Techno-economic analysis of a natural gas combined cycle integrated a Ca/Cu looping process for low CO₂ emission power production

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Abstract

A techno-economic analysis of a natural gas combined cycle integrated with a pre-combustion CO₂ capture process based on the Ca/Cu process is carried out. A throughout calculation of the balances of the whole power plant has been done, including the results obtained from a 1-D pseudo homogenous model for the fixed bed reactors that compose the Ca/Cu process. Moreover, a methodology developed by the authors is here presented for calculating the cost of the electricity produced and of the CO₂ avoided. This methodology has been used for performing the economic analysis of the Ca/Cu based power plant and to optimize Ca/Cu reactors size and pressure drops in critical heat exchangers. An electricity cost of 82.6 €/MWh has been obtained for the Ca/Cu based power plant, which is 2.2 €/MWh below the benchmark power plant based on Auto Thermal Reformer and MDEA absorption process for CO₂ capture. The improved performance of the Ca/Cu based power plant in terms of electric efficiency and reduced capital cost expenditure are the reasons for such reduced electricity costs. Moreover, a lower cost of CO₂ avoided is also obtained for the Ca/Cu plant with respect to the benchmark (80.75 €/t_{CO2} vs. 85.38 €/t_{CO2}), which features 89% of CO₂ capture efficiency.

Keywords: hydrogen, pre-combustion CO₂ capture, sorption enhanced reforming, chemical looping, combined cycle, economic analysis

Notation

E_{CO_2}	Specific CO ₂ emissions of the power plant with CO ₂ capture [kg _{CO2} /MWh]
$E_{CO_2,ref}$ [kg _{CO2} /MWh]	Specific CO ₂ emissions of the reference power plant without CO ₂ capture

Greek symbols

η	Electric efficiency
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Acronyms

ATR	Autothermal reforming
CEPCI	Chemical Engineering Plant Cost Index
CCA	Cost of CO ₂ avoided [€/t _{CO2}]
CCR	CO ₂ capture ratio [%]
ECO	Economizer
EVA	Evaporator
GT	Gas turbine
HP	High pressure
HRSG	Heat recovery steam generator
IF	Installation factor
LCOE	Levelized cost of electricity [€/MWh]
LP	Low pressure
NGCC	Natural gas combined cycle
MDEA	Methyldietanolamine
O&M	Operation and maintenance
RH	Reheater
SH	Superheater
SMR	Steam methane reforming
SP	scaling parameter
SPECCA	Specific primary energy consumption [MJ _{LHV} /kg _{CO2}]
TPC	Total plant cost
TDPC	Total direct plant cost
WGS	Water gas shift

1. Introduction

Contribution of the electricity production sector to the global CO₂ emissions is noticeable, accounting for around 14 Gt_{CO₂} in 2015 (i.e. around 42% of worldwide CO₂ emissions that year) (IEA, 2017). Coal represents the main primary energy source for electricity production worldwide, covering almost two-thirds of the electricity share in countries like China, Australia, South Africa or Poland. However, looking into the evolution of the coal and natural gas shares in the electricity production sector, it has been noticed a switch from coal to natural gas in the last years due to the lower CO₂ intensity and the decreasing cost trend of such fossil fuel due to the availability of shale gas reservoirs (IEA, 2017). Natural gas combined cycle (NGCC) is the state-of-the-art technology for power production from natural gas, characterised by high electric efficiency (more than 60% with state-of-the-art gas turbines), reduced environmental impact (thanks to the low-carbon fuel used and the improved efficiency, the low NO_x burners and zero SO_x and dust emissions), high flexibility and reduced investment cost with respect to coal-fired plants. On the other hand, carbon capture and storage (CCS) remains as the only option for drastically reducing CO₂ emissions to the atmosphere from fossil fuel power plants and energy intensive industry and achieve the ambitious goal of stabilizing global temperature increase to well below 2°C, as target of the recent Paris Agreement.

