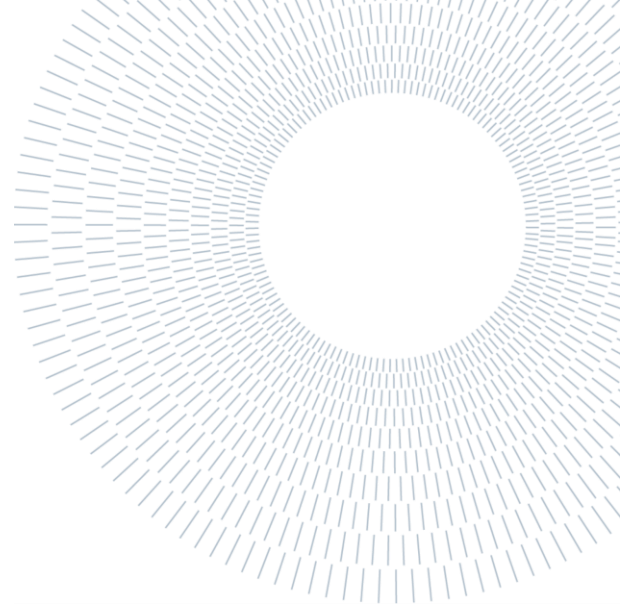




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EXECUTIVE SUMMARY OF THE THESIS

Alternative configurations for thermal integration in a Solid Oxide Cell (rSOC) system

TESI MAGISTRALE IN ENERGY ENGINEERING – INGEGNERIA ENERGETICA

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ACADEMIC YEAR: 2020-2021

1. Introduction

Finding solutions for climate change mitigation and discovering new ways to reduce greenhouse gas emission (GHG) are among the greatest challenges of our time. One of the most important strategies to achieve a significant reduction of GHG emissions involves a change in the power generation sector, for instance by developing an electric power plant infrastructure with an increasingly high penetration of power from wind power plants and solar PV. These power plants' production is highly dependent on the energy source availability, for this reason they are also referred to as Variable Renewable Energy (VRE). Timing of variable electricity supply and demand is not well matched (neither over the day nor between seasons). Integration of an increasing share of VRE (above 40% of the electricity mix) will enhance the need for operational flexibility. Increased electrification and limited storability of electricity requires adequate storage solutions. Among the various types of energy storage systems, chemical storage systems are gaining importance. One of the most promising is related

to hydrogen technologies, electrolysis (EC) and fuel cell (FC). Among all types of fuel cell, the high temperature Solid Oxide Cells are an interesting choice because of the high efficiencies and the possibility to be operated reversibly despite the still low Technology Readiness Level. Among the SOC technology, reversible cells have the advantage that with a unique electrochemical device the oxidation of H_2 and the reduction of water by electrolysis can be carried out. The exothermic nature of oxidation reactions in fuel cell mode and endothermic behavior during electrolysis makes way for the designing of an efficient thermal management system. A number of plant configurations have been compared in the literature and, this work presents two alternatives for thermal integration in a rSOC system. Both alternatives use pure hydrogen as fuel and air as an oxidant and implement a high temperature thermal storage as well as a linear Fresnel collector system.

2. Aim of the work

A reversible Solid Oxide Cell (rSOC) based on hydrogen/steam and air is coupled with two plant

configurations that use different thermal integration strategies. Both layouts implement a linear Fresnel solar concentrator to evaporate water during electrolysis operations. In the first configuration a high temperature thermal energy storage (TES) is connected to the stack via an isothermal heat pipe and an intermediate temperature TES using diathermal oil as heat transfer fluid is installed to preheat the steam flow. In the second configuration the air flow operates the thermal management of the adiabatic stack, a molten salt based high temperature TES transfers heat to and from the plant using reversible counter current heat exchangers. The objectives of this thesis are: the creation of two plant layouts that are able to correctly operate in both FC mode and EC mode at a nominal condition; calculate the heat exchanger areas involved and obtain an indication of feasibility; collect data on hydrogen and energy (electric and thermal) consumption and production; calculate performance indexes such as plant electric efficiency and plant first principle efficiency; perform a sensitivity analysis at part load; understand how solar irradiance influences plant operations and performances and ultimately compare the two configurations.

3. Methodology

3.1. rSOC model

To perform an analysis of the two plants a valid rSOC model has to be adopted. The rSOC model chosen for this work is a zero-dimensional cell representation that can describe electrochemical and thermal phenomena in both operating modalities. A state-of-the-art planar rSOC with Yttria Stabilized Zirconia electrolyte, an anode channel made by YSZ with Nickel and a cathode channel mainly composed by perovskite (LSM) mixed with YSZ to increase the triple phase boundary (anode and cathode are referred to fuel cell operation in this case). The model has been validated by the experimental work of Menon et al. and Kazempoor et al., respectively considering a stack temperature of 750°C and 850°C.

