is to eliminate the "HEW2" exchanger and design the "HEW1" exchanger to set the outlet temperature equal to that of the SOEC, changing the MITA and AR value.

To corroborate the above, four cases are studied in which the cell voltage (table 5), the AR and the MITA of the HEW1 block are varied. All other parameters are the same as the reference case.

**Table 5.** Data and result for the four cases investigated for the voltage sensibility analysis

	Reference Case	Case 1	Case 2	Case 3	Case 4	Unit
$MITA_{HEW1}$	10	10	10	10	25	°C
$AR$	10	10	5	22	10	-
$\Delta V/V$	-6.58	0	0	6.58	6.58	%
$V$	1.205	1.290	1.290	1.374	1.374	Volt
$j$	0.200	0.268	0.267	0.335	0.335	A/cm <sup>2</sup>
$\Delta T_{SOEC}$	-19.34	0	0	9.11	17	°C
$\eta_{system}$	22.70	21.69	21.97	20.41	20.77	
$\eta_{SOEC}$	72.63	72.63	72.62	66.13	66.12	
$\eta_{ORC}$	26.26	26.30	26.36	26.19	26.33	%
$\eta_{Boiler}$	92.26	92.27	92.27	92.27	92.27	
$\dot{m}_{H_2}$	100.70	96.23	97.46	90.56	92.13	
$\dot{m}_{H_2O}$	1040.27	994.20	1007.09	935.62	952.03	kg/h
$\dot{m}_{AIR}$	34452.20	32926.47	16676.79	68170.18	31529.72	

The parameter  $\Delta V/V$  can be defined as follows:

$$\Delta V/V = (V - V_{TN})/V_{TN} \quad (26)$$

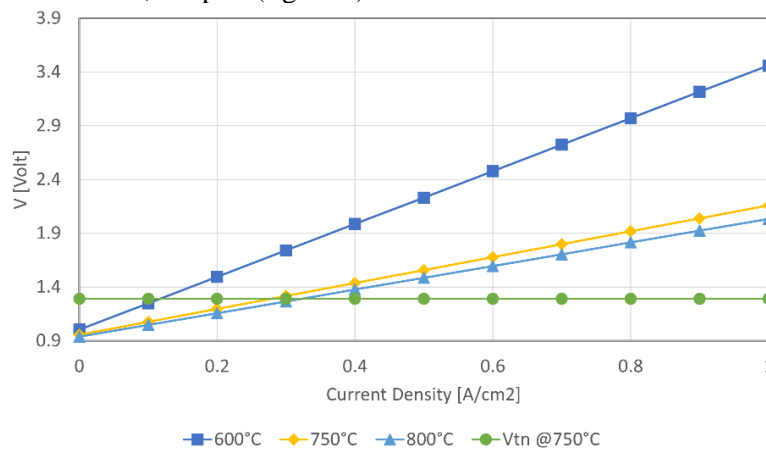
It is useful to define the thermal behaviour of the cell:  $\Delta V/V = 0$  for isothermal,  $\Delta V/V > 0$  for exothermic and  $\Delta V/V < 0$  for endothermic. It is noted that the higher efficiency is obtained in the endothermic case while, increasing the voltage, the performance decreases. Case 1 and case 2 show the behaviour of the system when the cell is isothermal. Since in this conditions air acts just as sweeping gas, cutting the AR improves the efficiency. Cases 3 and 4 refer to exothermic conditions and two parameters varies: the MITA of the HEW1 and the air ratio. Once again, the performance improves when the air ratio remains low.

In conclusion, with an exothermic behaviour, the efficiency of the system drops, and the cell undergoes greater thermal and mechanical stress. Accordingly, in the next analyses, only endothermic condition is investigated.

### 3.2.2 Temperature

A first consideration on the effect of temperature at constant pressure (1.01325 bar) on the system can be made by considering how the polarization curve varies at different the temperatures.