A pre-combustion CO₂ capture route is assessed in this work, where natural gas is converted into a hydrogen rich gas that feeds a NGCC to produce electricity with reduced CO₂ emissions. This pre-combustion route has some advantages with respect to the well-established post-combustion CO₂ capture route based on absorption with amines, namely the possibility of hydrogen co-production and the avoidance of emitting toxic products from amine degradation. A number of studies have assessed the natural gas-fired auto thermal reformer (ATR) based power plant with pre-combustion CO₂ capture, which is considered the benchmark technology for power generation with pre-combustion CO₂ capture. Different process conditions regarding the ATR operating temperature, the steam-to-carbon ratio (S/C) and the type of CO₂ capture process have been considered for these studies (Romano et al., 2010). Electric efficiencies range from 42% (or efficiency penalties of around 14%-points with respect to benchmark NGCC without CCS), to 49-51% (or about 7.5%-points of efficiency penalty) (Corradetti and Desideri, 2005; Ertesvåg et al., 2005; Fluor Daniel Inc., 2000; Lozza and Chiesa, 2002; Nord et al., 2009; Romano et al., 2010; Stork Engineering Consultancy, 2000). An air-blown ATR is the configuration most-widely considered in the reviewed literature, with the exception of the work by Stork Engineering Consultancy (2000) that considered an oxygen-blown ATR. Typically, chemical absorption with Methyldietanolamine (MDEA) has been the preferred option for CO₂ separation from the hydrogen-rich gas, except for the case where very high pressures or O₂-blown conditions were considered for the ATR, where physical separation through Selexol was chosen.

An emerging pre-combustion CO₂ capture route based on the combination of the CaO/CaCO₃ and Cu/CuO chemical loops is analysed in this work. The characteristic of this process is that it allows producing a high-pressure hydrogen rich gas stream while carrying out inherent CO₂ capture through the sorption enhanced reforming (SER) process of natural gas with CaO as a sorbent. The CaCO₃ produced in this hydrogen production stage is regenerated in situ in the same solid bed by means of the exothermic reduction of CuO with a fuel gas containing H₂, CO and/or CH₄, resulting in this way in a highly intensified process with no need of high temperature heat exchange surfaces or high cost air separation unit for sustaining sorbent regeneration (Abanades et al., 2010; Abanades and Murillo, 2009). This process fully exploits its advantages when carried out in a series of fixed bed reactors operating in parallel at different pressure and temperature. This hydrogen production process can be potentially cheaper than the benchmark ATR-based hydrogen production with CO₂ capture, since the highly cost-effective reforming steps (i.e. heated pre-reformers and ATR) are no longer present, the syngas conditioning section is noticeably simplified as it is based on a high temperature CO₂ removal process, which turns out into a higher efficiency and reduced equipment than the low temperature chemical absorption technology.

Since its first description in 2009 (Abanades and Murillo, 2009), this process has experienced a significant progress in terms of materials development, reactor modelling and process performance analysis as recently reviewed in Fernández and Abanades, (2017a, 2017b). The functional materials needed for the correct functioning of this process have been successfully synthesized and tested under conditions relevant to the Ca/Cu process (i.e. high loads of Cu and CaO, materials in pellet form and pressure/temperature conditions similar to those present at larger scale) (García-Lario et al., 2013; Grasa et al., 2017; Ridha et al., 2015). Moreover, significant progress has been made in the development of fixed bed reactor models operating in the Ca/Cu process (Alarcón et al., 2017; Fernandez et al., 2014, 2012; Martini et al., 2016), which have been useful for defining the reactors operating sequence and the process parameters to improve the process efficiency and reduce the CO₂ emissions (Fernández and Abanades, 2017c; Martini et al., 2017). Due to its inherent characteristics, the Ca/Cu looping process can be designed for integration in hydrogen production plants, reaching hydrogen production efficiencies as high as 77% with CO₂ capture efficiencies of around 95%, which are noticeably better than benchmark technology (Fernández and Abanades, 2017c; Martínez et al., 2014; Riva et al., 2018). Moreover, it has demonstrated a promising potential when integrated in ammonia production plants, allowing to reduce the specific primary energy consumption per unit of NH₃ produced by 14 % with respect to the commercial process (Martínez et al., 2017). This Ca/Cu process has been also studied as a decarbonisation process for blast furnace gas in steel mills in a process configuration based on atmospheric fluidized bed reactors (Fernández et al., 2017; Martínez et al., 2018). Regarding its application as a pre-combustion CO₂ capture route in NGCC-based

power production plants, Martínez et al. (2013) studied the overall performance of a stand-alone power production plant based on this technology, considering a simplified calculation behaviour for the Ca/Cu process reactors in agreement with Fernández et al. (2012a). Axial diffusion effects as well as reaction kinetics for the main reactions involved in each stage of the process were not included in this analysis, which have been latterly demonstrated to influence significantly the performance and have been useful for designing a more accurate operation strategy and operational window for the Ca/Cu process (Fernández and Abanades, 2017c; Martini et al., 2017).

