APPLICATION OF MCFC IN COAL GASIFICATION PLANTS FOR HIGH EFFICIENCY CO_2 CAPTURE

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ABSTRACT

Integrated gasification combined cycles (IGCCs) are considered the reference technology for high efficiency and low emission power generation from coal. In recent years, several theoretical and experimental studies in this field have been oriented towards capturing CO₂ from IGCCs through the integration of Solid Oxide Fuel Cells (SOFC) for coal-syngas oxidation, investigating the so-called Integrated Gasification Fuel Cell cycles (IGFC). However, Molten Carbonate Fuel Cells (MCFC) can also be a promising technology in IGFCs. After a rather comprehensive research carried out by the authors on modeling and simulation of SOFC-based IGFC plants, an interesting IGFC cycle based on MCFC is assessed in this work, where plant layout is designed to exploit the capability of MCFCs of transferring CO₂ and O₂ from the oxidant side to the fuel side. Syngas produced in a high efficiency Shell gasifier is cleaned and mainly burned in a combustion turbine as in conventional IGCCs. Turbine flue gas, rich of oxygen and carbon dioxide, are then used as oxidant stream for the fuel cell at the cathode side, while the remaining clean syngas is oxidized at the anode side. In this way the MCFC, while efficiently producing electricity, separates CO₂ from the gas turbine flue gas as in a post-combustion configuration; oxygen is also transported towards the anode side, oxidizing the remaining syngas as in an oxy-combustion mode. A CO₂-rich stream is hence obtained at anode outlet, which can be cooled and compressed for long term storage. This configuration allows to produce power from coal with high efficiency and low emission. In addition, as already highlighted in a previous study where a similar concept has been applied to natural gas-fired combined cycles, a limited fraction of the power output is generated by the fuel cell (the most expensive component), highlighting its potential also from an economic point of view. Detailed results are presented in terms of energy and material balances of the proposed cycle.

NOMENCLATURE

AGR	Acid Gas Removal
ASU	Air Separation Unit
CCR	Carbon Capture Ratio
CGE	Cold Gas Efficiency
DC	Direct current
FC	Fuel Cell
FGD	Flue Gas Desulfurizer
G	Mass flow rate (kg/s)
GT	Gas Turbine
HHV	Higher Heating Value
IGCC	Integrated Gasification Combined Cycle
IGFC	Integrated Gasification Fuel Cell cycle
LHV	Lower Heating Value (MJ/kg)
MCFC	Molten Carbonate Fuel Cell
NG	Natural Gas
Р	Power (MW)
р	Total pressure (bar)
R	Gas constant: R=8.314 J/(K·mol)
SOFC	Solid Oxide Fuel Cell
SPECCA	Specific Primary Energy Consumption for CO ₂ Avoided
Т	Temperature (°C or K)
TIT	First turbine rotor total inlet temperature (°C)
TOT	Turbine outlet temperature (°C)
Ua	Air utilization factor: U _a =O _{2,consumed} / O _{2,inlet}
U_{CO2}	CO ₂ utilization factor: U _{CO2} =CO _{2,transferred} / CO _{2,anode inlet}
U_{f}	Fuel utilization factor: $U_f = (H_{2,consumed})/(H_{2,equivalent in})$
V	Fuel cell potential (V)
WGS	Water Gas Shift
Х	Molar fraction
ΔH	Enthalpy change (kJ/mol)
ΔT	Temperature change (°C)

Subsc	ripts
a	air
e	electric
in	inlet
f	fuel
REF	referred to the reference case
th	thermal

1. INTRODUCTION

Application of Solid Oxide Fuel Cells (SOFC) and Molten Carbonate Fuel Cells (MCFC) appears particularly promising for future power plants with CO_2 capture. The reason of their potentiality is the inherent separation of oxygen from air which occurs during the fuel oxidation process. Consequently, the resulting oxidized gas from SOFCs and MCFCs is not diluted with nitrogen and CO_2 separation is easier and less energy consuming than in more conventional post-combustion and pre-combustion technologies. The products of fuel oxidation are hence similar to those resulting from oxy-combustion processes, where, however, O_2 is produced by cryogenic distillation with high electric consumptions. On the contrary, SOFCs and MCFCs are high efficiency systems where oxygen is separated without any extra consumption.

While SOFCs and MCFCs have been widely tested using natural gas as primary fuel, their application in coal-fired plants would represent a turning point in the power generation sector thanks to the possibility of generating power at high efficiency and with minimal emissions from a low-cost, abundant but typically polluting fuel like coal. Large R&D efforts have been focusing in recent years on the development of SOFC-based Integrated Gasification Fuel Cell (IGFC) cycles by research institutes [1-3] and manufacturers [4,5]. In previous works, the authors also investigated the potentialities of different SOFC-based IGFC configurations, with a rather comprehensive modeling and simulation activity [7-9], obtaining interesting results and net efficiencies of 54.3% and up to 51.6% for plants without and with CO₂ capture respectively.