3.2. Performance indexes

The most important indexes needed to acquire a good understanding of the cell stack and the plant performances are reported below. The electric efficiency of the cell stack is defined as:

$$\eta_{STACK,FC} = \frac{P_{STACK,FC}}{LHV_{H_2} \cdot \dot{m}_{H_2,cons}} \quad (1)$$

$P_{STACK,FC}$ is the electrical power generated by the stack in fuel cell mode, LHV_{H_2} is the lower heating value of hydrogen gas (120 kJ/kg) while $\dot{m}_{H_2,cons}$ is the hydrogen mass flow rate consumed inside the cell stack. The plant electric efficiency in fuel cell mode is defined as:

$$\eta_{PLANT,FC} = \frac{P_{NET,FC}}{LHV_{H_2} \cdot \dot{m}_{H_2,cons}} \quad (2)$$

Where $P_{NET,FC}$ is the net electrical power produced by the plant. The first principle plant efficiency is equal to:

$$\eta_{I,FC} = \frac{P_{NET,FC} + Q_{TES}}{LHV_{H_2} \cdot \dot{m}_{H_2,cons}} \quad (3)$$

Q_{TES} is the thermal power drawn from the plant and stored inside the thermal energy storage for further reuse.

The cell efficiency in EC mode is defined as the ratio between the lower heating value of hydrogen multiplied by the production rate and the electrical power consumed by the stack ($P_{STACK,EC}$).

$$\eta_{STACK,EC} = \frac{LHV_{H_2} \cdot \dot{m}_{H_2,prod}}{P_{STACK,EC}} \quad (4)$$

The plant electric efficiency in EC mode is defined as:

$$\eta_{PLANT,EC} = \frac{LHV_{H_2} \cdot \dot{m}_{H_2,prod}}{P_{TOT,IN}} \quad (5)$$

$P_{TOT,IN}$ is the sum of all electric power consumptions inside the plant.

The first principle plant efficiency ($\eta_{I,EC}$) is defined including all electric and thermal inputs to the system ($P_{TOT,IN}$ and Q_{IN}).

$$\eta_{I,EC} = \frac{LHV_{H_2} \cdot \dot{m}_{H_2,prod}}{P_{TOT,IN} + Q_{IN}} \quad (6)$$

A first principle efficiency excluding the linear Fresnel contribution ($\eta_{I,EC,NOLF}$) is also considered.

3.3. Heat pipe-based configuration (A)

The first option for thermal integration is a heat recovery system directly connected to the stack using heat pipes. This solution includes the Solid

Oxide Cell stack, six heat exchangers, a linear Fresnel solar collector, a condenser, a blower, a compressor, a thermal energy storage (TES) system based on thermal oil and a molten salt-based TES connected to the stack via an isothermal heat pipe system. The plant layout is reported in *Figure 1*.

3.3.1. EC mode

A water flow enters the plant and is sent to an economizer (HX-WATPR) where it is heated with the air flow exiting the plant. Water is further preheated by entering HX-FUELPR with a hot inlet flow composed by a mixture of hydrogen and steam. The linear Fresnel collector evaporates the water. Inside HX-OIL, the steam undergoes a further temperature increase by using hot thermal oil. The last superheating takes advantage of the hydrogen/steam flow exiting the cell. COCUR reduces the temperature difference between the two flows before entering the stack. Inside the cell, on the cathodic side, part of the water steam is

HX-AIRPR, the air stream undergoes a further cooling in the economizer.

3.3.2. FC mode

A hydrogen flow enters the plant, unreacted H_2 from the condenser is added to the main flow. The fuel stream enters HX-FUELPR, where the thermal content carried by the stream exiting the cell stack is exploited. Additional heat, at higher temperatures, is transferred to the cold flow inside HX-SUPHEAT, but in this case part of the hot flow coming from the cell is bypassed. After reducing the temperature difference inside a co-current heat exchanger between the two flows entering the cell, the heated fuel enters the anodic side of the stack and is oxidized. The reactions occurring in the cell are exothermic, a heat pipe system transfers heat from the cell stack to the high temperature thermal energy storage, electric power is generated. The steam-hydrogen flow exiting the cell is split in two streams, part of the flow goes inside HX-

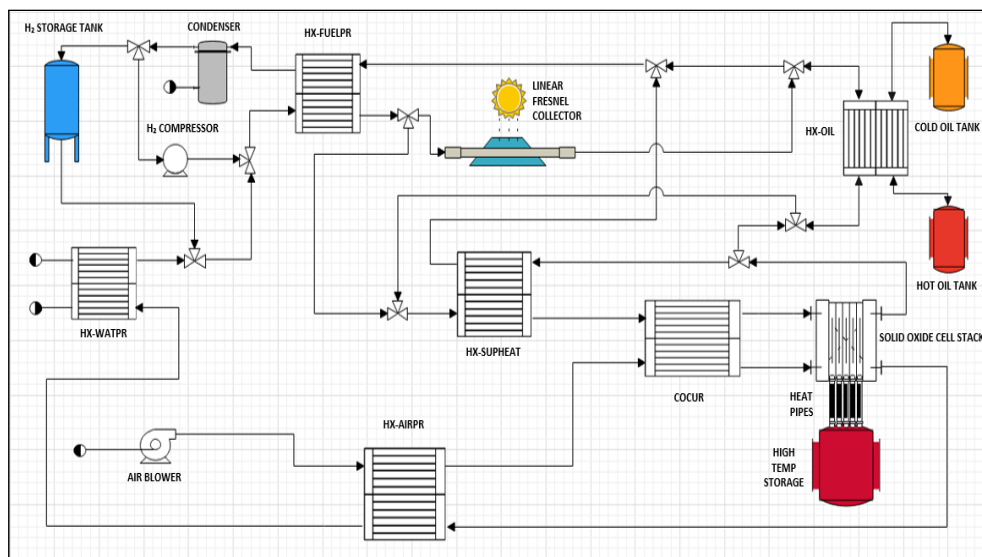


Figure 1: Plant layout, configuration A

reduced to hydrogen. A heat pipe system transfers heat from the high temperature salt storage to the cell stack and electric power is required for the chemical reaction to occur. The high temperature flow progressively reduces its temperature before entering a condenser where hydrogen gas is separated.