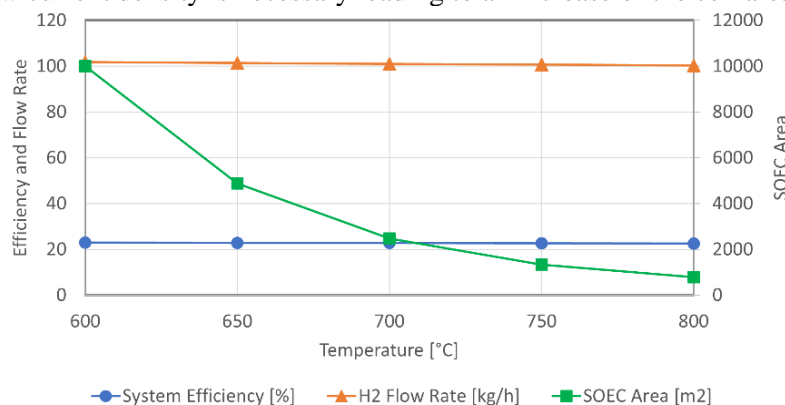
The slope of the curve is the parameter ASR which, as defined in (4), decreases as temperature increases; for this reason, the plot (figure 5) tends to have a minor inclination at higher temperatures.



**Figure 5.** Polarization curve at different temperature. SOEC parameter: 1.01325 bar,  $U_f=85\%$ ,  $AR=10$  and hydrogen molar fraction at inlet stack is set to 10%

The results in figure 6 highlight that the performance of the system does not significantly change at different temperatures. Indeed, SOEC and ORC are strictly coupled, so even if the SOEC working temperature decreases, the thermal power required by SOEC inlet heating remains almost constant, since it depends not only on the temperature drop but also on the water mass flow: lower SOEC working temperatures involve more thermal power available for ORC and, consequently, a larger electrical power in input to the SOEC itself, which implies a greater water flow inlet.

At an operational level, it can be highlighted that at lower temperatures the cell is less thermally stressed, and its lifespan would increase. Nevertheless, to achieve endothermic conditions at lower temperatures, a low current density is necessary leading to an increase of the cell area and, therefore, the cell cost.



**Figure 6.** Analysis on temperature – Results

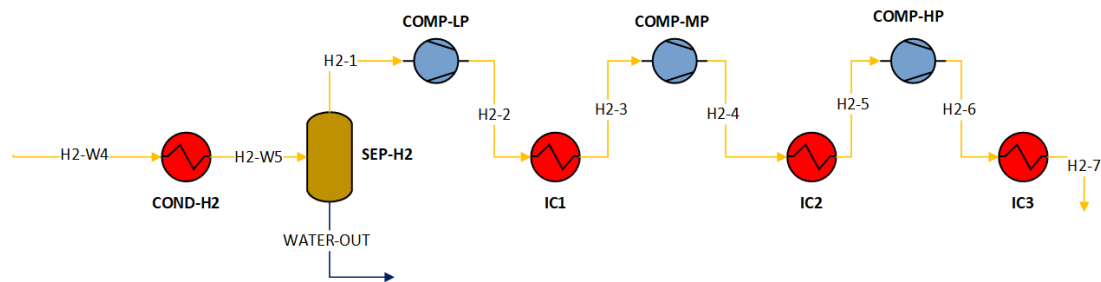
### 3.2.3 Moisture content

In the reference case, the combustion is simulated using raw material with a moisture content (MC) as received. However, it may be relevant to investigate the effect of a reduction of the moisture content on the heating value of the solid fuel during combustion. Using different drying techniques, the material can achieve a minimum moisture content of 10% for biomass [23] and 20% for waste [24], so these minimum values are considered in the sensitivity analysis.

The results prove that, thanks to a lower moisture content, the thermal power generated by combustion increases and consequently the electrical energy and the amount of produced hydrogen. Since the plant is optimized, the yields of the system, of the SOEC, of the ORC and of the boiler do not vary. Therefore, with a 50% reduction of MC in biomass and 42% in MSW, there is an increase in hydrogen production of 15% (biomass) and 33% (MSW).

### 3.3 H<sub>2</sub> purification and pressurization

An additional analysis is the implementation of hydrogen purification and pressurization processes. There are several separation and storage methods available [25]. For the sake of the discussion, it is assumed that the products stream is condensed (COND-H2) at ambient temperature to separate the liquid water from the gaseous hydrogen, then the latter is compressed using an intercooled compression station (figure 7) and sent to the grid, blended with natural gas, at 50 bar [26] (average pressure for the Italian natural gas pipelines).



**Figure 7.** Intercooler compression station design

The intercooled compression is simulated with three compressors (COMP-LP, COMP-MP and COMP-HP) all having an isentropic efficiency of 88%, organic one of 96% and the same pressure ratio [27]. After each stage, the intercoolers (IC1, IC2, IC3) chill the fluid at ambient temperature [28].