Molten Carbonate Fuel Cells were also proposed for application on coal derived syngas without [3,10,11] and with CO_2 capture [12]. In [12], for example, Jansen et al. considered a plant based on a Shell gasifier, high temperature syngas cleaning and a membrane-based separation of unconverted H₂ from CO_2 at anode exhaust separation, calculating a net efficiency of 47.5%, with 97.4% CO_2 capture.

Another option for the integration in power plants with CO_2 capture is to use the MCFC as an active (i.e. non consuming) CO_2 concentrator. In MCFCs, in addition to oxygen, also CO_2 must be provided to the cathode (Fig.1), since oxygen is transported as carbonate ion $CO_3^{=}$ to the anode side, where it is released for fuel oxidation. Hence, CO_2 in the anode exhausts derives both from the fuel oxidation and from the CO_2 used as O_2 carrier in the electrolyte. In stand-alone applications, a fraction of the anode exhaust stream is usually burned in a catalytic combustor and recycled at the cathode

inlet to sustain the formation of carbonate ions on the electrode. Another option would be to use the flue gas of a power plant, already containing CO_2 and O_2 , as oxidizing agent in a MCFC, concentrating in this way the CO_2 from the diluted flue gas side to the more concentrated anode side.



Figure 1 – Operating principle of a MCFC

This concept has already been proposed in some works [14,15] and gained the interest of manufacturers and power producers [13-17]. A comprehensive thermodynamic analysis was recently carried out in the research group of the authors, by considering the flue gas of a natural gas-fired gas turbine as cathode gas [18-20]. Different configurations to exploit the heating value of the unconverted fuel at anode outlet (oxy-combustion and cryogenic separation) and different MCFC characteristics (with and without internal reforming) were considered in these works. The results obtained confirmed the validity of the concept and interesting considerations could be made:

- High efficiencies (57.2-58.6%) were calculated for all the configurations, with penalties of 0-1.5% points with respect to the reference combined cycle without CO₂ capture. On the other hand, a power output increase of 20-25% was obtained when operation with a given gas turbine of a certain size is considered.
- Carbon capture ratios (CCR) of 60-85% were obtained. Such relatively moderate values depend on the limitations on CO₂ concentration at the anode, which cannot be too low for acceptable performance of the FC.
- As a result, much lower specific primary energy consumptions for CO₂ avoided (SPECCA) were obtained with respect to competitive post-combustion capture technologies.
- A favorable share of power generation between the plant components was obtained, with most of the gross power (80-85%) produced by the lower specific cost components (gas turbine and steam cycle) and the remaining 15-20% by the MCFC. A relatively limited

impact on investment costs can therefore be expected from the introduction of the MCFC.

The same concept can be considered also for coal-fired plants. However, the operation on flue gas from a coal boiler presents some drawbacks, namely: (i) flue gas should be deeply purified from particulate, SO_x and other species harmful for the FC; (ii) flue gas must be cooled for desulfuration, which is carried out at nearly ambient temperature with conventional limestone-gypsum FGD processes, and then heated up again to the MCFC operating temperature, requiring additional equipment and thermodynamic losses; (iii) a relevant heat input from a different clean fuel like natural gas should be considered to operate the MCFC. For these reasons, this concept seems better suited for application on gasification-based plants, where it is easier to obtain a deep purification of the syngas, which can be also used to feed both the MCFC.

The aim of the present work is the detailed investigation of an IGFC plant, where part of the coal-syngas is used to feed a MCFC, which uses the GT flue gas as cathode feed. Two solutions, based on two and three MCFC stacks, are proposed to optimize the CO_2 capture section. Results are discussed by considering efficiency, energy balances and preliminary technical and economical remarks.

