Economic analysis of CO_2 capture from natural gas combined cycles using Molten Carbonate Fuel Cells

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Nomenclature

BUA	bottom up approach
C	cost (€)
CCA	cost of CO_2 avoided (Eq. (1)) (€/kg_{CO2})
CCR	carbon capture ratio (%)
CCS	carbon capture and storage
COE	cost of electricity (€/MWh)
E	CO_2 specific emission rate (kg_{CO2}/kWh_{el})
EXP	expansion
HR	heat rate (kJ_{LHV}/kWh_{el})
HRSG	heat recovery steam generator
LHV	fuel lower heating value
LP	low pressure
MCFC	Molten Carbonate Fuel Cell
MEA	mono-ethanol ammine based CO_2 capture system
NG	natural gas
NGCC	natural gas combined cycle
SPECCA	specific primary energy consumption for CO_2 avoided (Eq. (3)) (MJ_{LHV}/kg_{CO2})
ST	steam turbine
TEC	Total Equipment Costs (€)
TIT	turbine inlet temperature (°C)
TPC	total plant cost (€)
TPI	total plant investment (€)
U_{CO2}	CO_2 utilization factor (%)
WGS	water gas shift
η	efficiency (%)

Subscripts

el	electrical
ref	reference

1. Introduction

During recent years, environmental issues associated with the use of fossil energy sources have been addressed, especially regarding carbon dioxide and its role in global warming, driving the search for electricity generation methods capable of limiting CO_2 emissions into the atmosphere. Besides renewable energy, one way is to capture the CO_2 produced by fossil fuels combustion and to store it in deep geological formations [1–3].

Low carbon energy conversion schemes for power production are generally classified in: (i) post-combustion capture, (ii) oxy-combustion capture and (iii) pre-combustion capture.

In the first case, CO_2 can be separated after the combustion using amine scrubbing [4,5], which is the state-of-the-art, or chilled-ammonia processes [6]. Oxy-combustion separation is based on an almost-stoichiometric O_2 combustion: main reaction products are therefore a mixture of CO_2 and steam which can be easily separated after condensation, even if CO_2 purity constraints may require additional stream processing to get rid of excess O_2 and other incondensable gases. The last category is pre-combustion decarbonisation which consists in transferring the heating value from primary fuel to a syngas and eventually to hydrogen, later used as a fuel in a combined cycle, without any CO_2 emission.

Besides, an additional option is offered by the use of fuel cells. Among different possible solutions using either Solid Oxide Fuel Cells (SOFC) or Molten Carbonate Fuel Cells (MCFC), the first ones are mainly investigated for their capability to permeate oxygen ions providing the possibility to arrange a sort of oxy-combustion capture plant [7,8]. The second ones have the potential to act as "active CO_2 concentrators" downstream a conventional power plant, behaving accordingly to a post-combustion approach [9,10].

Relevant studies exist in technical literature evaluating the economic outlook of SOFC application in coal fired power plants with CO_2 capture. Among the others, a comprehensive analysis has been recently released by the US DOE [11] with reference to Integrated Gasification Fuel Cell (IGFC) cycles. On the opposite, relatively few works were related to natural gas fired configurations and, to the authors' knowledge, no work has been done about the economic perspectives of applying MCFCs to CCS.

This paper stems from a previous work [12] that investigated the optimal integration of MCFCs within a natural gas combined cycle by a thermodynamic point of view and pointed out the remarkable advantages of the resulting plant in terms of energy efficiency and, moreover, very low penalty for CO_2 capture with respect to competitive technologies. Such advantages motivated to develop an analysis extension to thoroughly investigate the economic viability of this technology, comparing it with commercially ready options in terms of cost of electricity and cost of CO_2 avoided, and aiming to identify the fuel cell cost target – in terms of specific investment (€/kW_{el}) – that would make the NGCC + MCFC solution economically competitive.

2. MCFC technology and economic outlook

MCFCs are a well-known candidate for clean power generation from a variety of fossil fuels, including natural gas and biogas [13,14]. They are primarily developed for distributed power generation: the leading MCFC manufacturer (Fuel Cell Energy, USA) markets MW class fuel cell plants as on-site power sources for corporate buildings, hospitals, schools or factories [14]. The largest single unit is a 2.8 MW, rated at 47% net LHV efficiency from natural gas, while the largest installation is a 59 MW recently completed in South Korea, where also a 11.2 MW plant (made by 4×2.8 MW units) was previously installed [15,16]. As a counterpart to promising installation results, few other relevant MCFC manufacturers are still in operation (IHI in Japan and KEPRI and Doosan in Korea, in addition to POSCO Power, and FCE Solutions

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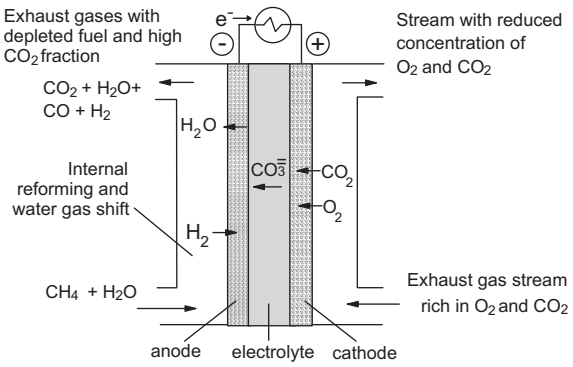

Fig. 1. Operating principles of a MCFC.

in Germany [14]), indicating the difficulties in MCFC technology development.

The operating principle of a MCFC (Fig. 1) involves that oxygen is taken from ambient air and is transported to the cell anode combined with CO_2 by carbonate ions ($CO_3^=$).

At the anode oxygen is released and oxidizes the fuel, primarily combining with hydrogen to produce steam. Hydrogen can be generated by an internal reforming process, either taking place at the anode or in a thermally integrated reactor, in both cases exploiting heat released by the fuel cell to sustain the reforming endothermic reactions. The carbon monoxide generated by reforming is converted to CO_2 by the concurrent water gas shift reaction which produces additional hydrogen, again oxidized in the process. In stand-alone applications, usually a fraction of the anode effluent stream is burned in a catalytic combustor and recycled at the cathode inlet, in order to sustain the formation of carbonate ions; while in the case investigated in this work, CO_2 is provided by the exhaust stream of a gas turbine.

As a result, the MCFC converts natural gas into electricity and simultaneously moves CO_2 from the cathode to the anode side, simplifying its separation. It is therefore possible to reduce the energy demand for CO_2 capture with a superior efficiency compared to conventional competitive CCS techniques.

In the last years, the technological evolution of MCFCs has always been targeting the issues of cost reduction and life extension increase. By the point of view of cost reduction, significant efforts have been made to lead to commercial and industrialized standards. The leading manufacturer has seen a decreasing cost history in parallel to the growth of production (above 56 MW in 2011 [15,16]), with a reported cost which is about $3 million for a MW of capacity (3000 $/kW) excluding installation costs (another 800–1200 $/kW) [13,18]. Although this price level is still much higher than conventional competitive technologies such as internal combustion engines or gas turbines, the same cost was about three times higher around the year 2000 and two times higher in 2005, demonstrating a remarkable decrease progression [13,14]. This may suggest that the long path towards full industrialization and competitive costs for MCFC systems might realistically find a success in the next years [15,16,17].