A preliminary step can be conducted if the purification and compression process is considered outside the system boundaries, meaning the electrical power for the compressors is received from the grid. In this scenario, the mix fluid (0.066 kg/s, 14% molar fraction of water and 86% of hydrogen) is cooled from 73°C to 25°C, then purified and compressed, leading to a total compression work of 210 kW and a thermal power in the intercoolers of 200 kW.

**Table 6.** Comparison between not self-sustainable and self-sustainable compression main results

	Not self-sustainable compression	Self-sustainable compression	Unit
$\eta_{system}$	22.70	21.57	%
$\dot{W}_{net}$	3.22	3.07	MW
$\dot{W}_{compression}$	210	183	kW
$\dot{m}_{H_2}$	100.70	95.69	kg/h

The simulation carried out with the integration of purification and pressurization processes, still considering optimal conditions and self-sustaining system, reveals (table 5) that, at the maximum ORC cycle pressure of 29 bar, the net generated electric power is equivalent to a hydrogen flow rate of 95.69

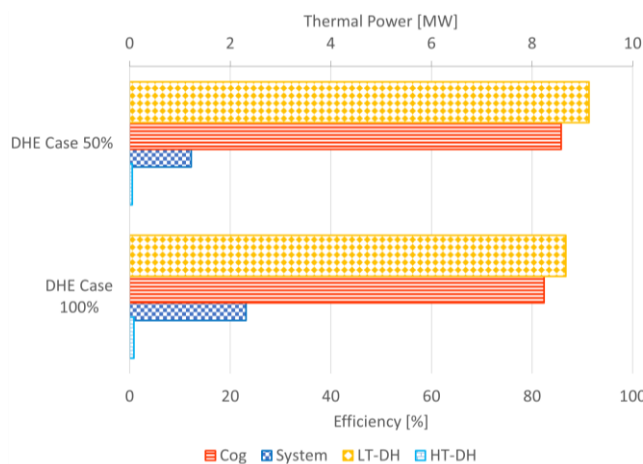
kg/h, a 21.57% system efficiency, a compression work of 183 kW and a intercoolers thermal power of 176 KW.

### 3.4 Cogeneration mode

The typical configuration of biomass and MSW incineration plant is the Combined Heat and Power (CHP) mode. In this section, the possibility of exploiting thermal power ( $\dot{Q}_{th}$ ) for district heating (DH) is presented. In the EU scenario, there are more than 6000 different DH systems, each one with different temperature and pressure ranges. However, there are two main categories of DH, both investigated: high-temperature DH (HT-DH) (pressurized water above 100 °C) [29] and low-temperature DH (LT-DH) (water at 40 °C) [30]. The accessible thermal power in the ORC condenser is considered as low temperature source, meanwhile the high-temperature source is the thermal power extrapolated from the flue gases after the ORC cycle, compatible with the no acid condensation constraint. To consider all the energy outputs, cogeneration efficiency has been defined in (27):

$$\eta_{cog} = \frac{\dot{m}_{H_2} LHV_{H_2} + \dot{W}_{ee} + \dot{Q}_{th}}{\dot{m}_{fuel} LHV_{fuel}} \quad (27)$$

Two cases are examined: in the first scenario 100% of electrical power produced by ORC is used for SOEC while in the second 50% of the electrical output is sent to the grid; both are referred to biomass feeding. The results prove (figure 8) that cogeneration efficiency is very high in all scenarios, varying between 80% to 85%. The richest thermal source for DH is the low temperature one, because the high temperature one is penalised by the power cycle that fully utilizes the thermal energy of the flue gases.



**Figure 8.** Cogeneration Mode – Results for the Direct Heat Exchange (DHE) Case

### 3.5 Techno-economic analysis

In this section, a preliminary analysis of the Levelized Cost of Hydrogen (LCOH) is carried out. The study is based upon several assumptions: i) results are referred to the reference case and considering hydrogen separation and compression, ii) material collection and handling related cost and fuel cost are not taken into account, iii) interest rate  $i$  is supposed 5%, iv) capacity factor  $cp$  for the power plant is 63.5% [31], v) plant lifetime  $N$  is 25 years [32] while SOEC lifetime is 5 years [33]. Indeed, the strong degradation due to the high-temperature environment has a negative influence on the electrolyser lifespan, while SOEC auxiliaries (compressor, pump and heat exchangers), which do not undergo these burdensome conditions, are assumed to last 25 years.