2. PLANT CONFIGURATION

The configuration of the proposed plant is shown in Fig.2. Gasification is based on a dry-feed, oxygen-blown, entrained flow Shell type gasifier operating at 44 bar and 1550°C, which allows for high carbon conversion and cold gas efficiency. It is a slagging gasifier with membrane walls cooled with 54 bar evaporating water and insulated by the slag layer which minimizes heat losses and heat flux towards the membrane walls. A low sulfur South African coal (64.44% C, 3.95% H, 7.40% O, 1.49% N, 0.85% S, 9.20% H₂O, 12.67% Ash; 24.62 MJ/kg LHV) is used, pulverized and dried before feeding with a stream of air, heated by means of saturated water from the HP steam drum. Coal is loaded by means of lock-hoppers where CO_2 is used as pressurizing gas (stream 34) to reduce dilution of the final CO_2 stream with N_2 . CO_2 -based stream used for coal loading is extracted from the final CO₂ compression train, at a proper pressure before the final pump. In order to assure easier start-up and higher flexibility, oxygen is produced in a



Figure 2 – Schematic of the proposed MCFC-based IGFC plant.

stand alone ASU, generating a 95% purity oxygen flow. A pumped oxygen plant with a consumption of 325 kWh per metric ton of oxygen produced is used, according to data reported by Air Products [22]. Nitrogen produced in the ASU is largely used for syngas dilution for NO_x control (stream 23), the remaining vented to the atmosphere. Syngas produced is quenched to 900°C by means of cold (200°C) syngas recirculation (stream 12). Molten slag entrained by the gas stream solidifies and syngas is then cooled down in a convective syngas cooler, generating HP steam and heating up HP feedwater. Most of the ash entrained in the raw syngas leaving the syngas cooler are removed in a high pressure, high temperature filter [23]. Syngas is then partly recirculated for quench, partly sent to a scrubber, where remaining solids and soluble contaminants are removed. Liquid water from the scrubber is clarified in a sour water stripper by means of LP steam and subsequently recycled back to the scrubber. Syngas exits the scrubber at about 130°C (stream 14) and is heated up to 180°C by means of water from the HP drum and sent to a catalytic bed for COS hydrolysis. Here COS is converted to H₂S, allowing for a deep sulfur removal in the subsequent acid gas removal (AGR) section. H₂S is removed in the AGR section by means of a MDEA-based system and sent to the sulfur recovery unit. No net steam output is assumed to be obtained in the sulfur recovery unit, i.e. the steam raised by H₂S combustion in the Claus plant is balanced by the heat required to keep S molten and to regenerate the SCOT solvent. AGR section was not simulated and assumptions for its calculation were obtained from [23-25].

Clean syngas is preheated and humidified in a saturator using low temperature heat. Saturation with water helps to reducing both NO_x production in the GT combustor and the risk of carbon deposition in the MCFC. Saturated syngas is then further heated up to 330°C by a closed loop of pressurized water transferring heat from syngas coolers. This solution prevents from the risk of sulfur contamination of the clean syngas, which is not in direct contact with the raw syngas stream, in case of a heat exchanger failure. Most of the syngas is then mixed with nitrogen and burned in the gas turbine combustor (stream 24). The remaining part of the syngas is expanded, reheated to 330°C and sent to the atmospheric MCFC module (described in more detail in the next paragraph) together with the gas turbine exhausts. During syngas preheating, a sulfur guard should be introduced to reduce sulfur compounds below the concentration tolerated by the MCFC. Further syngas cleanup processes could be introduced to remove other traces of potentially harmful compounds (e.g. HF, HCl, NH₃, Hg).

Cathode gas (stream 5) is then cooled down by producing steam for the steam cycle, while anode exhaust is sent to the CO_2 purification and compression unit after cooling and heat recovery (stream 21). Following the approach proposed in [19], CO_2 is purified from incondensable gas by means of a cryogenic process based on a two-step flash separation [21]. In this unit, while obtaining a high purity CO_2 stream suitable for final liquefaction and permanent disposal, most of the CO and H₂, which left the fuel cell unconverted, are recovered at high pressure and burned in the GT combustor (stream 22).

The heat recovery steam cycle is based on a single pressure level with reheat configuration (just a small additional amount of 54 bar steam is evaporated in the gasifier membrane walls). Lower evaporation levels are not required for an efficient heat recovery, since low temperature heat is used for water preheating. A larger amount of high pressure steam is in fact generated in this plant with respect to a conventional NGCC because of the high temperature heat available from syngas coolers and MCFC intercoolers. In particular, 27-44% of the total high temperature heat (i.e. suitable for the production of high pressure steam) to the steam cycle is recovered from MCFC cooling, 28-25% from the syngas coolers and the remaining 45-31% from the FC cathode exhausts.

2.1. The MCFC module

The detail of the MCFC module layout is shown in Fig.3. Inlet flows are the cleaned and preheated syngas (stream 19) and the GT flue gas (stream 4), which provides O_2 and CO_2 to the cathode. FC exhausts are the CO_2/H_2O -rich stream (20) sent to the CO_2 purification unit and the O_2 and CO_2 depleted GT flue gas (stream 5) sent to the stack after heat recovery.