The industrialization perspectives of MCFC are a key issue for the applications investigated in this work: although the integration of this kind of high temperature fuel cell in CCS plants could be very effective in terms of efficiency [12], it would require the deployment of large power FC installations (e.g., close to 100 MW class FC systems for a 500 MW NGCC, as shown in the next sections), not too far but significantly above the maximum power already demonstrated in stand-alone applications. By this point of view, the purpose of this work is to identify the cost range which should be targeted by MCFCs to become fully competitive when applied to large scale CCS from natural gas combined cycles in terms of cost of electricity ad cost of CO_2 avoided.

3. NGCC + MCFC integration

The power cycles discussed in this work are based on a natural gas combined cycle (NGCC), where a MCFC is placed downstream the gas turbine and ahead the heat recovery steam generator (HRSG). The gas turbine exhaust gases are used as cathode feed for the fuel cell, where CO_2 is transferred to the anode side, concentrating the CO_2 in the anode effluent. As already mentioned, the MCFC works with internal reforming; natural gas feeding the MCFC must be desulfurized by a proper treatment since reformer catalysts and MCFCs do not tolerate the presence of sulfur compounds (including NG odorizing additives) above 0.5–1 ppmv. Several options including Zinc-oxide absorption beds or active carbon filters could be considered; we have simulated here the economic outlook of the second option, relying on active carbons with metal impregnation, which does not require sulfur hydrogenation and can be regenerated with steam below 400 °C [19,20].

After the fuel cell, the anode effluent requires additional purification processes to recover the unconverted fuel species (still containing about 2 MJ/kg in terms of LHV) and achieve the stipulated CO_2 purity (i.e., >96% [21]). This paper compares two NGCC–MCFC plant configurations, based on two different CO_2 separation processes:

(1) Cryogenic option (case Cryo)

Provided that the CO_2 concentration at MCFC anode outlet is around 80% on a dry basis, the first possible configuration adopts a cryogenic system for performing CO_2 separation from the incondensable species included in the mixture. The anode stream is therefore cooled down to a temperature approaching the triple point of CO_2 (−56.6 °C), until most of the CO_2 condenses and can be separated by gravity; other fuel species with much lower boiling point remain in the gas phase. Details about this process, based on internal refrigeration cycles, and the related calculation assumptions are presented in [9], where it is shown that the process is able to attain an 89.3% CO_2 separation efficiency, delivering a stream whose purity approaches 98.8% mol. Along with separating CO_2, the cryogenic plant releases a remaining exhaust stream still containing the combustible species which were present at MCFC anode outlet. This stream is then sent along with natural gas to the gas turbine, representing about 9% of the total fuel input (LHV basis), increasing slightly the CO_2 fraction at GT outlet (5% vs. 4% of no capture case). This option is shown in Fig. 2, together with some examples of thermodynamic conditions and chemical composition in relevant points. For a complete list of properties the reader could refer to [12].

(2) Oxy-combustion option (Oxy)

In the second case, residual fuel components are burnt in a boiler with pure oxygen to avoid CO_2 dilution. An air separation unit (ASU) provides the required oxygen stream: it is assumed to produce 98% purity oxygen (the rest being 0.67% N_2 and 1.33% Ar) at atmospheric pressure through a cryogenic, double-column air separation process with a specific consumption of 0.295 kWh/kg_{CO2}, a value consistent with literature Ref. [22]. The resulting combustion products are mainly composed of H_2O and CO_2, and they are cooled down to recover heat for steam generation; after water condensation, the

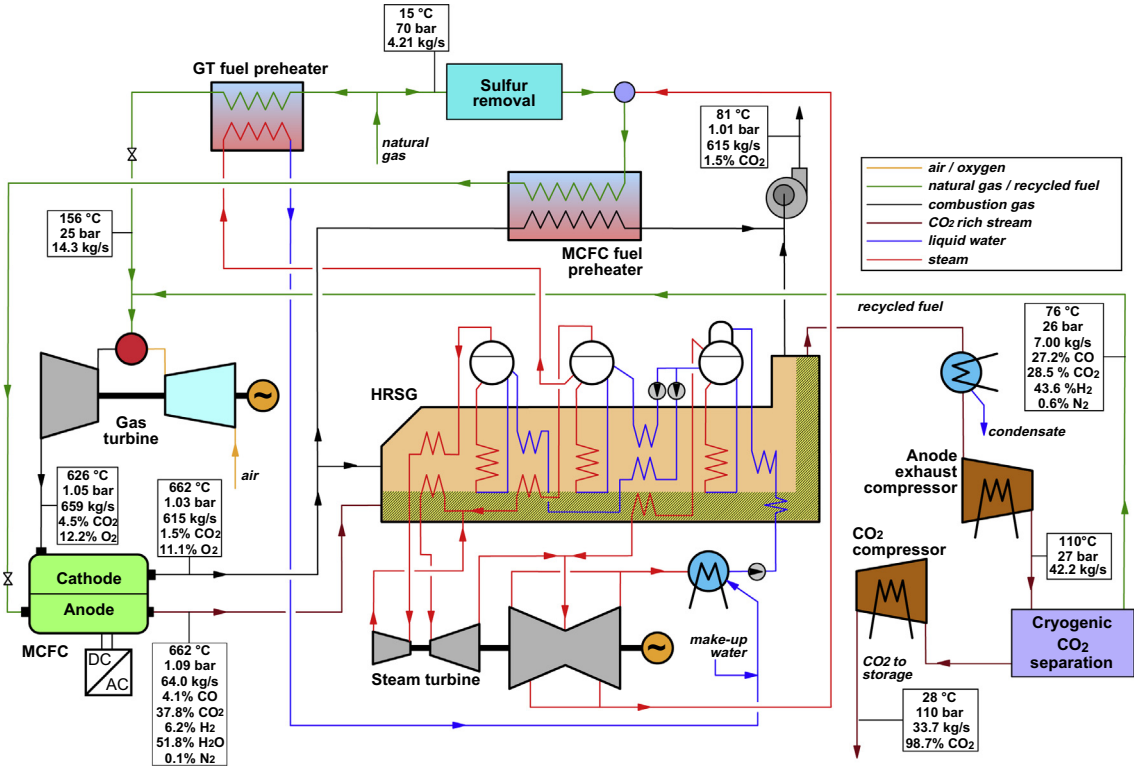

Fig. 2. NGCC + MCFC plant configuration with cryogenic CO_2 capture (Cryo).