In the air circuit, the slightly pressurized air enters HX-AIRPR and reaches the required temperature for electrolysis operations in the cell, the heat required is harvested from the air flow leaving the cell stack. The oxygen produced inside the stack is swept by the air flow. Before leaving the plant, after the temperature reduction inside

SUPHEAT. The second stream is directed inside HX-OIL, a countercurrent heat exchanger that uses thermal oil as energy storage mean, for cooling the steam. The heated oil is stored in the hot tank. The exiting stream flow is further cooled down in the economizer before entering the condenser. In the condenser, before leaving the plant, the steam flow is brought into its liquid form and separated from the incondensable fraction. The unreacted hydrogen is compressed and reused at the plant inlet.

Inside the air circuit, a flow of air enters an air-to-air heat exchanger (HX-AIRPR), by exploiting the thermal content of the high temperature air

flow exiting the cell, the temperature of the oxidant flow is increased. Part of the oxygen inside the cell is consumed, the air flow exiting the cell has a smaller flow rate and a low oxygen content, after cooling down inside the air-to-air heat exchanger, the flow exits the plant.

3.3.3. Plant design specifications

The plant layout has initially been studied analyzing electrolysis and fuel cell mode separately.

In fuel cell mode the temperature of the stack outlet flows is set to 850°C while a temperature of 750°C is chosen during electrolysis operation. The temperature difference gives the possibility to operate the cell during EC mode at a low current density, the cell stack has an endothermic behavior, resulting in a higher efficiency. The thermal input needed comes from a high temperature storage, that can be recharged during fuel cell operation. The heat transfer between the rSOC stack and the molten salts vessel is made possible using a heat pipe system, A suitable storage media is identified in a 68-32% molar ratio KCl-MgCl₂ eutectic molten salt, capable of operating at temperatures above 800°C. The intermediate temperature TES in configuration A uses as heat transfer fluid (HTF) a thermal oil (biphenyl (C₁₂H₁₀) and diphenyl oxide (C₁₂H₁₀O)). The fluid operating temperature range in this application is 230-400°C.

The nominal current density in EC mode is set to 0,513 A/cm² well below the thermoneutral point (0,69 A/cm²). In FC mode the current density is set to 0,230 A/cm². As a consequence, the ratio between the electrical power output in FC mode and the electric power input in EC mode is one third.

The cell works in EC mode with a 65% water utilization factor (UF), the air flow rate is fixed to have an oxygen content at the plant outlet that is twice the one entering the cell. A 10% hydrogen molar fraction is present in the vapor stream. In fuel cell mode the UF of H₂ is 85%, the air entering the blower has a fixed flow rate so as to obtain an oxygen UF across the cell stack of 40%. The linear Fresnel concentration system is dimensioned to evaporate the water and contribute to the superheating up to 220°C. To perform an extensive analysis on the heat exchanger area a plant size has to be assigned, in nominal conditions the plant is set to have an electric power input of 3 MW at the stack in SOEC mode, while a 1 MW electric power

output at the stack in fuel cell mode is obtained. The electrochemical system is made of roughly 12000 planar cells considering that each one has an active area of 400 cm².

The optimal heat exchanger dimensions are calculated in nominal conditions separately in fuel cell and electrolysis operations by setting the approach temperature in each heat exchanger. For example, during EC operations:

- in the economizer (HX-FUELPR) the approach temperature between the cold inlet and the hot outlet is set to 10°C; the hot fluid thermal capacity is significantly lower than the one of liquid water;
- the thermal oil heat exchanger (HX-OIL) increases the steam temperature from 220°C up to 360°C (approach temperature is 40 °C);

For surface areas computation the assumption of a reasonable overall heat transfer coefficients (U) is necessary. Typical values of U are selected considering the different characteristics of the fluids. The plant is slightly pressurized in both operational modes and the cell stack works at ambient pressure.

The system described above is implemented in the suite Aspen Plus. The Design Spec tool is employed to comply with the described design parameters. It allows to adjust an input parameter within the plant, such as a flow rate, a stream temperature or composition, in order to impose a value to a dependent output taken as objective. Hydrogen, water, and air flow at the input are regulated to match the previously described specifications. Stack thermal duty and condenser refrigeration duty are also modified to satisfy the components exit temperature requirement.

3.4. Air cooled configuration (B)

The second option selected for thermal integration uses the air flow to recover heat from the adiabatic stack. This solution includes the Solid Oxide Cell stack, six heat exchangers, a linear Fresnel solar collector, a condenser, a blower, a compressor, and a molten salt-based TES. The layout is depicted in *Figure 2*.

3.4.1. EC mode

A water flow enters the plant and is mixed with a small amount of hydrogen that is recirculated from the condenser outlet. The mixed flow is directed inside an economizer (HX-WATPR) where the remaining thermal content of the air that is exiting the plant is exploited. The water flow

enters the linear Fresnel system where it is fully evaporated. The steam enters a countercurrent heat exchanger (HX-FUELPR) to further increase its temperature using the heat coming from the humid hydrogen directed to the condenser. In order to achieve a higher steam temperature, the hot molten salts accumulated during fuel cell operations are used inside a heat exchanger (HX-S-

mass flow rate and oxygen content. Before leaving the plant, after cooling down inside HX-AIRPR, the air stream undergoes a further temperature reduction inside the economizer.