The following limitations were considered to define the working conditions of the MCFC module:

CO₂ utilization factor at the cathode (U_{CO2} , defined as the ratio between the flow rate of CO₂ transferred through the cell as carbonate CO₃⁼ ions and the CO₂ flow rate introduced at the cathode inlet) is limited by the minimum allowable CO₂ concentration of 1.5%vol.;



Figure 3 – Schematic of the MCFC module.

• the allowable operating temperature range in the fuel cells is 580-665°C.

Considering the above limitations, configurations with two and three (as in Fig.3) FC stacks in series were investigated. If only one FC module were used, very low CCR would result. In fact, if the maximum CO₂ capture potential of the MCFC from GT flue gas is aimed (corresponding to the maximum allowable U_{CO2}), the corresponding fuel consumption (for each mole of CO_2 separated, 0.5 moles of O_2 are transported to the anode for fuel oxidation) would lead to unacceptable temperature increases along the fuel cell. For this reason, cases with two MCFC in series (leading to CCR of almost 60%) and three MCFC (CCR above 90%) with intermediate cooling and HP steam production (stream 36) were considered. Such a configuration is not required when using NG as fuel and MCFC with internal reforming, since a large amount of heat is converted into chemical energy by the endothermic steam methane reforming reaction [18,19].

Before entering the fuel cell, syngas is preheated to the minimum allowable temperature of 580°C, by mixing with part of the hot anode exhausts. Considering the high temperature, recycle is carried out by means of an ejector, driven by the higher pressure stream from syngas expander (stream 19).

Carbon deposition conditions in the FC were also verified by chemical equilibrium calculations. Thanks to the syngas saturation and the anode recycle, steam to carbon ratio at FC inlet is high enough to avoid carbon formation and further dilution with steam is consequently not necessary.

Fuel and CO_2 utilization factors determine the repartition of the syngas fuel between the gas turbine and the MCFC. Fuel utilization factor also influences CO_2 and H_2O concentration in the anode exhausts, which increase with U_f . Higher U_f hence lead to an easier CO_2 liquefaction and to a better humidification of the fuel at MCFC inlet after mixing with the recycle.

3. CALCULATION TOOLS AND ASSUMPTIONS

The thermodynamic model of the assessed power cycles is carried out with the modular simulation code "GS", a tool developed since several years at the Energy Department of Politecnico di Milano, which has proved to yield highly accurate results in a variety of complex plant configurations, including all kinds of gas turbine cycles, combined cycles and hybrid cycles [26-33]. The code integrates models for the prediction of steam turbines [26], cooled gas turbines [27] and FC performance [30-33].

Main assumptions for plant calculation are reported in Tab.1. The MCFC is modeled through a lumped volume approach, which calculates energy balances, thermodynamic properties and chemical composition of anode and cathode outlet gases as a function of reactant utilization factors (U_f , U_{O2} and U_{CO2}), inlet compositions and voltage, with no attempt of predicting the fuel cell temperature profiles. Water gas shift reaction at the anode is calculated with the hypothesis of chemical equilibrium. Overall fuel utilization factor in the MCFC module was assumed equal to 85%. Fuel cell potential

is calculated as a function of the reversible Nernst potential and the current density as described in [18]. A current density of 1000 A/m^2 was assumed and potentials of 0.71 V (for the first stack operating with higher H₂ and CO₂ fractions), 0.67 V (second stack) and 0.57 V (third stack, if present) were obtained. The rather low voltage of the last stack is a result of the model voltage prediction, given the assumption of a constant current density; in real stacks, different operating conditions with lower current density and a higher voltage could be selected, leading to higher FC active area and cost but also to a higher plant efficiency. For this reasons, further investigation should be carried out about the unconventional operating conditions of the MCFCs, especially in the threestack configuration.

Pressure at syngas expander outlet was determined to have the pressure required for the primary flow to drive the anode ejector. This pressure was calculated by solving momentum and energy balance equations [34-36] for a constant area mixing channel and assuming nozzle and diffuser isentropic efficiencies of 97% and 50% respectively. The value of diffuser isentropic efficiency takes into account friction losses occurring during the mixing process and was calibrated on the specifications of an industrial product [37].

The cooled gas turbine model [27] was calibrated to reproduce the advanced industrial gas turbine Siemens SGT5-4000F (TIT = 1335°C, β = 17), under natural gas firing conditions. The selected turbine also defines the actual size of the reference IGCC without CO₂ capture used as a reference [7], which results in a thermal input of 950 MW (coal LHV), also used for the FC-based plant configurations. The flow rate of N₂ used for syngas dilution was varied to obtain a stoichiometric flame temperature of 2220 K, like in the reference IGCC, which should be sufficiently low to have acceptable NO_x emissions with a diffusive flame [38].