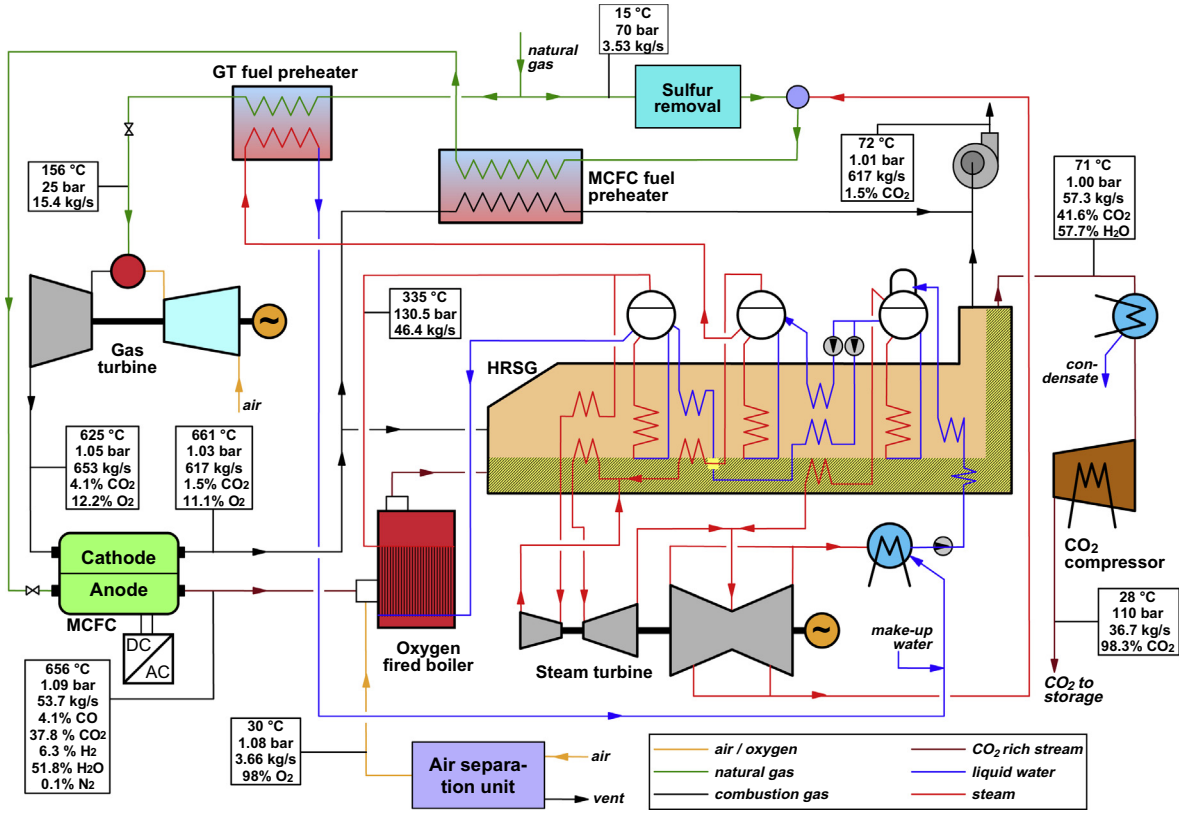

Fig. 3. NGCC + MCFC plant configuration with oxy-combustion CO_2 capture (Oxy).

fraction of incondensable species in dry CO_2 is below 4%.[2] This option is shown in Fig. 3.

In both cases the high purity CO_2 stream released by the separation process is compressed above supercritical pressure for subsequent storage. Additional possible configurations could be discussed; for instance, the MCFC could work with an external reforming process [23,24], or the stream exiting the HRSG could be partially recycled to the inlet of gas turbine compressor to increase the CO_2 concentration in the exhaust gases. These solutions have been investigated in previous works [9,10,12], showing positive and negative aspects. However, they generally feature a higher complexity; for the purpose of the economic analysis developed in this work we therefore focus on the two configurations introduced above, which can be considered more attractive according to the author's sensitivity and understanding.

3.1. Calculation model and MCFC simulation

Plant schemes are simulated with two software tools. The thermodynamic model of the assessed power cycles is carried out with the modular simulation code "*GS*", a tool developed at the Energy Department of Politecnico di Milano. It has provided highly accurate results in a variety of complex plant configurations, including all kinds of gas turbine cycles, combined cycles and hybrid cycles [25]. The code integrates models for the prediction of fuel cell performance [26,27] as well as built-in rules to calculate turbo machines (gas and steam turbines, compressors) efficiency as a function of their operating conditions [28], while the turbine cooled expansion is calculated by a stage by stage model described in detail in a previous paper [29]. For the sake of brevity we address the reader to the referenced publications to discuss the model details. Energy balance of the CO_2 separation section, including cryogenic processes and compression, is simulated with ASPEN Plus™ [30].

When assessing the plant thermodynamic and economic performances, reference is made in this paper to a natural gas fired combined cycle based on two heavy duty gas turbines, two heat recovery steam generator and one steam turbine. This 2 + 1 arrangement is chosen as an economic benchmark since it has become perhaps the most popular configuration among utilities, adding operational flexibility as required in a competitive electricity market [31]. The gas turbine calibration is based on "F Class" performances as discussed in [32] achieving a net electric efficiency of 58.34% at 829.9 MW_{el} power output (with specific emissions of 351.8 kg_{CO2}/kWh_{el}). This reference case also reflects the simulation assumptions adopted within latest EU FP7 projects on CO_2 capture by the European Benchmark Task Force (EBTF), where a NGCC based on a generic F-class gas turbine was considered [21]. Once included in the NGCC + MCFC plant configuration, the gas turbine slightly varies its operating conditions with respect to a NGCC due to: (i) a small increase of expander back pressure caused by the addition of fuel cell pressure losses and (ii) an increase in fuel mass flow rate due to the recycling of spent fuel to the gas turbine[3] in the plant configuration based on cryogenic separation (Fig. 3), which could be tolerated by the gas turbine control system, for instance, by slightly closing the compressor inlet guide vanes [33]. In this second case, a small increase of the gas turbine power output (partially balanced by the slightly higher back pressure) is expected as a consequence of the increased fuel mass flow rate. The power plant increases its power output, first of all thanks to the MCFC; due to that, and not to exceed the 100 MW size in the MCFC section, we have considered a case with a single GT in all NGCC + MCFC configurations. In the HRSG, the steam cycle operates with two heat sources featuring a higher total thermal input with respect to a baseline NGCC, due to reactant heating across the MCFC: (i) the exhaust gases exiting the MCFC cathode after CO_2 separation and (ii) the CO_2 concentrated stream exiting the cell anode. The second goes through oxygen-firing in the Oxy plant configuration of Fig. 3, resulting in an additional post-firing effect which further enhances the thermal power available in the HRSG. Globally, the power output of the steam section increases with respect to a baseline NGCC. A detailed list of calculation assumptions is reported in [12].

Simulation of the MCFC is carried out based on reactant properties at FC inlet (temperature, pressure, chemical composition and mass flow rate) and reactant utilization factors, using a lumped volume approach. The cell voltage is calculated from the reversible Nernst potential considering losses proportional to cell current density, with the approach discussed in detail in [9]. MCFC simulation parameters have been assumed to respect realistic operating ranges. Fuel utilization factor is set to 75%. The carbon dioxide utilization U_{CO2} (defined as the ratio between the flow rate of CO_2 transferred across the electrolyte through carbonate ions $CO_3^=$ and the CO_2 flow rate entering at the cathode) depends on the cycle configuration; it is adjusted during the simulations in order to keep the minimum CO_2 fraction at cathode outlet above 1.5%, an assumed operational limit suggested by a MCFC manufacturer [24].