3.4.2. FC mode

A hydrogen flow enters the plant, unreacted hydrogen coming from the condenser is added to

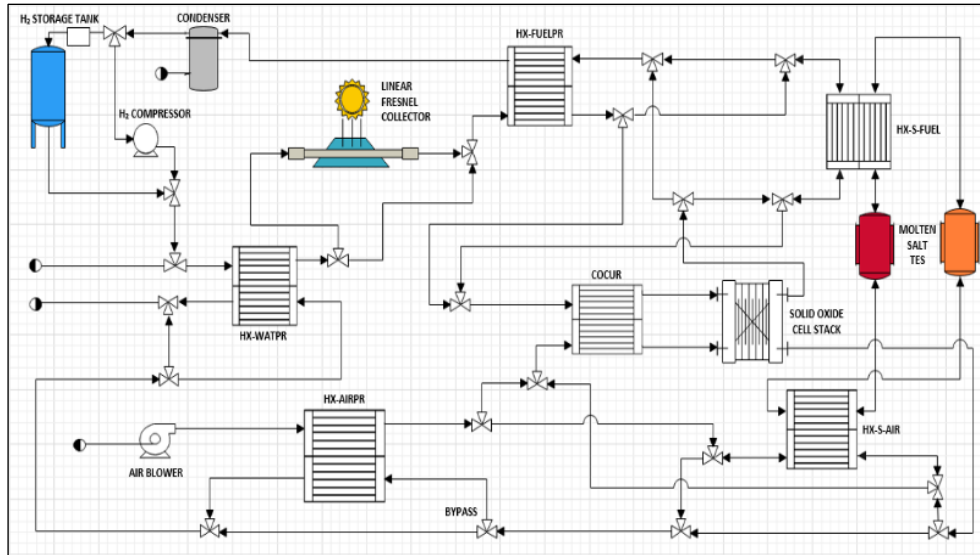


Figure 2: Plant layout, configuration B

FUEL). The hot steam and the sweep air flow are sent in a co-current heat exchanger (COCUR) to reduce their temperature difference. Inside the cell, on the cathodic side, part of the water steam is reduced and hydrogen is formed. The cell stack can be considered adiabatic, no external thermal input is applied at the stack. Electric power is required for the chemical reaction to occur. The high temperature humid hydrogen flow exits the cell and enters HX-FUELPR, the flow progressively reduces its temperature before it is sent to a condenser where its water component returns to the liquid form allowing to separate the high purity hydrogen fraction. Liquid water is reused in the water flow fed at the plant inlet.

Inside the air circuit, the slightly pressurized flow is heated inside an air-to-air countercurrent heat exchanger (HX-AIRPR), exploiting the thermal content of the high temperature air flow exiting the cell. In order to reach the minimum required temperature for electrolysis operations inside the cell, the air flow inside HX-S-AIR absorbs a high temperature thermal flow coming from the hot molten salts accumulated during FC mode. Inside the cell stack (anodic side), the produced oxygen is swept by the air flow. The air flow exiting the cell experiences an increase in

the main flow. The fuel stream is sent in a countercurrent heat exchanger where the thermal content carried by the water steam exiting the cell stack is exploited. After reducing the temperature difference with the air flow, the heated fuel enters the anodic side of the stack and is oxidized. The reactions occurring in the cell are exothermic, therefore the flows passing through the stack experience a temperature increase, the intensity of such gradient depends on the current density. The cell stack is considered adiabatic, meaning that there is no external removal nor addition of heat, electric power is generated. The steam-hydrogen flow exiting the cell is cooled down inside HX-S-FUEL, a countercurrent heat exchanger that takes the molten chloride salts at the cold intake. The salt flow is accumulated in the hot storage. The cooled steam is sent to the condenser where unreacted hydrogen is compressed and reused.

In the other circuit the slightly pressurized air enters an air-to-air countercurrent heat exchanger (HX-AIRPR), its temperature is increased by exploiting the thermal content of the air flow exiting the stack. Increasing the air temperature before entering the stack guarantees that the maximum temperature difference across the stack in fuel cell operations is respected. Part of the

oxygen in the air flow is consumed. The oxygen depleted air flow is cooled down inside HX-S-AIR with the use of molten salts previously accumulated in the cold storage during electrolysis operation, the hot molten salt flow is stored. After a further cool down it is discharged into the outside environment.

3.4.3. Plant design specifications

The electrochemical system is identical to the one described in configuration A (3.3.3), both in terms of number of cells and arrangement. The cell design specifications concerning electric power and outlet temperatures are also the same.

Part of the plant thermal input during EC mode comes from a high temperature TES that can be recharged during fuel cell operation. A suitable HTF for this application is found in a 68-32% molar ratio KCl-MgCl₂ molten salt (also used in configuration A). The cold salt storage is set to 720°C while a temperature of 820°C is selected in the hot molten salt storage.

The cell is designed to work in electrolysis mode with a 65% water UF, the air flow rate is set to a value that guarantees the adiabaticity of the cell stack. A 10% hydrogen molar fraction is present in the vapor stream entering the stack. In fuel cell mode the UF of H₂ is 85%, the air entering in the plant has a flow rate that can be regulated in order to obtain 850°C at the cell stack outlet. The linear Fresnel concentration system is dimensioned to evaporate the water and bring the steam to a temperature of 120°C. A critical aspect to keep in mind is that in EC mode the cell has an endothermic behavior and therefore the inlet streams must have a temperature well above the outlet one fixed to 750 °C. Reaching a higher temperature is only possible with the use of hot chloride molten salts, stored during fuel cell operation at 820°C.