4. RESULTS

The energy balance and the main operating parameters of the assessed IGFC plant are shown in Tab.2 and compared with a reference IGCC without CO_2 capture. The characteristics of the main streams are also reported in Tab.3 referred to the case with three intercooled MCFC stacks.

Efficiencies of 47.1 and 46.0% and corresponding efficiency penalty of 0.12-1.25% points, with CCR of 58.8 and 91.4% were obtained for the cases with 2 and 3 MCFC stacks respectively. A measure of the energy cost related to CO_2 capture is given by the Specific Primary Energy Consumption for CO_2 Avoided (SPECCA), where efficiency and specific emission of the assessed plant are compared in a single index with those of a reference technology. SPECCA is defined as:

$$SPECCA = \frac{HR - HR_{REF}}{e_{REF} - e} = \frac{3600 \cdot (1/\eta - 1/\eta_{REF})}{e_{REF} - e}$$
(1)

where HR is the heat rate of the plants, expressed in kJ_{LHV}/kWh_{el} and e is the CO₂ specific emission, expressed in

Table 1 – Main assumptions for plants calculation

Gasifier		Auxiliaries
Gasification pressure, bar	44.0	Pulverisers and coal handling, kJe/kgcoal
Gasification temperature, °C	1550	Slag handling, kJ _e /kg _{ash}
Heat losses in gasifier, % of input LHV	0.7	Miscellaneous BOP, % of input LHV
Steam for coal drying, kJ per kg of evaporated H ₂ O	2950	Syngas expander
H_2O in coal after drying, % wt.	2	Polytropic efficiency, %
Carbon conversion, %	99.0	Electrical/mechanical efficiency, %
Moderator steam, kg _{H20} /kg _{coal}	0.06	MCFC
Moderator steam pressure, bar	48	Maximum outlet temperature, °C
Oxygen pressure, bar	48	Maximum gas temperature increase, °C
Temperature of O_2 to gasifier, °C	15	Current density, A/m^2
Heat to membrane walls, % of input coal LHV	2	Overall fuel utilization factor, %
Lock hoppers CO ₂ pressure	88	Minimum CO ₂ concentration at anode, % vol.
Lock hoppers CO_2 temperature, °C	80	Heat loss, % of fuel LHV
Lock hoppers CO ₂ to dry coal ratio, l/kg _{dry-coal}	2.6	Air and fuel side pressure losses, %
ASU		Pressure loss in intercoolers, %
Oxygen purity, % mol.	95	Electrical efficiency, %
Pressure of delivered oxygen, bar	48	Ejectors
Pressure of delivered nitrogen, bar	1.2	Nozzle isentropic efficiency, %
Temperature of delivered O_2 and N_2 , °C	15	Diffuser isentropic efficiency, %
Electric consumption, kWh/t_{O2}	325	Gas turbines and combustor
Syngas quench		Combustor pressure loss, %
Quenched syngas temperature, °C	900	Compressor polytropic efficiency ^a , %
Cold recycled syngas temperature, °C	200	Turbine polytr. efficiency cooled/uncooled sta
Recycle compressor polytropic efficiency, %	75	Stoichiometric flame temperature, K
Recycle compr. electrical/mechanical efficiency, %	92	Mechanical loss of compressor/turbine, %
Heat exchangers		Gas turbine auxiliaries, % of power output
Pinch point ΔT in gas - water heat exchangers, °C	10	Electric generator efficiency, %
Pinch point ΔT in syngas coolers, °C	20	Steam cycle
Sub-cooling ΔT in evaporators, °C	5	Pressure levels, bar
Pressure losses in SH/RH, %	8	HP/IP live steam temperature, °C
Pressure losses in economizers, bar	30	Condensing pressure, bar
Gas side pressure loss on MCFC flue gas heat recovery, %	3	Power for heat rejection, MJ _e /MJ _{th}
Overall pressure losses between gasifier and saturator, %	15	Turbine mechanical efficiency ^b , %
Heat losses, % of heat transferred	0.7	Electric generator efficiency, %
Bulk sulfur removal (MDEA process)		CO ₂ compression and conditioning
Temperature of absorption tower, °C	35	\dot{CO}_2 delivery pressure, bar
Syngas pressure loss, %	1	Number of compression stages before/after pu
Steam consumption (net of Claus plant),		Inter-cooling temperature, °C
MJ of LP steam per kg of H ₂ S removed	16	Inter-coolers pressure loss, %
Sulfur removal and recovery auxiliaries, MJ _e /kgH ₂ S	1	Compressors isentropic / mechelectric efficie
Sour water stripper		Pump hydraulic / mechelectric efficiency, %
Steam consumption, kJ _{LP steam} / MJ inlet coal LHV	12	Minimum ΔT in cryogenic heat exchanger, °C
· · · · · · · · · · · · · · · · · · ·		Incondensable gas separation temperature, °C