In this work a techno-economic analysis of a NGCC integrated with a pre-combustion CO₂ capture process based on the Ca/Cu process is carried out. Mass and energy balances of the whole power plant have been calculated, based on the results of a 1-D pseudo homogenous model for the Ca/Cu fixed bed reactors. Performance results obtained for the proposed NGCC configuration have been assessed and compared with a benchmark NGCC power plant based on Auto Thermal Reforming and on MDEA solvent process for CO₂ absorption. An economic analysis is also performed for calculating the cost of the electricity and of the CO₂ avoided and for comparing the assessed Ca/Cu process with the benchmark. The developed economic model has also been used to optimise the size of the Ca/Cu reactors and the pressure drops in some critical heat exchangers, whose optimal design is strictly related to a Capex-Opex trade-off.

2. Process description and assumptions

The proposed integration of the Ca-Cu process in a full-scale combined cycle-based power plant is shown in Figure 1, which can be conceptually divided into four islands: (i) a natural gas treating section, (ii) a syngas production section based on the Ca/Cu process, (iii) a CO₂ island and (iv) a power island.

The main assumptions used for calculating the heat and mass balances of this power plant are given in Table A1 in the Appendix.

After desulphurisation, the higher hydrocarbons present in the NG need to be decomposed into CH_4 , H_2 and CO before being introduced into the fixed bed reactors of the Ca/Cu process, in order to avoid coke formation. A pre-reforming unit consisting of an adiabatic solid bed filled with a commercial Ni-based catalyst and operating at temperatures of 380-500°C is used. Desulphurised NG is split into two streams. The main NG fraction (stream #17 in Figure 1) is mixed with steam at 27 bar bled from the steam turbine (stream #18) in a steam-to-carbon (S/C) ratio of 5 and then heated up to 490°C to be sent to an adiabatic pre-reformer operating at 26.4 bar. The outlet pre-reformed gas (stream #21) is further heated to 575°C and fed to the reforming stage A' of the Ca/Cu process. The other NG fraction, which is ultimately introduced into the reduction/calcination stage C of the Ca/Cu process (stream #4), follows a similar route as the previous one but a lower S/C=1 is used in the adiabatic pre-reformer since the objective of such pre-reformer is decomposing the hydrocarbons while keeping a minimum amount of steam at the inlet of the reduction/calcination stage for avoiding carbon formation from CH_4 . After being pre-reformed (stream #8), the mixture is expanded and heated up to the reduction/calcination Ca/Cu stage feed temperature.

Concerning the choice of the pre-reforming unit placed before the reduction/calcination stage, a HP adiabatic pre-reformer has been considered for this study instead of a LP heated pre-reformer as done previously by Martini et al. (2017), in order to reduce the capital cost of the pre-reforming unit. Finally, it is worth mentioning that the possibility of having CaO hydration in the initial part of the solid bed during stages A' and C of the Ca/Cu process has been discarded. As discussed in previous works about process design of Ca-Cu technology (Fernández and Abanades, 2017c; Martínez et al., 2014; Martini et al., 2017), the combination of high steam partial pressure and high temperature at stage A' inlet may result in the hydration of the CaO present in the initial part of the solid bed, which may cause particle breakup and deactivation. However, as will be seen in the next section about Ca/Cu process description, the presence of stage A before stage A' results in a fraction of CaCO_3 in the initial part of the solid bed at the beginning of stage A', which allows carrying out this stage without CaO hydration.

2.2. *Hydrogen production island based on the Ca-Cu process*

The process scheme of the different stages considered for the Ca/Cu process is shown in Figure 2. The Ca/Cu process consists of a system of packed bed reactors operating at different temperatures and pressures, corresponding to the stages of sorption enhanced reforming (A and A'), Cu oxidation (B), bed heating (B') and CaCO_3 calcination/CuO reduction (C). The operating conditions of the Ca-Cu process are indicated in Table 1. These stages of the Ca-Cu process have been calculated with a 1-D pseudo-homogeneous reactor model, which has been described in a

previous work (Martini et al., 2016). This model calculates the performance of an adiabatic packed bed reactor, accounting for axial mass and heat dispersion as well as for reaction kinetics. No radial temperature and concentration profiles are considered in this model, which neglects also the mass and heat transfer resistances between gas and solid phases, resulting in a negligible temperature difference between both phases along the reactor. Specific information about the equations used in this model can be found in the works by Martini et al. (2017, 2016).