Typical values of U are selected considering the different characteristics of the fluids. The optimal heat exchanger dimensions are found in nominal conditions separately in fuel cell and electrolysis operations by setting conditions on heat exchangers approach temperatures. For example:

- during FC operations, in the cold molten salts/steam heat exchanger (HX-S-FUEL), the hot flow outlet temperature is set to 740°C, or equivalently an approach temperature at the cold side inlet of 20°C. A smaller temperature difference is possible, the choice is a compromise between the

inevitable increase of heat exchanger area and having a higher hot molten salt flow rate;

- considering the air circuit during EC mode, at HX-S-AIR cold outlet the temperature of the sweep flow reaches 805°C;

The described system is implemented in the suite Aspen Plus. The Design Spec tool is employed to comply with the described design parameters. In this case refrigeration duty hydrogen, water, and air flow at plant inlet, as well as heat exchanger surface areas are adjusted to match the former specifications. The water/hydrogen conversion in the cell is computed through a Calculator tool that implements the rSOC model described in Section 3.1.

4. Results

4.1. Configuration A

The results of the required heat exchanger surface area reported in *Table 1* are obtained independently in both EC and FC mode. Two possible values for each of the five heat exchangers installed in configuration A (HX-WATPR is only used in EC mode) are obtained. The largest size between the two is chosen to fulfill the heat exchanger requirements in both conditions. On the opposite functioning mode, the system operates with an over dimensioned heat exchanger.

Table 1: U coefficients and calculated heat exchangers area (Conf. A)

| | FC MODE | | EC MODE | |
|------------|------------------------|----------------|------------------------|----------------|
| | U [$\frac{W}{m^2K}$] | AREA [m^2] | U [$\frac{W}{m^2K}$] | AREA [m^2] |
| HX-FUELPR | 130 | 6,93 | 600 | 4,12 |
| HX-OIL | 400 | 3,50 | 400 | 12,55 |
| COCUR | 30 | 0,02 | 130 | 5,23 |
| HX-SUPHEAT | 130 | 5,98 | 250 | 23,15 |
| HX-AIRPR | 30 | 512,2 | 30 | 324,5 |
| HX-WATPR | | | 40 | 37,03 |

Stream property results in nominal conditions are obtained with the implementation of the plant layout and specifications in AspenPlus considering previously calculated heat exchangers areas instead of fixing approach temperatures.

In fuel cell operation, 1 MW of electric power is produced at the stack, roughly 41 kilograms of hydrogen gas are oxidized each hour, the air blower absorbs 23 kWe. The heat pipe system harvests 300 kW of high temperature heat and stores it inside the TES. Thermal oil flow rate accumulated inside the intermediate temperature

TES is approximately 425 kg/h harvesting 47 kW of thermal power from the hot steam.

In electrolysis operation 3 MW of electric power are consumed at the cell stack, approximately 92 kilograms of hydrogen gas are produced each hour, the air blower absorbs 21 kW. The linear Fresnel collector system exchanges 734 kW of thermal power to the evaporating water inside the plant. Thermal oil flow rate utilized to heat up the water vapor is around 976 kg/h transferring 108 kW of heat to the steam. The heat pipe system transfers 240 kW of high temperature heat from the TES to the cell stack. Performance indexes results defined in section 3.2 are reported in *Table 2*.

Table 2: Performance indexes (Conf. A)

| $\eta_{STACK,FC}$ | $\eta_{PLANT,FC}$ | η_{LFC} | $\eta_{STACK,EC}$ | $\eta_{PLANT,EC}$ | η_{LEC} | $\eta_{LEC,NOLF}$ |
|-------------------|-------------------|--------------|-------------------|-------------------|--------------|-------------------|
| 72,94% | 71,26% | 96,96% | 102,21% | 101,51% | 74,74% | 91,04% |

4.1.1. Part load

A sensitivity analysis at partial load conditions is performed. In fuel cell mode, the oxidized hydrogen flow is lowered and consequently the

In electrolysis mode, at partial load conditions, the current density is reduced, causing a fall in the power input required at the cell stack and in the hydrogen production. As power input decreases, the water vapor UF inside the cell remains equal to the value assigned under nominal condition. To maintain a steam utilization factor constant the liquid water entering the plant is reduced and the linear Fresnel collector becomes oversized. A current density reduction causes an increase in cell efficiency and the overall plant efficiency is also improved. The hot thermal oil required to heat up the steam flow decreases with power input and so does the heat exchanged (*Figure 3*). A higher steam temperature can be reached at HX-OIL outlet due to the presence of over dimensioned heat exchangers at part load conditions. The thermal power required at the cell stack has a maximum between 60% and 70% of the cell nominal electric power, at a lower current density, Joule losses are reduced and the endothermicity of the reaction becomes more relevant. In other words, the ratio between the plant electric power consumption and