Auxiliaries	
Pulverisers and coal handling, kJe/kgcoal	50
Slag handling, kJe/kgash	100
Miscellaneous BOP, % of input LHV	0.15
Syngas expander	
Polytropic efficiency, %	88
Electrical/mechanical efficiency, %	98.7
MCFC	
Maximum outlet temperature, °C	665
Maximum gas temperature increase, °C	85
Current density, A/m^2	1000
Overall fuel utilization factor, %	85
Minimum CO_2 concentration at anode, % vol.	1.5
Heat loss, % of fuel LHV	2
Air and fuel side pressure losses, %	2
Pressure loss in intercoolers, %	1
Electrical efficiency, %	97
Ejectors	
Nozzle isentropic efficiency, %	97
Diffuser isentropic efficiency, %	50
Gas turbines and combustor	
Combustor pressure loss, %	3.0
Compressor polytropic efficiency ^a , %	92.2
Turbine polytr. efficiency cooled/uncooled stages ^a , %	93.3/93.5
Stoichiometric flame temperature, K	2220
Mechanical loss of compressor/turbine, %	0.135
Gas turbine auxiliaries, % of power output	0.35
Electric generator efficiency, %	98.7
Steam cycle	
Pressure levels, bar	130/54
HP/IP live steam temperature, °C	565
Condensing pressure, bar	0.04
Power for heat rejection, MJ _e /MJ _{th}	0.01
Turbine mechanical efficiency ^b , %	99.5
Electric generator efficiency, %	98.7
CO ₂ compression and conditioning	
CO_2 delivery pressure, bar	150
Number of compression stages before/after purification	3/2
Inter-cooling temperature, °C	30
Inter-coolers pressure loss, %	2
Compressors isentropic / mechelectric efficiency, %	82/94
Pump hydraulic / mechelectric efficiency, %	75/90

^a Applies to large size stages at the optimum specific speed. Actual efficiencies are calculated according to the model reported in [27].

^b Isentropic efficiency of the steam turbine is calculated, stage-by-stage, according to the method described in [26].

kg_{CO2}/kWh_{el}. The SPECCA obtained here is equal to 0.05 and 0.31 MJ/kg_{CO2}. As a comparison, a value of about 2.8 MJ/kg_{CO2} can be considered for an IGCC with pre-combustion CO₂ absorption with 90% of CO2 captured and 9% points of efficiency penalty [23,24].

It is interesting to discuss the results obtained for this IGFC plant in comparison with those reported in [18-20] for the NG-fed plants:

Considering the limit of the minimum allowable concentration of CO₂ at the anode (assumed equal to 1.5% vol.), a higher CCR can be obtained in the IGFC

plants. This results from the higher CO₂ concentration in the GT flue gas at the FC cathode inlet and to the consequently higher CO₂ flow rate which can permeate through the MCFC electrolyte.

The efficiency penalty obtained for the IGFC cases is similar than the penalty of the NG-fed case when compared to the reference plants without CO₂ capture, but obtained with a lower current density (1000 A/m^2) vs. 1500 A/m²). Therefore, a 50% higher MCFC active surface is needed in this case to separate 1 mole of CO_2 to have similar performance. This results from the lower

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 H_2 content of the syngas, which leads to a lower MCFC polarization curve.

- In the IGFC plant, high CCR can only be reached by adopting an intercooled multi-stack configuration. The higher CO₂ specific production of coal together with the lower voltages, lead to higher temperature increase in the fuel cell for a given CO₂ utilization factor. In addition, the heat sink of the steam reforming reaction cannot be exploited with syngas (despite a methanation island is introduced, with an increase of plant cost and complexity).
- In the NG-fed plant, the MCFC-based active separator can be considered for the retrofitting of an existing combined cycle, with an increase of the net power output due to the CF introduction. On the contrary, in the IGFC plant, a retrofitting option is not as easy because a larger gasification and syngas treating island would be required to produce the fuel for the MCFC.

This and the previous two points could be overcome by using natural gas as MCFC fuel.

Like in the NGCC+MCFC plants, a favorable power share between the components is obtained in the IGFC cases, where the MCFC produces 18-26% of the gross power output. Most of the power is hence again produced by the gas turbine and the steam cycle, which are the lower specific cost components in terms of ϵ/kW , with beneficial effects for the expected investment costs.