Figure 2 Simplified scheme of the Ca/Cu process configuration assessed in this work

Table 1 Assumptions used for calculating the reference case of the Ca/Cu process with the 1-D pseudo-homogeneous model

Functional materials	
CaO active content in the CO ₂ -sorbent [% wt.]	30
Cu active content in the Cu-based material [% wt.]	65
Ni active content in the reforming catalyst [% wt.]	18
Particle size (m)	0.01
Ca-based sorbent apparent density [kg/m ³]	1539
Cu-based carrier apparent density [kg/m ³]	2089
Ni-based catalyst apparent density [kg/m ³]	1866
Operating conditions Ca/Cu stages	
Inlet temperature of gas to stage A/A' [°C]	700/575
Inlet pressure of stage A' [bar]	25.2
S/C molar ratio at pre-reformer of stage A' inlet [-]	5
Inlet gas velocity of stage A [m/s]	0.58
Inlet gas velocity of stage A' [m/s]	0.44
Inlet temperature of gas to stage B [°C]	340
Inlet pressure of stage B	21
Outlet pressure of stage B [bar]	19
O ₂ fraction in the gas fed to stage B [% vol.]	3.0
Inlet gas velocity of stage B [m/s]	1.28
Inlet temperature of stage C [°C]	670

Inlet pressure of stage C [bar]	1.8
S/C molar ratio at pre-reformer of stage C inlet [-]	1
Inlet gas velocity of stage C [m/s]	1.47
Reactor dimensions	
Reactor diameter [m]	4.5
Reactor Length [m]	10
Number of reactors in stage A/A'	2/2
Number of reactors in stage B/B'	4/4
Number of reactors in stage C	2
Initial solid composition	
Total CaO [% wt]	38.4
Active CaO [% wt]	11.5
Cu [% wt]	28.1
Ni [% wt]	2.1
Al ₂ O ₃ [% wt]	31.4

At the beginning of the first stage of the Ca/Cu process (A and A'), solid bed contains the three functional materials in their reduced form, whose composition is given in Table 1 (Martini et al., 2017). The SER stage has been split into two intercooled stages (A and A') and operated at high pressure and high S/C in order to achieve high CO₂ capture efficiency approaching 90% (Martini et al., 2017). The key objective of this scheme is to remove in stage A the carbon that has not been captured during stage A', increasing in this way the carbon capture efficiency in the whole process. The feed gas from the adiabatic pre-reformer (stream #21 in Figure 1) is heated to 575°C and fed to stage A', whose initial conditions correspond to the final condition of stage A. Steam methane reforming (SMR), water gas shift (WGS) and carbonation reactions take place in this stage. The H₂-rich gas generated in stage A' is cooled down from an average temperature of around 785°C to 700°C before being introduced into stage A, where the carbon slipped from stage A' reacts with the CaO in the bed. As discussed by Martini et al. (2017), this intermediate gas cooling is needed to boost the CO₂ capture through the carbonation reaction in stage A, which is favored at low temperature.

Figure 3 and Figure 4 show the dynamic profiles along the bed for the axial temperature, CaO conversion (X_{CaO}) and gas molar fraction of CH₄, CO, CO₂ and H₂ during stages A and A' operation, respectively. CaO conversion X_{CaO} represented in Figure 3(b) has been calculated according to Eq. (1) as the ratio between the moles of active CaO that have been converted into CaCO₃ and the total moles of fully calcined active CaO. Therefore, X_{CaO} of 1 indicates that all the active CaO has been converted into CaCO₃.