3.2. Fuel desulfurization

Natural gas must be desulfurized prior to feeding the fuel cell to avoid poisoning of catalytic materials used at the anode as well as in the pre-reforming compartments integrated within the stack. Maximum tolerance to sulfur compounds is generally reported to be below 0.1 and 0.5 ppm to avoid performance degradation [13,14]. Different solutions are considered in literature for fuel desulfurization, including Zinc-oxide absorption beds or active carbon filters. We have considered here to simulate the technical features and the economic outlook of the second option, adopting a filtering unit based on activated carbon beds with metal impregnation [19]. With respect to ZnO beds, they do not require to preheat the fuel before cleaning (only the bed regeneration phase requires heating, as discussed below), and they do not require a preliminary mixing with hydrogen for sulfur compounds conversion to H_2S. A preliminary sizing, specifically developed with an industrial manufacturer [34] based on the plant specifications of Figs. 2 and 3, yields the results of Table 1. Such filters are dimensioned according to the volumetric flow rate of gas to be treated (about 17,750 Nm^3/h in the case Oxy, with a fuel mass flow rate of 4 kg/s). Assuming a concentration of sulfur compounds (primarily mercaptans odorizers and traces of H_2S) in input equal to 30 ppm, it is possible to reach an exit concentration of the order of 0.01 ppm, thus making the gas clean enough to enter the fuel cell.

The resulting desulfurization unit includes 4 filters, of which 2 in series under operation, and 2 other replacing the previous

Table 1
Design parameters for NG desulfurization with activated carbon beds [19,34].

NG flow rate (Oxy/Cryo) (kg/s)	4/4.6
Sulfur compounds at inlet/outlet (ppm)	30/0.01
Filter number/diameter (m)	4/0.9
Height of activated carbon bed (m)	2.20
Mass of activated carbon (@540 kg/m^3, kg)	3025
Gas speed (m/s)	0.3
Pressure losses (2 filters in series, kPa)	8.8

[2] According to literature on CO_2 sequestration in geological sites [27], further purification is not required for long-term storage in saline aquifers. Residual impurities in the CO_2 stream are essentially due to N_2 contained in natural gas, N_2 and Ar from the ASU and excess O_2 in the combustion process (added to ensure complete fuel oxidation).

[3] In previous works it is also considered the option of recycling part of the residual fuel stream to the fuel cell instead of to the gas turbine, which remains closer to the baseline NGCC design point.

Table 2
NGCC + MCFC plant performances.

		Ref. NGCC	Ref. MEA	Cryo	Oxy
MFCF U_f/U_{CO2}		–		75/69.5	75/65
MCFC cell voltage	V	–		0.709	0.702
MCFC current density	A/m^2	–		1000	1000
MCFC active area	m^2	–		140,762	118,159
GT electric output	MW_{el}	272.1(×2)	272.1(×2)	274.6	270.9
ST electric output	MW_{el}	292.8	215.7	161.3	184.2
MCFC electric output	MW_{el}	–		94.8	78.8
Auxiliaries					
HRSG/steam cycle	MW_{el}	−3.5	−3.4	−2.1	−2.5
Air separation unit	MW_{el}	–	–	–	−3.8
CO_2 compressor	MW_{el}	–	−22.6	−13.6	−12.4
Cryogenic compressors	MW_{el}	–	–	−5.5	–
Exhaust gas blower	MW_{el}	–	−15.0	−1.2	−1.6
Heat rejection	MW_{el}	−3.7	−4.4	−2.3	−2.7
MEA pumps and BOP	MW_{el}	–	−4.8	–	–
Net power output	MW_{el}	829.9	709.7	505.9	510.9
Fuel input to GT (LHV)	MW_{th}	1422.6	1422.6	664.1	715.0
Fuel input to MCFC (LHV)	MW_{th}	–	–	195.2	164.1
Net electric efficiency	%	58.34	49.90	58.88	58.11
CO_2 specific emissions	g/kWh_{el}	351.8	41.1	98.0	97.1
CO_2 avoided	%	–	88.3	71.3	71.6
Carbon capture ratio – CCR	%	–	90.0	71.9	72.5
SPECCA MJ/kg_{CO2}		–	3.36	−0.08	0.24

during a regeneration process. The latter lasts 4 h daily and requires superheated steam at 400 °C. In terms of mass flow rate, steam consumption would be equivalent to extracting continuously a flow rate of ~0.07 kg/s from the steam turbine; this would entail a loss of steam turbine power below 90 kW, therefore negligible compared to the net power output of the turbine, which exceeds 160 MW.

3.3. Plant performances

Results of plant performance calculations are summarized in Table 2. In all cases, the proposed NGCC + MCFC cycles with CO_2 capture feature remarkable performances with relevant advantages with respect to competitive CCS technologies, allowing to:

- Keep a very high plant net electrical efficiency: referring to NGCC, efficiency decay reduces to about 0.2% in the Oxy case and becomes negative with an efficiency gain of 0.5% in the Cryo configuration. Such advantage is obtained due to the efficiency gain achieved by MCFC integrated in the power cycle, thanks to the energy recovery from fuel cell exhaust heat operated by the HRSG and by the steam bottoming cycle.
- Limit the fuel cell share on total net power output at about 20%, much below typical values for other kind of fuel cell hybrid cycles based on gas turbines [8,26,28], where the FC contribute to power output is generally much higher (e.g., 60–80%). This condition yields a positive impact on plant economics which would otherwise suffer the high specific costs (€/kW_{el}) nowadays suffered by fuel cells.
- Obtain an increase of plant power output with respect to a baseline NGCC with the same gas turbine and no CO_2 capture. The variation would be around 20–30% considering a NGCC with a single gas turbine as reported in [12].

On the other hand, Table 1 shows also the plant carbon capture ratio CCR, which stands at about 72%, significantly lower than what can be achieved with other CCS strategies. This result is due to the MCFC operational limits which do not allow working with too high utilization factor of CO_2 fed to the cathode. As discussed in previous works [10,12] a possible way to overcome this limitation would be rising the CO_2 concentration in the exhaust fed to MCFC by a recirculation of flue gases to the GT inlet; this solution entails an efficiency decay and brings about a larger MCFC active area with higher costs, so that it is not considered for the economic analysis carried out here.

Along with traditional performance indexes, Table 2 also reports the SPECCA (Specific Primary Energy Consumption for CO_2 Avoided) index, that aims at defining with a single value plant achievements in terms of electric efficiency and CO_2 specific emissions. SPECCA is therefore defined as:

$$\mathrm{SPECCA}\left[\frac{MJ}{kg_{CO_2}}\right] = \frac{\left(\frac{1}{\eta}-\frac{1}{\eta_{ref}}\right)}{E_{ref}-E}\times 3600 \quad (1)$$

where η is the net electric efficiency of system and E is the CO_2 specific emission (kg_{CO2}/MWh_{el}) of the system with CO_2 capture, while 'ref' subscript refers to the reference NGCC plant. It measures in a comprehensive way the amount of fuel thermal energy required to avoid the emission to ambient of one kg of CO_2. It is much lower for the MCFC plants than for a conventional MEA system (at least one order of magnitude), and it is negative for the Cryo case according to the fact that Cryo has a higher efficiency than NGCC even though a significant fraction of the CO_2 produced is captured. Thus, the SPECCA index shows that the NGCC + MCFC solution is by far more effective than conventional MEA option in separating CO_2, despite its lower total CCR. This concept will be also addressed afterwards the economic analysis in Section 5.