The issue of reducing the MCFC power share is relevant also when considering the large expected footprint and the bulky volume requirements of the fuel cell section of the power plant. A first gross estimation of the volume required by the vessels hosting MCFC stacks and related equipments, considering the first configuration (case with groups of 2 stacks), would result in about 8000 m³ assuming a conservative ratio of 1 m3 for each 20 m² of active surface (resulting here in

Table 2 – Power balance and	operating parameters	of the assessed	plants
	operating parameters		P

	IC	IGCC	
	2 MCFC stack		
FCs single passage fuel utilization, %	31.1-55.7	22.1-32.6-41.1	-
Overall fuel utilization, %	84.5	84.5	-
FCs O ₂ utilization, %	10.8-14.9	13.6-18.1-18.8	-
Overall O ₂ utilization, %	24.1	42.6	-
FCs CO ₂ utilization, %	22.2-35.3	27.9-44.5-68.4	-
Overall CO ₂ utilization, %	49.7	87.4	-
O_2/CO_2 concentration at last FC cathode outlet, % mol.	8.6/5.5	7.0/1.5	-
Anode recirculation, % of MCFC inlet flow	78.0	79.1	-
MCFC potential, mV	714-664	712-670-570	-
Total active MCFC surface, m ²	159 355	227 905	-
Fraction of syngas to the MCFC, %	36.7	52.5	-
Flow at GT compressor inlet, kg/s	365.0	293.7	546.0
Electric power, MW _e			
MCFCs	106.1	143.8	-
Gas turbine	232.2	184.6	328.6
Steam turbine	213.0	217.2	207.2
Syngas expander	11.50	15.22	-
Air separation unit (ASU)	-37.61	-37.61	-36.17
Dilution N ₂ compressor	-34.79	-29.90	-34.66 ^a
Compression of CO_2/N_2 for lock-hoppers	-1.79	-1.79	-4.56
Syngas cooling recycle fan	-0.82	-0.82	-1.20
Steam cycle pumps	-4.28	-3.97	-3.44
CO_2 compression and liquefaction	-28.94	-42.72	-
Auxiliaries for heat rejection	-2.86	-2.98	-2.99
Pulverizers and coal handling	-1.79	-1.79	-1.79
Slag handling	-0.49	-0.49	-0.49
Auxiliaries for sulfur recovery	-0.35	-0.35	-0.35
Miscellaneous BOP	-1.42	-1.42	-1.42
Gross power output, MW _e	562.8	560.8	535.8
Net power output, MW _e	447.6	437.0	448.7
Fuel input LHV, MW _{th}	950.0	950.0	950.0
CGE, %	76.75	76.75	78.07
Carbon capture ratio, %	58.79	91.44	0
Net LHV efficiency, %	47.12	45.99	47.24
SPECCA, MJ/kg _{CO2}	0.05	0.31	-
Efficiency penalty, % points	0.12	1.25	-
Specific CO ₂ emission, g/kWh	298.5	63.5	723.3