$$X_{CaO} = \frac{n_{CaCO_3}}{n_{active\ CaO}} \quad (1)$$

At the beginning of stage A, the solid bed already contains a fraction of carbonated sorbent that results from the operation of the previous stage of the Ca/Cu process (i.e. the reduction/calcination). This CaCO₃ represents a conversion of 16.7% (molar basis) of the total

active CaO present in the solid bed at the beginning, and the Cu-based material and reforming catalysts are fully reduced. As the H₂-rich gas coming from stage A' is fed to this stage, the CaO is slowly converted into CaCO₃ due to low carbon content in the stream (average CO and CO₂ contents of 2.1 and 1.73% vol. respectively). A heat plateau is formed at around 800°C due to the exothermic carbonation reaction and the hydrogen fraction increases due to the CO₂ removal, as shown in Figure 3. Regarding stage A' (Figure 4), its initial temperature and solid conversion profiles correspond to those at the end of stage A. At the beginning of this stage, the initial part of the solid bed is already carbonated (36.2% of conversion of total active CaO in the solid bed). Therefore, the feed gas reacts according to SMR and WGS reactions (promoted by the presence of the Ni-based catalyst) until the equilibrium of both reactions is reached, which makes the temperature decrease to a lower value than the inlet temperature. When the gas reaches the part of the solid bed where active CaO is available (i.e. $X_{CaO} < 1$) the carbonation reaction starts to take place, the temperature increases due to the exothermicity of the reaction and the H₂ fraction increases due to the shifting of the equilibrium caused by the CO₂ removal. This stage A' ends when all the active CaO in the reactor is converted to CaCO₃.

Figure 3 Dynamic profiles of (a) axial temperature, (b) CaO conversion and (c)/(d) gas molar fraction inside the bed during stage A operation

Figure 4 Dynamic profiles of (a) axial temperature, (b) CaO conversion and (c)/(d) gas molar fraction inside the bed during stage A' operation

Once stage A' has finished, reaction step B starts, where diluted air at high pressure is fed into the solid bed reactor for oxidizing the Cu to CuO. An O₂-depleted gas at high pressure and high temperature is obtained at stage B outlet, which is split into two fractions as indicated in Figure 2. During stage B operation, solid bed maximum temperature must be controlled to avoid problems related to undesired Cu reactions, Cu melting problems and CaCO₃ calcination. A reasonable value for this maximum temperature is around 830-860°C when operating at 15-30 bar, which is controlled through the inlet gas temperature and the O₂ concentration of the stage B gas feed (J.R. Fernández et al., 2012; Martini et al., 2016), which are fixed to the values given in Table 1. Compressed air from the GT compressor (stream #37 in Figure 1) is therefore mixed with a recycled fraction of O₂-depleted gas exiting stage B after being cooled down (stream #44) to keep the O₂ content in stage B feed under such reduced value.

¡Error! No se encuentra el origen de la referencia. shows the dynamic profiles of temperature, CaO and Cu conversion and gas molar fraction of CO₂ and O₂ along the bed during the operation of stage B. Cu conversion X_{Cu} indicated in **¡Error! No se encuentra el origen de la referencia.**(b) has been defined according to eq.(2) as the ratio between the moles of Cu oxidized to CuO and the total moles of fully reduced Cu.

$$X_{Cu} = \frac{n_{CuO}}{n_{Cu}} \quad (2)$$

When O₂ reaches the part of the solid bed with unconverted Cu the temperature increases up to around 850°C due to the exothermic oxidation reaction, which causes the release of some CO₂ due to the undesired calcination of CaCO₃. However, a small fraction of CO₂ is always present in the feed gas thanks to the recycle of the N₂-rich gas exiting this step, which reacts with the CaO formed by calcination in step B. Due to the low temperature of the diluted air fed to stage B, the packed bed remains at a temperature of 340°C at the end of this stage, and therefore an intermediate heating stage between stages B and C needs to be included to allow stage C to initiate. Moreover, since the recycled O₂-depleted gas should be cooled down to be mixed with the compressed air, the solid bed is heated during stage B' by extracting heat from this recycled O₂-depleted gas (Figure 2).

Figure 5 Dynamic profiles of (a) axial temperature, (b) solid conversion and (c) gas molar fraction of CO₂ and O₂ inside the bed during stage B operation

Figure 6 shows the profiles inside the bed for this solid bed heating stage. As indicated in Figure 2, this stage is fed counter-currently to the rest of stages of the Ca/Cu process and therefore the temperature heat fronts within the solid bed move from the end to the beginning of the solid bed

(i.e. from the right to the left hand side of

Figure 6 (a). Part of the CO₂ present in the recycled stream (that has been released during stage B) reacts with the CaO formed in the previous step, reducing in this way the carbon loss caused by the oxidation reaction. At the end of this step, all the solid bed will be left at a constant temperature of around 695°C, which was the temperature of the fed gas to this stage, and almost fully carbonated. At stage B' outlet, average temperature of the gas is 545°C (stream #42 in Figure 1), which is too high to be mixed with the compressed air from the compressor. Therefore, a heat exchanger is introduced to cool down this recycled gas to 340°C before being mixed with the compressed air. Non-recycled O₂ at stage B outlet is directly sent to the GT combustor to be expanded (stream #45).