4. Economic analysis

The economic assessment performed in this work is based on the 'bottom-up' approach, assumed by several institutions and research groups as a reference method for investigating the economic outlook of a new technology. It is based on the principle of splitting the plant costs based on all its subcomponents, isolating all the contributions and finally highlighting the role of the most uncertain costs in the overall estimation. Based on this analysis, we will be able to evidence the role of the MCFC on the plant

Table 3
Equipment costs for main components [31,35,39–42].

Plant component	Scaling parameter	Reference erected cost C_0 (M€)	Reference size, S_0	Scale factor, f	N
HRSG, ducting and stack	$U \times S$ (MW/K)	32.6	12.9	0.67	1
Steam turbine, generator and auxiliaries	$ST_{Gross\ Power}$ (MW)	33.7	200.0	0.67	1
Cooling water system and BOP	Q_rejected (MW)	49.6	470.0	0.67	1
CO_2 compressor and condenser	Compressor power (MW)	9.9	13.0	0.67	1
Desulfurization filters	Thermal input LHV (MW)	0.66	413.82	0.67	2
Air separation unit (ASU)	Oxygen produced (kg/s)	26.6	28.9	0.7	1
Cryogenic heat exchangers	Cooling duty (MW)	0.8	32	0.9	1
Heat exchangers	Heat transferred (MW)	1.8	57.2	0.9	1

economics, first in terms of capital costs (due to the presence of the MCFC itself as well as to the change of costs brought about on other components) then in terms of final cost of electricity and cost of CO_2 capture for the proposed plant configurations.

The step-by-step procedure resulting from this approach to the economic analysis reflects the European Benchmarking Task Force (EBTF) methodology [35,36]. The EBTF was created by the European Commission to unify the modeling methodologies of the European projects involved in Carbon Capture within the 7th framework R&D Programme. The two main economic parameters chosen by EBTF as term of comparison among different CO_2 capture technologies are the cost of Electricity (COE) and the cost of CO_2 avoided (CCA), which are also assumed here as preferred indexes for the evaluation of the economic viability of the proposed power plants.

The COE is calculated with the IEA methodology [37,38] which sets the net present value (NPV) of the power plant to zero. This can be achieved by varying the kWh price until the revenues balance all the expenses over the whole life time of the power plant. The cost of electricity in this analysis does not include CO_2 transport and storage costs.

The cost of CO_2 avoided CCA is defined as the additional COE cost to avoid one kg of CO_2:

$$\mathrm{CCA}\left[\frac{€}{kg_{CO2}}\right] = \frac{(\mathrm{COE})_{CO2\ cap} - (\mathrm{COE})_{ref}}{(kg_{CO2}\ KWh^{-1})_{ref} - (kg_{CO2}\ kWh^{-1})_{CO2\ cap}} \quad (2)$$

where CO_2 *cap* is the considered power plant with CO_2 capture and *ref* stands for the reference plant without CCS.

As for COE; the cost of CO_2 avoided does not take into account CO_2 transport and storage costs. Reference costs for transport and storage are in the range of 1–4 $/$t_{CO2}$ and 6–13 $/$t_{CO2}$ respectively, depending on the type of storage assumed and on power plant distance from the storage site [39].

The economic assessment requires the definition for each considered plant of the total plant investment, fixed operation and maintenance, consumable and fuel costs.

The calculation of total plant investment (TPI) is carried out with the so-called bottom-up approach (BUA) which consists of breaking down the power plant into basic equipment or components, assuming a proper costing model for each of them, and finally adding installation, indirect costs and contingencies.

The first step in TPI evaluation consists of calculating the Total Equipment Costs (TEC) as the sum of the cost of every plant components. For each subsystem or component, a scaling parameter is selected and the actual cost C is derived starting from the cost C_0 of a reference component having size S_0 by the relationship:

$$C[\mathrm{M€}] = n\ C_0[S/(n\ S_0)]^f \quad (3)$$

where S is the actual size, f is the scale factor and the coefficient n is the number of components for the base case. Adopted equipment costs are shown in Table 3.

Two components are excluded from the law of Eq. (2) and Table 3:

- The gas turbine, whose equipment cost has been kept constant at 49.4 M€ for all the plants, since the gas turbine size is always the same. Assumption of a constant cost reflects the possibility to consider the proposed MCFC cycles as a 'retrofit' at least with respect to the gas turbine, which remains as a topping power plant – basically untouched with respect to the original combined cycle. This assumption neglects only the small difference in power output deriving from change in exhaust backpressure and fuel composition (the latter just for the Cryo case as explained in Section 2). The specific gas turbine equipment cost was taken from reference literature [40,41] and consistently with [31,36].
- The MCFC, whose Total Equipment Cost (TEC) was assumed equal to 2700 €/kW_{el} according to indications for current costs in [43,44]. A large number of parallel MCFC modules have been assumed to achieve the total power output required in the plants, provided that the current largest commercial module is 2.8 MW. For this reason no scale economy has been considered for cost evaluation (i.e., $f = 1$). However, much lower costs are projected for future evolutions of the FC technology: as a possible term of comparison, a US DOE study dedicated to fuel cell based CCS plants assumes a cost of 500 $/$kW_{el}$ for SOFCs [11]. Aiming to discuss this perspectives, we will carry out a sensitivity analysis in Section 4 where the fuel cell cost is decreased down to 1000 €/kW_{el}.

As far as the HRSG and steam plant is concerned, their operating conditions and power duties change remarkably with respect to the baseline combined cycle due to the addition of the MCFC, which releases exhaust heat to the bottoming steam plant through both the cathode and the anode stream. We therefore include them into the economic variables through a model which evaluates:

(a) HRSG, ducting and stack costs as a function of the total heat exchange relative duty, expressed in MW/K, given by the product $U \times S$ of the total heat exchange coefficient (U, MW/m^2/K) and the heat exchangers surface (S, m^2). The total $U \times S$ can be found by adding the $(U \times S)i$ of each heat exchanger in the triple pressure, reheat type HRSG, which are calculated based on the corresponding thermal power and gas/steam operating temperatures, obtained by the simulation model discussed at point 3.1.
(b) Steam turbine, generator and auxiliaries costs as a function of the steam turbine gross power. Details about steam turbine simulation are given in [12] and omitted here for brevity.

All the values coming from different sources have been referred to the same year (2008) by means of the CEPCI deflator index [35], while a 0.8 €/USD change rate has been assumed.

Installation costs (INST) such as piping, erection, external connections outside battery limits, etc. are calculated as a percentage of TEC equipment. This percentage is set according to a unit can be

envisaged as a chemical process (requiring higher installation costs) or a power block. Coefficients, evaluated as weighted average from values adopted in [39,40], were set equal to 80% of TEC for MCFC and CO_2 removal section and 68% for power section. As far as the sulfur removal is concerned, costs are assumed after the preliminary sizing considered in Section 3.2.

Indirect costs (IC) which account for yard improvement, service facilities, buildings and engineering are calculated as a fixed percentage (14%) of total direct plant costs (TDPC = TEC + INST). Engineering, procurement and construction costs (EPC) is the sum of TDPC and IC. The total plant cost (TPC also designed as overnight capital) results from EPC plus owner's cost and contingencies (OCC) due to plant planning, designing and commissioning together with contingencies. OCC are fixed to 15% (a typical value for a *n*-th of a kind plant) of the EPC cost for all the technology options according to [36].