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	T,	р,	G,	Molar composition, %							
point	°C	bar	kg/s	Ar	CO	CO_2	H ₂	H ₂ O	N ₂	O ₂	H_2S
1	15.0	1.01	293.7	0.92	0.00	0.03	0.00	1.03	77.28	20.73	0.00
2	407.6	17.00	293.7	0.92	0.00	0.03	0.00	1.03	77.28	20.73	0.00
3	1393	16.54	328.0	0.92	0.00	12.29	0.00	5.94	72.29	8.56	0.00
4	609.5	1.09	389.9	0.92	0.00	10.29	0.00	5.14	73.11	10.55	0.00
4a	665.0	1.07	367.2	0.97	0.00	7.75	0.00	5.37	76.40	9.52	0.00
4b	665.0	1.05	341.1	1.02	0.00	4.53	0.00	5.66	80.57	8.22	0.00
5	665.0	1.04	319.0	1.07	0.00	1.50	0.00	5.94	84.50	7.00	0.00
6	69.7	1.01	319.0	1.07	0.00	1.50	0.00	5.94	84.50	7.00	0.00
7	15.0	1.01	142.1	0.92	0.00	0.03	0.00	1.03	77.28	20.73	0.00
8	15.0	44.00	35.75	Dr	y coal (%w	vt.: 69.5 C,	0.9 S, 1.6 I	N, 4.3 H, 8	.0 O, 2.0 H	20, 13.7 A	sh)
9	15.0	48.00	34.04	3.09	0.00	0.00	0.00	0.00	1.91	95.00	0.00
10	300.0	54.00	2.14	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
11	900.0	44.00	167.7	0.99	61.62	9.15	16.92	9.80	1.22	0.00	0.30
12	206.8	44.44	85.59	0.99	61.62	9.15	16.92	9.80	1.22	0.00	0.30
13	200.0	42.68	82.10	0.99	61.62	9.15	16.92	9.80	1.22	0.00	0.30
14	155.2	41.83	84.39	0.96	59.40	8.82	16.30	13.06	1.18	0.00	0.29
15	180.0	41.00	84.39	0.96	59.40	8.82	16.30	13.06	1.18	0.00	0.29
16	35.0	38.57	75.85	1.10	68.45	10.16	18.79	0.15	1.36	0.00	0.00
17	153.7	37.42	84.74	0.95	58.92	8.75	16.17	14.05	1.17	0.00	0.00
18	330.0	37.04	84.74	0.95	58.92	8.75	16.17	14.05	1.17	0.00	0.00
19	330.0	3.33	44.45	0.95	58.92	8.75	16.17	14.05	1.17	0.00	0.00
19a	590.0	1.09	213.2	0.69	20.88	55.63	6.23	15.72	0.85	0.00	0.00
19b	580.0	1.07	235.9	0.65	13.86	63.98	6.07	14.64	0.80	0.00	0.00
19c	580.0	1.05	261.9	0.61	8.85	70.34	3.77	15.68	0.75	0.00	0.00
20	665.0	1.04	115.3	0.58	4.99	75.22	2.07	16.42	0.71	0.00	0.00
21	76.1	1.02	115.3	0.58	4.99	75.22	2.07	16.42	0.71	0.00	0.00
22	330.0	22.54	8.26	3.92	41.91	28.14	20.19	0.00	5.84	0.00	0.00
23	553.8	20.47	47.61	0.00	0.00	0.00	0.00	0.00	100.00	0.00	0.00
24	429.8	20.47	96.16	0.75	30.35	6.28	9.03	6.43	47.16	0.00	0.00
25	29.0	1.50	123.0	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
26	180.0	156.0	46.12	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
27	565.0	119.0	46.69	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
28	565.0	119.0	152.4	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
29	228.9	30.00	98.97	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
30	229.5	54.00	9.89	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
31	174.6	1.50	7.32	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
32	324.0	130.0	6.85	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
33	29.76	110.0	82.96	0.24	1.20	98.20	0.18	0.00	0.18	0.00	0.00
34	30.0	88.00	15.30	0.24	1.20	98.20	0.18	0.00	0.18	0.00	0.00
35	325.0	130.0	98.71	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
36	335.0	130.0	98.71	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00
37	300.0	54.00	7.74	0.00	0.00	0.00	0.00	100	0.00	0.00	0.00

Table 3 – Temperature, pressure, flow rate and composition of the streams of the plant in Fig.2,3, with 3 intercooled MCFC stacks.

about 14 kW/m³ of overall volumetric power density) [39].

It also interesting to compare these results with those obtained with SOFC-based IGFC cycles already discussed in previous works. By the point of view of efficiency, the SOFC-based cycles yield similar results (47.1-47.6%) when considering the most comparable configurations (without methanation process and with anode exhaust oxycombustion [8]), but with higher CO₂ capture rate (~97.5%) and a much worse power share among the components (SOFC generates about 60% of the gross power). On the other hand, the more complex configurations with methanation and cathode exhaust hydrogen post-firing considered in a recent work [9] reach

51.6% efficiency, still with a less favorable power distribution (53% of the gross power is generated by the SOFC).

5. CONCLUSIONS

This work presented the thermodynamic assessment of an IGFC where an MCFC is used as an active CO_2 separator. Two cases utilizing two and three intercooled FC stacks were calculated. Detailed energy and mass balances of the assessed cases are presented and compared with a reference IGCC without capture and with a NG-fired plant with CO_2 capture, exploiting the same MCFC-based post-combustion capture concept. Results show that the proposed IGFC cycles reach efficiencies of 46.0-47.1%, 0.1-1.25% points less than the

reference IGCC cycle, while achieving 58-91% lower specific CO₂ emissions.

The contribution of the MCFC on the plant power balance is between 18 and 26% of the gross power, confirming the favorable power share obtainable with this concept, where most of the power is generated by the lower cost components (gas and steam turbine). For this reason, despite the lower efficiency and the higher emission obtained with respect to other concepts where the FC (MCFC or SOFC) produces most of the gross power, such a concept can be competitive when plant economics and the cost of CO₂ avoided are considered.

Further investigations are however required to verify the unconventional operating conditions of the MCFC, especially when considering: (i) the operation with a CO-rich fuel, which leads to increased risk of carbon deposition; (ii) the operability of the MCFC module, with 2 or 3 stack in series and (iii) the uncommonly low voltage of the last FC in the 3-stack configuration. Such issues, together with economic assessments, will be the subject of future works.

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