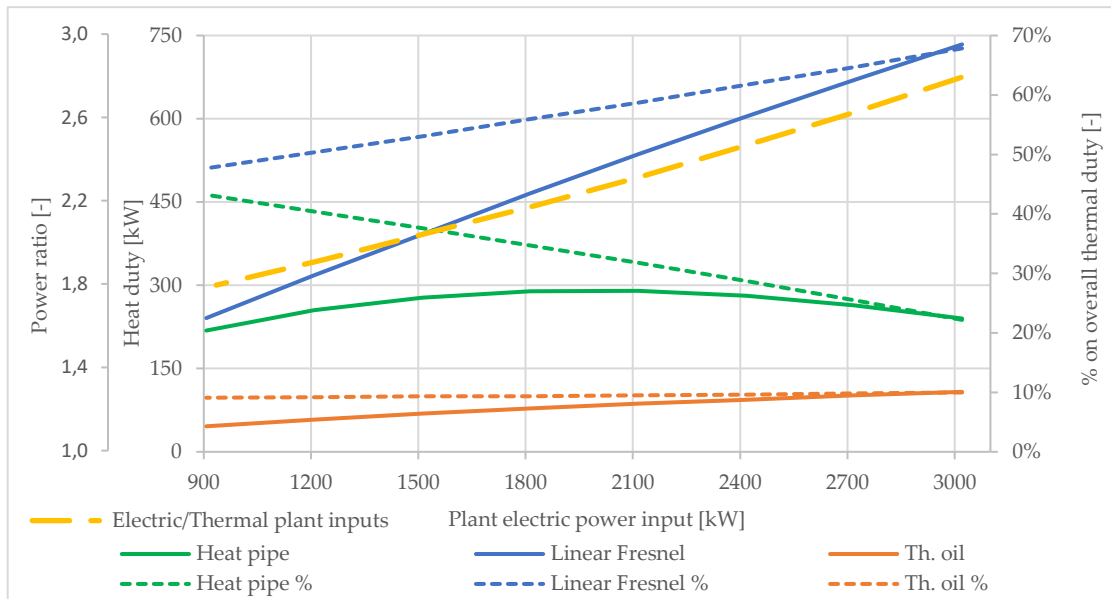


Figure 3: Plant thermal contributions and ratio between electric and thermal power inputs as a function of the plant electric consumption (EC MODE)

stack power output and the current density are reduced. A similar trend is experienced by the oxidizing agent flow entering the plant, in fuel cell mode as power output is decreased hydrogen and oxygen UF are set constant to the assigned nominal value. The reduction of the thermal power collected in the high temperature TES is both due to the reduction of the amount of oxidized hydrogen and the increase of cell efficiency.

the plant thermal request decreases, the trend is shown by the dotted yellow line.

4.1.2. Solar irradiance effect

The system is able to operate also in case of partial or no solar irradiation, its nominal value is set to 1000 W/m². In this configuration, as solar radiation decreases the percentage contribution to the overall heat requested by the plant in electrolysis operation is provided by the

intermediate temperature TES. At nominal load, the heat exchanged inside HX-OIL ranges from 100 to over 800 kW depending on solar thermal power availability. In Figure 4 is reported the total thermal power necessary to operate in EC mode, divided

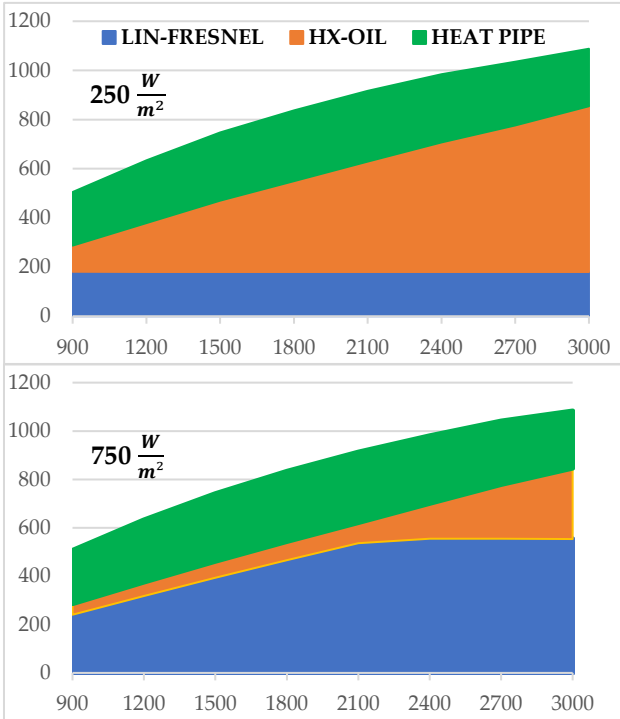


Figure 4: Thermal contributions [kW] as a function of stack electric power input [kW] for two levels of solar irradiance (EC MODE)

between the three contributions, for different plant loads and two levels of solar irradiance.

4.2. Configuration B

The results of the required heat exchanger surface area are reported in Table 3, the same approach used in configuration A is used.

Table 3: U coefficients and calculated heat exchangers area (Conf. B)

| | FC MODE | | EC MODE | |
|-----------|------------------------|----------------|------------------------|----------------|
| | U [$\frac{W}{m^2K}$] | AREA [m^2] | U [$\frac{W}{m^2K}$] | AREA [m^2] |
| HX-FUELPR | 130 | 16,73 | 250 | 32, 25 |
| HX-S-FUEL | 600 | 2,03 | 600 | 7,68 |
| COCUR | 30 | 11,70 | 130 | 0,00 |
| HX-AIRPR | 30 | 1289,8 | 30 | 558,9 |
| HX-S-AIR | 60 | 175,80 | 60 | 214,54 |
| HX-WATPR | | | 40 | 156,11 |

The fact that HX-AIRPR area is twice as large as it is required in EC mode, causes the sweep flow at HX-AIRPR exit to have a higher temperature than the cold salt storage. For this reason, approximately 15% of the flow bypasses HX-AIRPR and is sent directly downstream where it is mixed back with the main flow. Stream property

results are obtained with the same procedure used in configuration A.