Interest during construction (IDC) is assumed at 7% of TPC, based on a 3-year construction schedule with 40%, 30%, 30% allocation of annual payments, and a discount rate of 8%/year. Finally total plant investment is calculated as TPI = TPC + IDC.

As final comment about this procedure, we note that BUA is the usual approach followed when the plant include novel components (the MCFC in the present case) that prevent adopting a Top-Down Approach (TDA) based on equipment supplier estimates of the entire EPC costs. BUA was similarly adopted by NETL in a recent work [11] concerning integration of fuel cells in coal fired power plant.

Other main economic assumptions for COE calculation are shown in Table 4 along with fuel, operation and maintenance (O&M) and consumables costs.

Labor costs refer to an average European situation, assuming that about 20 people are usually necessary to run a natural gas fired combined cycle power station. Capture cases require a higher labor cost because of the plant complexity that needs dedicated employees to supervise the CO_2 removal section. The assumed labor cost is lower than the one assumed in EBTF, but it reflects the actual situation. Maintenance and insurance costs are taken from IEA [35,36].

An additional yearly operating cost equal to 5% of the TPC of the fuel cell modules [43,46] has been added to consumables for the MCFC based plants to account for the short lifetime of the stack, assuming it requires a complete replacement every 5 years; this corresponds to a total cost for stack substitution equal to 25% of the TPC occurring every 5 years (vs. 25 years plant operating span) [45]. A sensitivity analysis in the last section will discuss the effect of this assumption on final results.

5. Results and comparison with conventional capture methods

Table 5 reports the main performance and the economic evaluation regarding four plants:

Table 4
Main economic assumptions.

Discount rate	%	8
First year operating hours	h	5700
Rest of lifetime operating hours	h	7500
Operating lifetime	years	25
Natural gas costs	€/GJ_{LHV}	6.5
O&M costs		
Labor costs, no capture case	M€/y	1.2
Labor costs, capture case	M€/y	1.8
Maintenance costs	% of TPC/y	2.5
Insurance	% of TPC/y	2
Consumables		
MCFC stack periodical substitution	% of TPC/y	5%
Evaporative tower blow-off	% of evap.water	100
Cooling water make-up cost	€/m^3	0.35
HRSG water blow-off cost	€/m^3	1
Process water cost	€/m^3	2

- NGCC: a natural gas fired combined cycle with two gas turbines and one steam turbine configuration, without CO_2 capture, assumed in this paper as reference technology for economics and performances in electricity production from natural gas [42].
- MEA: a natural gas fired combined cycle with the same two GT and one ST configuration of the baseline NGCC, with the addition of post-combustion CO_2 capture by amine scrubbing, removing 90% of CO_2 from the exhaust, assumed as benchmark for evaluation of the potential of MCFC-CC technology in CCS application from natural gas [42].
- The two configurations Cryo and Oxy calculated in [12] and assessed in the present paper, focusing on a single GT configuration.

A first observation is that we compare here power plants having a different power output. On one hand, the rather diffused 800 MW_{el}-class NGCC configuration with two GTs and one steam cycle, and the 700 MW_{el} plant resulting from the application of CCS with MEA capture. The latter suffers a heavy power reduction due to steam extraction from the steam turbine for solvent regeneration as well as due to the heavy auxiliary consumption (see Table 2; for instance the exhaust gas blower requires about 15 MW due to the high pressure losses in the capture loop). On the other hand, we are considering a NGCC + MCFC plant with a single gas turbine, with about 500 MW_{el} of net power, 80–95 MW of which from the fuel cell. The latter has been chosen as a more plausible solution than a two-GT NGCC + MCFC plant, where the overall net power would have exceeded 1 GW with a twice larger fuel cell section.

Design specs and performance of NGCC and MEA plants are congruent with the same EBTF assumptions [32,36] used for Cryo and Oxy plant assessment, and reflect the calculation model discussed in Section 3.1 and in [12].

As already evidenced in Section 3.3, thermodynamic results showed the remarkable potential of MCFC technology for capturing the CO_2. By the point of view of economics, the situation is different, at least with contemporary MCFC costs.

The breakdown of costs shown in Table 5 allows to highlight the changes in terms of capital costs brought about by the presence of the MCFC. On one hand the Total Equipment Costs show the heavy role of the MCFC itself (212 and 256 M€ in Oxy and Cryo respectively), on the other hand also the steam turbine and HRSG sections are changing their share of TEC since the HRSG handles the additional thermal power input released by the MCFC in the cathode and anode exhaust streams. This last effect is particularly important in the Oxy plant where the residual heating value of the anode exhaust is converted into heat in the oxygen-fired boiler ahead the HRSG, so that also the total cost of the HRSG is higher. As evidenced in Fig. 4, the percentage distribution of TEC changes substantially with respect to the original NGCC as well as with respect to the NGCC + MEA plant. The latter features a particularly high BOP cost which includes the section of CO_2 scrubbing through the MEA solvent, including absorption and regeneration columns and related auxiliary components.

As far as the costs for auxiliary components are concerned, in all cases heat rejection includes costs of condenser, evaporative towers (assumed as cooling source), draining system and related pumps. CO_2 compressors power and costs for the MCFC plants are higher in the Cryo case since they include the contribute of the anode exhaust compressor placed ahead the CO_2 separation process.

Results in the lower part of Table 5 show that, at the assumed Total Equipment Cost (TEC) of the MCFC (2700 €/kW_{el}), COE of the MCFC based plants are significantly higher than MEA due to their elevated capital component. This is can be regarded to as a contemporary or short-term perspective, where MCFC are installed

Table 5
Results of energy and economic performance assessment (MCFC specific cost = 2700 €/kW_{el} in TEC calculation).

Main plant component power (from Table 2)	NGCC	MEA	Cryo	Oxy
GT electric output (MW_{el})	272.1(×2)	272.1(×2)	274.6	270.9
ST electric output (MW_{el})	292.8	215.7	161.3	184.2
MCFC electric output (MW_{el})	–		94.8	78.8
Auxiliary power (MW_{el})	−7.1	−50.2	−24.8	−23
Net power output (MW_{el})	829.9	709.7	505.9	510.9
Plant component TEC, M€	**NGCC**	**MEA**	**Cryo**	**Oxy**
Gas turbine	98.8	98.8	49.4	49.4
Steam turbine	43.2	35.1	29.2	31.7
MCFC	–	–	256.0	212.8
HRSG	45.7	44.8	30.2	33.3
Air separation unit (ASU)	–	–	–	6.2
Heat rejection	49.4	54.9	35.8	37.5
CO_2 compressor	–	14.4	10.3	9.7
Exhaust gas blower (only NGCC + MCFC)	–	–	1.3	1.3
BOP (includes exhaust blower for MEA)	0.4	59.4	7.8	13.1
TEC: Total Equipment Cost, M€	237.5	307.4	420	391.1
EPC: engineering, procurement and construction cost, M€	454.9	598.7	840	791.4
TPC: total plant cost, M€	523.1	688.5	966	898.6
IDC: interest during construction, M€	36.2	48.2	71	66.0
TPI: total plant investment, M€	559.3	736.7	1038	964.6
Specific total plant cost, €/kW_{el}	630.4	969.9	1910	1777
COE components, €/MWh				
Capital	9.55	15.59	32.50	29.83
Fixed O&M	3.85	5.24	5.62	5.24
Consumables	0.59	1.43	6.80	5.79
Fuel	40.11	46.89	39.74	40.67
Total cost of electricity, €/MWh	54.76	69.10	84.66	81.53
Cost of CO_2 avoided, €/ton_{CO2}	–	48.5	120.4	107.7