Figure 6 Dynamic profiles of (a) axial temperature, (b) solid conversion and (c) gas molar fraction of CO_2 inside the bed during stage B' operation

Last stage of the Ca/Cu process (stage C) consists in the calcination of the $CaCO_3$ formed during stage A and A' by means of the exothermic reduction of the CuO formed during stage B. A proper Cu/CaO molar ratio should be chosen so that the heat released by the exothermic reduction of CuO with CH_4 , CO and/or H_2 is enough for driving the endothermic calcination of the $CaCO_3$ (J.R. Fernández et al., 2012). To avoid calcination temperatures higher than 900-950°C that may affect CaO and Cu reactivity, operating pressure of stage C should be around atmospheric. Therefore, a depressurization stage followed by a rinse step for solid bed cleaning may be introduced between stages B' and C as already considered by Martínez et al. (2014). On the other hand, operating temperature during calcination should be as low as possible with the aim of minimizing the amount of Cu needed in the Ca-Cu process. Typical values proposed for this reduction/calcination step fall down in the range of 850-870°C, which is enough for ensuring complete $CaCO_3$ calcination and sufficiently fast CuO reduction kinetics under the conditions of this reduction/calcination stage (Díez-Martín et al., 2018; Fernández et al., 2016). As anticipated before, pre-reformed NG coming from a high pressure adiabatic pre-reformer is

fed to stage C at 670°C. The S/C ratio used for this pre-reforming is around 1, which is high enough for avoiding carbon formation during stage C.

Figure 7 shows the dynamic profiles of temperature, CaO and Cu conversion and molar fraction of CH_4 , CO , CO_2 and H_2 within the solid bed during this stage. As this stage proceeds and CuO has already been reduced to Cu in the initial part of the solid bed (X_{Cu} becomes 0), SMR, WGS and carbonation reaction occurs causing the carbonation of a small fraction of CaO in the initial part of the solid bed. The gas reaching the reaction front where there is unconverted CuO contains mainly H_2 and CO , and a Cu/CaO molar ratio of 2.1 has been chosen to reach a temperature in the bed of around 870°C needed for a sufficiently fast reaction rate. When the gas reaches the unreacted CuO , the reduction reaction to Cu happens, the temperature increases and the CaCO_3 is calcined (X_{CaO} decreases), resulting in CO_2 and H_2O as main reaction products. The reaction front is faster than the heat exchange front due to the increase in the flow rate caused by the CO_2 released by the sorbent, causing the formation of a heat plateau which remains in the solid bed after this stage is finished to the next one (see Figure 3 (a)). Gas at stage C outlet is basically CO_2 and H_2O (stream #11 in Figure 1), which is cooled down before being sent to CO_2 compressor.

Figure 7 Dynamic profiles of (a) axial temperature, (b) solid conversion and (c)/(d) gas molar fraction inside the bed during stage C operation

2.3. Power island and heat recovery

As shown in the process flowsheet in Figure 1, the power island in the proposed configuration consists of a combined cycle that uses the H_2 -rich gas produced in the Ca/Cu process as fuel in the combustor of the GT. Air needed in the oxidation stage B of the Ca/Cu process is split from the compressed air from the GT compressor and further compressed by a booster compressor (stream #36).

Flue gas from the gas turbine (stream #34) is cooled down in a heat recovery steam generator (HRSG) in order to produce superheated steam at 565°C and 123.4 bar that is sent to a steam turbine. It must be pointed out that most of the steam expanded in the steam turbine is not generated in the HRSG downstream the GT (here it is partly economized/heated and partly superheated/reheated), but from the N_2 -rich gas cooling in heat exchangers G and H shown in Figure 1. From the total amount of superheated steam sent to the high-pressure steam turbine (208 kg/s), almost 91% is produced in these heat exchangers (71.2 and 117.4 kg/s in heat exchangers G and H, respectively). As shown in Figure 8, which represents the composite curves

for all the hot gas streams available in the Ca/Cu process, 215 MW_{th} are recovered in the evaporators placed in the N₂-rich gas coolers G and H. A small amount of superheated steam (8.6 kg/s) is generated from the intermediate gas cooler between stages A' and A where the H₂-rich gas at 785°C (stream #23) is cooled down to 700°C before being introduced into stage A for recovering 17 MW_{th}. Finally, a fraction of the CO₂-rich gas from stage C (stream #12) is also used for producing a small amount of additional saturated steam at 130 bar and 331°C (stream #54) that is ultimately superheated within the HRSG. The remaining fraction of the CO₂-rich gas (stream #13) is used for pre-heating the NG fed into the Ca/Cu process in heat exchanger K of Figure 1.