Hydrogen production and consumption rate as well as electric power values at the stack are the same one found for configuration A, the two electrochemical devices operate identically.

In fuel cell mode the air blower electric consumption is 48 kW. The heat exchanged inside HX-S-FUEL between the superheated steam and the molten salts is 35 kW, an order of magnitude higher is the thermal power exchanged between the oxygen depleted flow exiting the cell and the chloride molten salts (266 kW). The heat transfer fluid mass flow rates stored in the TES are 1576 kg/h coming from the anodic side and 11900 kg/h from the cathodic side. A higher amount of molten salt flow passes inside HX-S-AIR compared to HX-S-FUEL, in this configuration, the air flow regulation is used to correctly extract heat from the cell stack and let out streams at 850°C.

In electrolysis operation the air blower absorbs 41 kW of electric energy. The linear Fresnel collector system exchanges 736 kW of thermal power to the evaporating water inside the plant. The chloride molten salts exchange 137 kW inside HX-S-FUEL superheating the water vapor, the mass flow rate required to transfer it is around 6120 kg/h. The hot molten salts flow rate, needed to bring the sweep air flow to a temperature of 804°C inside HX-S-AIR, is around 9650 kg each hour. Efficiencies results are reported in Table 4.

Table 4: Efficiencies results (Conf. B)

| $\eta_{STACK,FC}$ | $\eta_{PLANT,FC}$ | $\eta_{I,FC}$ | $\eta_{STACK,EC}$ | $\eta_{PLANT,EC}$ | η_{LEC} | $\eta_{LEC,NOLF}$ |
|-------------------|-------------------|---------------|-------------------|-------------------|--------------|-------------------|
| 72,94% | 69,43% | 91,41% | 102,21% | 100,83% | 74,29% | 90,25% |

4.2.1. Part load

In fuel cell mode at part load conditions, the oxidized hydrogen flow is lowered and consequently are the power output and the current density. The air flow that in this configuration is regulated to guarantee 850°C at the cell stack outlet, decreases for two reasons: firstly because of the reduction of the hydrogen oxidized at the anodic side and secondly for the reduction of irreversibilities inside the cell. On the other hand, on the anodic side, the hydrogen utilization factor is kept constant.

During EC operation, the most critical part inside the plant, is found in the sweep air circuit. As current density is lowered, an increment in cell efficiency is present, causing an increase in the heat provided by the air flow inside the stack. The

maximum heat, provided by the molten salts, is found between 60% and 70% of the plant nominal electric power consumption. The ratio between the plant electric consumption and the plant heat request increases with the load. The plant response to this additional heat request at partial load is an increase in the air flow entering the plant. The linear Fresnel contribution to the overall thermal input decreases (*Figure 5*).

In order to limit the sweep air mass flow rate increase, an electric heater downstream of HX-S-AIR is installed. Its role at part load conditions is to increase the sweep air flow temperature to 810°C: the electric consumption is minimal and the benefit of limiting the sweep flow is worth the extra

considered constant when a variation in the solar irradiance level is introduced, this is due to the fact that the sweep air flow rate, as a first approximation, does not change when different levels of thermal power from the sun are considered. This makes the electric heater consumption also independent from what occurs inside the linear Fresnel. When solar irradiance is lower than the nominal one, the plant request to the high temperature TES increases and the hot molten salt additional flow is directed inside HX-S-FUEL.

As solar thermal power availability increases, the plant utilizes a lower amount of molten salts causing an increase of several percentage points in

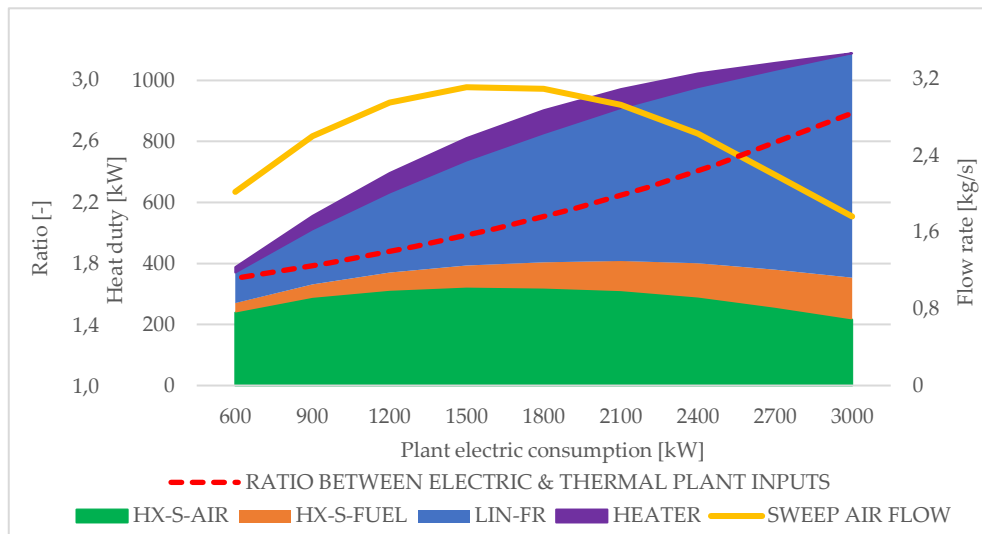


Figure 5: Thermal power contributions, sweep flow rate and ratio between plant electric over thermal power consumption as a function of plant electric consumption. (EC MODE)

electric energy consumption.