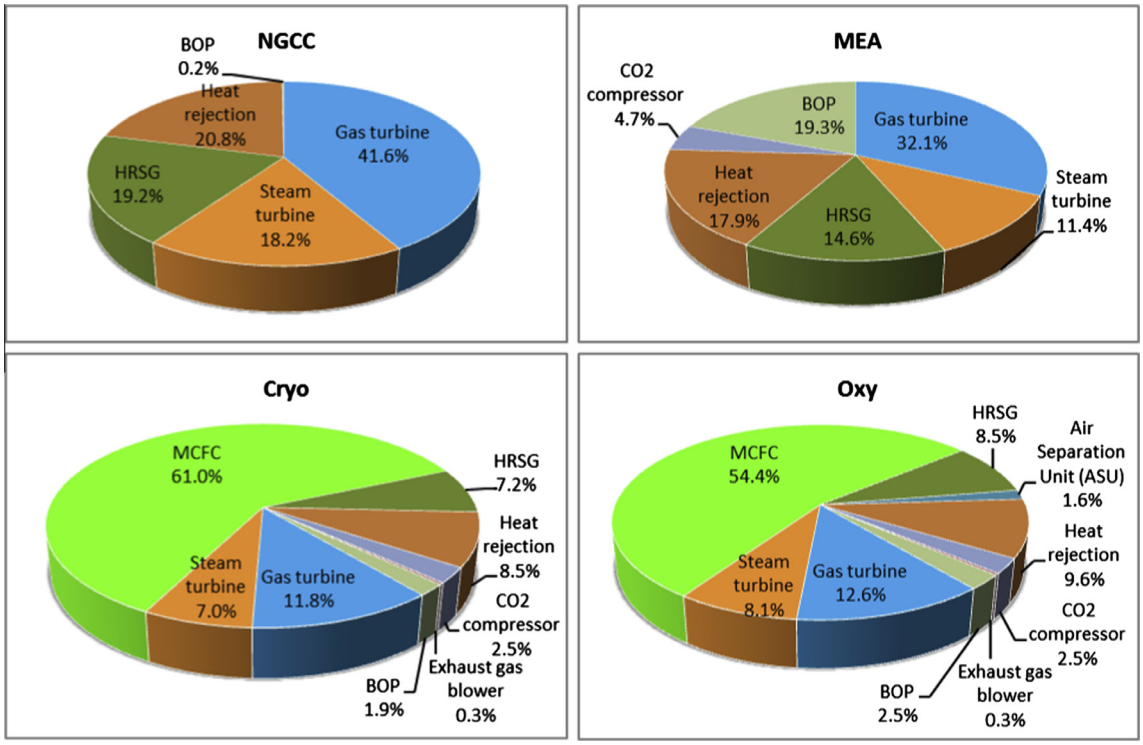

Fig. 4. Total Equipment Costs (TEC) share of the main components in the assessed plants.

at their current high cost. The resulting cost of CO_2 avoided is higher than 100 €/ton, more than double the cost of MEA plant which takes advantage from a higher CO_2 capture ratio.

Under the current assumptions the Oxy configuration achieves lower COE and CCA than Cryo. This is essentially due to the lower MCFC power output (78.8 vs. 94.8 MW) leading to a significant reduction of the associated costs (initial equipment purchase and ensuing replacement).

The reason of this difference is that in the Cryo case a fraction of the spent fuel at the MCFC anode side exit is eventually

recirculated to the gas turbine combustor. More CO_2 is contained in the gas turbine exhaust gases and hence a larger MCFC capacity has to be installed to achieve a comparable CO_2 capture ratio. At the assumed NG price (6.5 €/GJ) these cost components cannot clearly be traded off by the lower fuel cost allowed by the higher efficiency of the Cryo solution.

Of course, the picture can change and even be completely reversed if the MCFC cost decreases substantially, as foreseen by studies related to the mid-term and long-term future evolution of fuel cell technology [43]. Moreover the results are strongly influenced by the fuel cost assumptions. Fig. 5 shows the results of a sensitivity analysis aiming at evaluating the cost of electricity as MCFC cost and fuel price are varied.

At NG price of 4–6.5 €/GJ, the equivalence in the cost of electricity between MEA and Oxy is reached for a specific cost of the MCFC unit around 1300–1400 €/kW_{el}, that is about half the current quotation. This equivalence level moves to 1600 €/kW_{el} for a scenario with high NG price (9 €/GJ), as a consequence of the higher efficiency of the Cryo plant which reduces the COE component due to fuel consumption.

More meaningful is the result illustrated in Fig. 6 that compares the different low CO_2 emission plant options in terms of CCA assuming an intermediate fuel price of 6.5 €/GJ.

Because of the higher CO_2 capture ratio achieved by the MEA configuration, the MCFC based plants reach an equivalent CCA of the MEA plant for an equipment cost of MCFC unit in the range 1000–1100 €/kW_{el}.

The decrease of CCA with fuel cell costs is rather sharp, so that a substantial advantage would arise for MCFC specific costs in the range of 700 €/kW_{el}. It is interesting to note that at such low values of the MCFC cost, the higher efficiency of the Cryo configuration leads to a better economic outcome than the Oxy solution.

As already pointed out in a previous work [12], one major drawback of the NGCC + MCFC plant is the significantly lower overall CO_2 capture ratio (70–75% in Table 2) if compared to other strategies, like the 90% level easily reached by the MEA option of Table 5. This happens since it is not possible to transfer across the MCFC a too high portion of the CO_2 fed at the cathode, or in other words U_{CO2} is limited, due to: (i) on one hand, the difficulties in operating at very high reactant utilization factors featured by any fuel cell: when operating with diluted reactants, their consumption can not be too high to avoid excessive local depletion of reactants at cell reaction sites, which adversely influences the cell voltage;

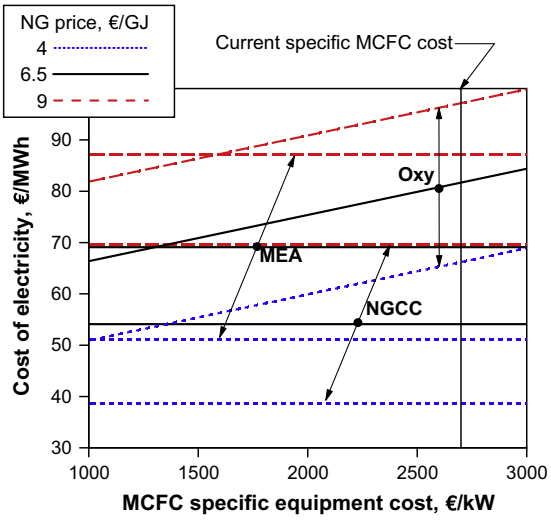

Fig. 5. Sensitivity analysis of COE with MCFC specific equipment cost (used to calculate TEC).