LCOH in €/kg is calculated as follows [34]:

$$LCOH = \frac{C_{capital\ cost} + C_{O\&M}}{M_{annual,H_2}} \quad (28)$$

$$M_{annual,H_2} = \dot{m}_{H_2} cp H \quad (29)$$

$$C_{capital\ cost} = C_{total} CRF \quad (30)$$

$$CRF = \frac{i(1+i)^N}{(1+i)^N - 1} \quad (31)$$

$$C_{total} = \sum C_{eq,i} + C_{add} + C_{BoP} \quad (32)$$

where  $C_{capital\ cost}$  is the investment cost rate in k€/y, depending on the total investment cost  $C_{total}$  (in k€) and the Capital Recovery Factor  $CRF$ .  $C_{total}$  is comprehensive of the cost for all the equipment  $\sum C_{eq,i}$ , additional cost  $C_{add}$ , such as construction and engineering fees, and the balance of plant (BoP) cost  $C_{BoP}$ .  $C_{O\&M}$  are operation and maintenance costs. The annual hydrogen production  $M_{annual,H_2}$  in kg/y is calculated as a product of the capacity factor, the hourly hydrogen production and the total amount of hours in a year  $H = 8760\ h/y$ .

The equations for the cost of the equipment are consistent with the literature [35], [34], [36], [37].

For the ORC power plant,  $C_{add}$ ,  $C_{O\&M}$  and  $C_{BoP}$  are the 66%, 5% [38] and 40% [37] of the cost of equipment, while for the SOEC, the  $C_{BoP}$  is the 66% of the cost of equipment, the installation cost is the 20% of the sum of BoP and equipment cost [39]. Labour, maintenance, and other costs are estimated as 0.3%, 5%, and 1% of the sum of equipment costs (SOEC system, BoP, and installation) [39].

LCOH is determined for three different scenarios: i) basic reference case, ii) reference case where the energy for intercooled compression is sourced from the grid and iii) reference case with self-sustainable intercooled compression (avoiding external contributions). For the second one, the electric energy price is set to 0.13 €/kWh (average value from October 2020 to April 2022 of the Italian price indicated in IPEX - Italian Power Exchange [40]).

In the first scenario, LCOH results in 5.56 €/kg, in the second one the cost increase to 6.12 €/kg (+8%) and in the self-sustainable scenario the LCOH is 6.22 €/kg (+10% comparing to the first scenario, +2% comparing the second one). This minor difference between the latter ones occurs because in the self-sustainable scenario the turbomachinery cost is higher since more hydrogen must be compressed.

It is important to emphasize that the present estimated LCOH should be considered only as a lower limit: as mentioned above, the pressure drops and the wastes handling costs are not considered, as well as BoP energy consumption.

Biomass and residual waste gasification is considered the most mature and economical process in hydrogen-from-waste sector, so it is logical to compare this method and the innovative route presented in this paper. Depending on size plant, waste gate fee or feedstock, gasification LCOH is in the range between 2.3-5.2 \$/kg for biomass and 1.4-4.8 \$/kg for wastes [41] (average exchange rate Euro – US Dollar is 1.16 €/€ in the June 2020 - June 2022 period [42]).

#### 4 Conclusions

In this work, a hydrogen production plant exploiting a solid oxide electrolyser cell and a power plant, which works with an organic fluid instead of steam, is analysed. This cycle is powered by the combustion of biomass and MSW.

The parameters used for the model are consistent with the literature and a reference case is defined, in which the cell operates at a temperature of 750 °C and a pressure of 1.01325 bar in endothermic conditions. The simulation produces a system efficiency of 22.70%, a SOEC cell efficiency of 72.63% and an ORC cycle efficiency of 26.26% for the biomass case and the hourly production of hydrogen was 100.70 kg/h, while the performance for residual wastes is lower due technological constraint. A preliminary analysis of the LCOH shows a hydrogen production cost of 5.56 €/kg (reference case), 6.12 €/kg (not self-sustainable compression) and 6.22 €/kg (self-sustainable compression).