The plant efficiency increases under part load conditions, however the improvement inside the plant electric power range between 50% and nominal condition is not significant. Once the point of maximum air flow rate, or analogously the limit of maximum heat requested by the stack, is reached the plant efficiency increases at a higher pace.

4.2.2. Solar irradiance effect

Electrolysis operation for different levels of solar irradiance intensity is investigated. As the available thermal power provided by the sun and exchanged inside the linear Fresnel collector system decreases, depending on the plant consumption, part of the evaporation might occur inside HX-FUELPR and HX-S-FUEL. The thermal power exchanged in HX-S-AIR and therefore the hot molten salt flow used in the air circuit can be

$\eta_{LEC,NOLF}$. This performance index can be useful when heat provided by the solar evaporator is considered a free source of thermal power.

5. Conclusions

This work focuses on the investigation of different thermal integration options in a reversible Solid Oxide Cell system.

By setting a current density in nominal condition during electrolysis operation well below the thermoneutral point the cell stack has a strongly endothermic behavior. On the other hand, choosing an electric power ratio at the stack between FC mode and EC mode of one third reduces the amount of thermal energy accumulated inside the TES per unit of time and per unit of hydrogen consumed. In both configurations, the fuel and the air sides require

several heat exchangers. They are sized analyzing the different operating modes and selecting the most critical operating condition for each in particular the air-to-air preheaters in both configurations are sized in fuel cell mode despite the low nominal stack electric output. In both layouts the linear Fresnel plays a crucial role, more than 67% of the overall thermal inputs during EC operation is provided by it (high radiation).

The use of a heat pipe system in configuration A allows to keep the air flow rate under control reducing the electric blower size and consumption in both EC and FC mode. In this layout, despite operating with an endothermic behavior of the SOEC, the steam and air flows entering the cell stack can have a temperature below the one set at the stack outlet, in configuration B on the other hand, both flows have to be well over 750°C.

In configuration B the streams entering the SOEC can be only heated up to the required condition with the use of a high temperature TES. To achieve a temperature high enough, during FC mode the cold molten salts must exploit the thermal content of the stack outlets right after the exit. An important result is also that most of the thermal energy collected by the molten salts is exchanged inside the air cooler (HX-S-AIR), the role of the steam cooler (HX-S-FUEL) in fuel cell operation is surprisingly marginal. During fuel cell mode the thermal power absorbed, and the heat stored inside the two TES by configuration A is 16,6% higher, configuration A also requires slightly less thermal power from its thermal energy storages compared to configuration B.

Having a lower air flow rate entering the plant, like in configuration A (roughly half of configuration B air flow both in EC and FC mode), causes in absolute terms a higher difference in mass flow rate between the two streams across the stack in both modes of operation. This high difference makes the presence of a TES on the air circuit pretty much unnecessary.

In both configurations the plant efficiencies during electrolysis operation are over 100% (101,5% case A, 100,8% case B) and around 70% in fuel cell mode (71,3% case A, 69,5% case B).

At part load conditions both layouts experience a plant efficiency increase in EC mode, but this improvement is sharper in configuration A compared to B. At part load, during electrolysis operations, the contribution of the heat coming from the linear Fresnel collector to the overall plant

thermal requirements reduces more rapidly in configuration B (from 67,5% at nominal load to 32,5%) compared to configuration A. As load is reduced in EC mode the sweep air flow increases in configuration B and linearly decreases in configuration A, influencing the blower electric consumption. At a stack power input of 900 W (30% of the nominal value) the sweep air flow in configuration A is one ninth of the air flow in configuration B. A solution proposed in this thesis is the installation of a heater in configuration B. This component operates at part load, allows for the size of HX-S-AIR to be set in nominal conditions and brings down the value of the maximum sweep air flow rate.

Considering both configurations at nominal conditions, if the plant balance between produced and consumed hydrogen over a period of time is null, the plant can be considered a net thermal energy producer (about 4,8 kWh per kg of hydrogen consumed in case A, and 3,5 kWh in case B). When a thermal energy balance over a time span is imposed instead, the plant becomes a net hydrogen producer.

Given the high plant thermal request during electrolysis operation in both configurations, even at nominal conditions, the TES management over time is highly dependent on the water evaporated inside the linear Fresnel collector system. Even assuming thermal losses equal to zero in the plant and inside the TES, an average level of solar irradiance not lower than 500 W/m² must be guaranteed during electrolysis operation, otherwise either an external source of heat or hydrogen has to be provided to the plant.

In conclusion, this work proves that a rSOC plant with integrated heat storage is a valid electricity storage system alternative. Both configurations have a convincing thermal integration system that guarantees competitive plant electric efficiencies and a thermal management of the plant in both operations. These are important results especially because this work includes one of the most detailed operating maps for this kind of systems in the investigated literature. Configuration A has overall better performances, especially at part load. The heat pipe system is effective in lowering the air request, but it adds complexity from a stack design point of view. Configuration B on the other hand has only one TES but significantly larger heat exchangers installed on the air circuit.