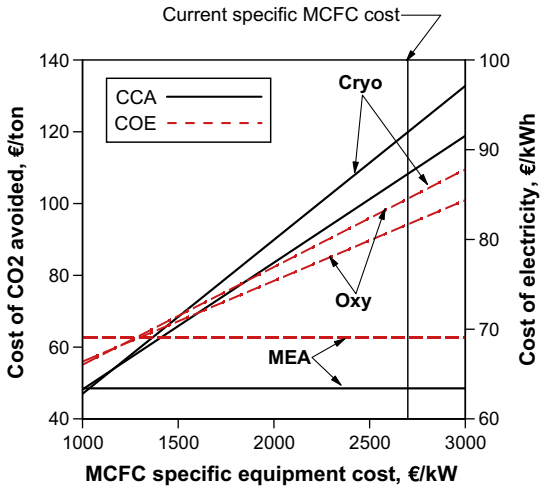

Fig. 6. Sensitivity analysis of CCA with MCFC specific equipment cost (used to calculate TEC).

and (ii) a specific technological limitation of MCFCs where a minimum fraction of CO_2 is requested to keep stable conditions of the molten carbonates facing the cathode stream and to avoid a fast degradation of the electrolyte, preventing a severe voltage decrease. Only few experiments are reported in literature about the behavior of MCFCs in the operating conditions required for CCS [23,24,47,48], so that future development of the MCFC technology could allow pushing further this limitation.

A possible way to improve the CCR, keeping a constant CO_2 utilization factor, would be to raise the CO_2 concentration in the gas turbine exhaust which are fed to MCFC by adopting a recirculation of flue gases to the GT inlet. This solution can increase CCR up to about 85–90%, with the negative drawback of efficiency decay (e.g., 1.5–2% in terms of electrical efficiency, due to the a rapid decrease of the oxygen content in the gas turbine exhaust which negatively affects MCFC voltage) and requiring larger MCFC active area as investigated in a previous work [10]; the economic outlook of this solution, which entails higher costs in the MCFC section, could be discussed in future works. Alternatively, it could also be proposed to add a MEA section at MCFC cathode outlet, working on the residual CO_2 fraction.

Anyhow, it must be evidenced that the CO_2 capture ratio is only one of the indexes describing the effectiveness of CCS systems, and not the most comprehensive. The best indication is given by the SPECCA index, which – as already pointed out – shows that the NGCC + MCFC plants would be by far more effective than conventional MEA plants in separating the CO_2, up to the fraction that they can separate. In other words, even a 'low' carbon capture rate, if obtained very 'easily' (by an efficiency point of view, thus implying low energy penalties), makes the system attractive; and this is the situation for the NGCC + MCFC plant.

Another significant issue in the successful deployment of the MCFC technology is related to life extension. Due to progressive decay of performances, we have assumed here that the MCFC stacks are substituted every five years, accounting for 25% of the fuel cell section TPC. Aiming to better evaluate the effects of this assumption, it is possible to develop a sensitivity study where the MCFC life extension and periodical stack substitution period are changed from every three to every ten years, verifying the results on overall balances. Results in terms of COE are shown in Fig. 7 focusing on the Oxy plant, evidencing that a significant increase of MCFC cost could be tolerated while keeping competitiveness towards the MEA solution when extending the stack lifetime to 10 years.

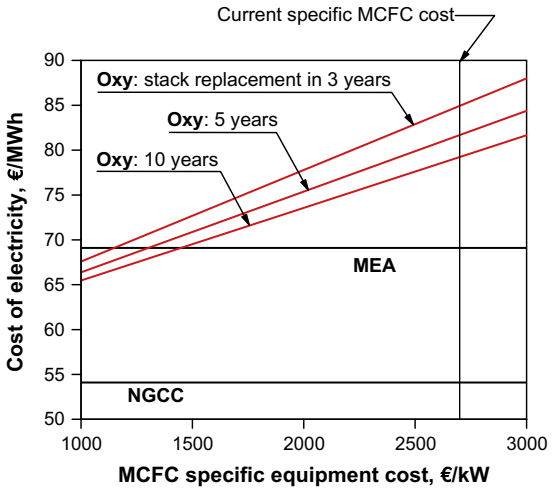

Fig. 7. Sensitivity analysis of COE with stack lifetime and replacement interval for the Oxy plant, as a function of MCFC specific equipment cost (used to calculate TEC).

The other relevant issue regards the technical feasibility of the NGCC + MCFC solution, provided that current largest installations have a size (see Section 2) sensibly lower than that required for this application; only small-scale MCFC units have been dedicated up to now to demo plants for CCS [49] although new activities are foreseen within DOE projects [50]. Moreover, as shown in Figs. 4 and 5, the high specific costs presently featured by this type of fuel cells heavily affects the economics of the NGCC + MCFC concept, whereas conventional CCS strategies nowadays offer lower investment costs and a substantially better economic feasibility (a wider technical comparison with pre-combustion and post-combustion strategies is discussed in [12]).

Even taking into account all these considerations, the potential of the NGCC + MCFC concept seems very promising.

6. Conclusions

This work discusses the economic perspectives of an innovative solution for CO_2 capture from combined cycles, relying on the integration of Molten Carbonate Fuel Cells. Two NGCC + MCFC plants are considered, following a previous detailed modeling activity which individuated the most promising configurations.

In both cases, the fuel cell is placed downstream the gas turbine and ahead the heat recovery steam generator (HRSG). The gas turbine exhaust gases are used as cathode feed for the fuel cell, where CO_2 is transferred across the electrolyte, concentrating the CO_2 in the anode effluent. Exhaust heat in the cell effluents is recovered by the bottoming steam cycle.

The two plants follow different approaches for purification of the CO_2-rich stream exiting the cell anode: (i) a cryogenic process that separates CO_2 from the residual combustible species which are recycled back to the gas turbines, or (ii) an oxy-combustion of residual combustible species, followed by heat recovery, cooling and water separation by condensation. Both these plant configurations can capture up to 70–75% of CO_2 with negligible efficiency variation compared to a baseline combined cycle, while increasing by about 20% the overall power output. Thanks to these positive features, the NGCC + MCFC integration could have remarkable advantages when compared to competitive carbon capture technologies.

The economic analysis is based on a detailed bottom-up approach to determine the distribution of equipment costs and the resulting total plant costs.

Results show that at current MCFC specific costs (2700 €/kW_{el} in terms of TEC), which are still very high, competition with reference CCS technologies based on ammine scrubbing (MEA configuration) is not attractive: COE of the MCFC based plants and the resulting cost of CO_2 avoided are significantly higher than MEA due to their elevated capital component. Under this cost assumptions the Oxy configuration achieves lower COE and CCA than Cryo, thanks to the lower MCFC power output (78.8 vs. 94.8 MW) leading to a significant reduction of the associated costs.

A successful competition with conventional technologies could be achieved for a lower MCFC specific cost, falling in the range 1000–1500 €/kW_{el} (depending on natural gas cost), a target which could be reached by future evolution of the MCFC technology. In these forward-looking cases the higher efficiency of the Cryo configuration leads to a better economic outcome than the Oxy solution. The trade-off with conventional NGCC + MEA is at 1600 €/kW_{el} for an economic scenario featuring a very high natural gas cost (9 €/GJ), while it goes around 1300–1400 €/kW_{el} for a NG price of 4–6.5 €/GJ. In these cases, the NGCC + MCFC concept would become a winning solution also by the point of view of economics.

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