The sensitivity analyses reveal that the best performances are achieved in endothermal condition and low temperature, however the cell voltage must be close to the thermo neutral one and at low temperature the cost of the cell increase.

Although the performance of the system is significantly lower than the most widespread hydrogen production technologies, other aspects must be considered. First, the ORC system can be further optimized by testing other working fluids other than toluene, characterized by a higher thermal stability, thus yielding comparatively higher conversion efficiencies. Moreover, while in this analysis the focus is on the production of hydrogen, the proposed layout, in principle, allows to obtain three useful effects from the combustion of solid fuels: electricity and thermal energy (CHP mode) and hydrogen achieving an overall efficiency up to 85%.

## Nomenclature

<b>Acronyms</b>		<b>Roman letters</b>	
<b>0D</b>	Zero dimension	$\dot{Q}$	Thermal power, <i>Watt</i>
<b>BoP</b>	Balance of Plant	$U_f$	Utilization Factor, -
<b>CHP</b>	Combined Heat and Power	$\dot{W}$	Mechanical and electrical power, <i>Watt</i>
<b>DH</b>	District heating	$\dot{m}$	Mass flow, <i>kg/s</i>
<b>DHE</b>	Direct heat exchange	$\dot{n}$	Mole flow, <i>mol/s</i>
<b>EU</b>	European Union	$A$	Area, <i>m<sup>2</sup></i>
<b>GHG</b>	Greenhouse gases	$AR$	Air Ratio, -
<b>HfW</b>	Hydrogen from waste	$ASR$	Area specific resistance, <i><math>\Omega cm^2</math></i>
<b>HT-DH</b>	High-temperature district heating	$C$	Cost, <i>€/y</i>
<b>IPEX</b>	Italian Power Exchange	$CFR$	Capital Recovery Factor, -
<b>LCOH</b>	Levelized Cost of Hydrogen	$F$	Faraday's constant, <i>C/mol</i>
<b>LT-DH</b>	Low-temperature district heating	$H$	Total hour in a year, <i>h</i>
<b>MC</b>	Moisture content	$I$	Electric current, <i>A</i>
<b>MITA</b>	Minimum internal temperature approach	$K$	Thermodynamic equilibrium constant, -
<b>MSW</b>	Municipal solid waste	$LHV$	Lower Heat Value, <i>J/kg</i>
<b>RDF</b>	Refuse-derived fuel	$M$	Mass production, <i>kg/y</i>
<b>SOEC</b>	Solid oxide electrolyte cell	$N$	Lifetime, <i>years</i>
<b>TIT</b>	Inlet turbine temperature	$P$	Pressure, <i>bar</i>
		$R$	Universal gas constant, <i>J/molK</i>
		$V$	Voltage, <i>Volt</i>
		$cp$	Capacity factor, -
		$i$	Interest rate, %
		$j$	Current density, <i>A/cm<sup>2</sup></i>
		$z$	Number of electrons, -
<b>Chemical formula</b>		<b>Greek letters</b>	
$H_2$	Hydrogen	$\Delta G(T)$	Variation of free Gibbs's energy of reaction at temperature T, <i>J/mol</i>
$H_2O$	Water	$\Delta H(T)$	Variation of enthalpy of reaction at temperature T, <i>J/mol</i>
$O_2$	Oxygen	$\Delta T$	Temperature gradient, <i>°C</i>
		$\eta$	Efficiency, %
		$\nu$	Stoichiometric coefficient of the reaction, -
<b>Subscripts</b>			
<b>Nernst</b>	Nernst Equation		
<b>SOEC</b>	SOEC		
<b>TN</b>	Thermo neutral		
<b>add</b>	Additional		
<b>an</b>	Anode		
<b>annual</b>	Annual		
<b>cog</b>	Cogeneration		
<b>comb</b>	Combustion		
<b>compressor</b>	Compressor		
<b>ee</b>	Electric Energy		
<b>eq</b>	Equipment		
<b>fan</b>	Fan		
<b>fg</b>	Flue gases		
<b>fuel</b>	Fuel		
<b>in</b>	Input parameter		
<b>max</b>	Maximum		
<b>net</b>	Net value		
<b>prod</b>	Produced		
<b>pump</b>	Pump		
<b>steam</b>	Steam		
<b>system</b>	System		
<b>th</b>	Thermal		
<b>turbine</b>	Turbine		



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