Because of the large fraction of high temperature heat available in the chemical island (i.e. at temperature above the HP evaporation level), the steam cycle features a single evaporation pressure level and a series of four regenerative feedwater preheaters heated by steam bled from the steam turbine. Operating conditions for the deaerator, condenser and feedwater pump are given in Table A1 in the Appendix.

Figure 8 Temperature-Heat transferred curves for (a) H_2 -rich gas from stage A' (unit F in Figure 1); (b) H_2 -rich gas from stage A; (c) N_2 -rich gas from stage B' (unit H); (d) N_2 -rich gas to stage B (unit G); (e) CO_2 -rich gas from stage C (following stream #12 in Figure 1); (f) CO_2 -rich gas from stage C (following stream #13) and (g) HRSG of the GT flue gas

3. Description of the benchmark power production plant with CO₂ capture

A reference power production plant with CO₂ capture is considered for the sake of comparison. This plant, which reproduces the flowsheet and design criteria discussed by Romano et al. (2010), includes an ATR unit, a two-stage WGS section, a MDEA-based CO₂ absorption unit and combined cycle power plant (Figure 9).

Figure 9 Flowsheet of the benchmark NGCC with pre-combustion CO₂ capture by ATR and MDEA process

After being desulfurized and moisturized, NG is mixed with steam to get a S/C molar ratio of 2.5 and then heated up to 450°C to be processed in a heat exchange catalytic pre-reformer (stream #9 in Figure 9). Such pre-reformer allows an efficient recovery of high temperature heat released by cooling syngas exiting the ATR, and it operates with an outlet temperature of 700°C. Steam reforming is carried out in a nickel-based catalytic Auto Thermal Reformer (ATR). Compressed air boosted from the GT compressor is used as oxidizing agent and its mass flow rate is adjusted for achieving an ATR outlet syngas temperature of 950°C, which ensures a good methane

conversion that is essential for an effective carbon capture. The ATR operates with an outlet pressure of 34.5 bar. The presence of a large amount of nitrogen in the syngas product at ATR outlet (22.7% vol. in stream #11) is helpful for hydrogen dilution to reduce NO_x emissions from gas turbine combustion.

CO-rich syngas from the ATR is sent to a two-stage WGS section with intermediate cooling by HP steam production. At LT-WGS reactor outlet (stream #17), over 96% of carbon in the natural gas input is finally oxidized to CO₂.

Syngas from the WGS section is cooled down to nearly ambient temperature and sent to the MDEA-based CO₂ absorption unit. In the process considered (Meissner and Wagner, 1983; Romano et al., 2010), the solvent is physically and thermally regenerated in a LP flash and in a stripper, producing semi-lean and lean solvent streams used in an absorber for CO₂ separation. 95% of CO₂ capture efficiency is obtained for this process, with thermal and electric consumptions of 0.99 MJ_{th}/kg_{CO2} and 0.12 MJ_e/kg_{CO2} respectively. The >99.9% purity CO₂ stream (dry basis) at 1.1 bar released by the process is compressed for storage with no further treatment. The decarbonized H₂-based fuel obtained after the MDEA process is moisturized in a saturator by using low and medium temperature heat for NO_x emission reduction and then heated up to 200°C to be burnt in the gas turbine combustor. A single train, state-of-the-art combined cycle, calculated with the assumptions given in Table A1 in the Appendix, has been considered.

The steam cycle is based on a two-pressure level steam generator with reheat. A low-pressure level is not necessary for a good heat recovery because a large amount of high-pressure steam is generated in syngas coolers and low temperature heat is therefore required in economizers. In this case, the intermediate pressure level has been set to 38 bar in order to match the HP turbine discharge pressure to the ATR operating pressure, allowing to extract the steam for reforming from the cold reheat. Parameters used for calculating the steam cycle correspond to those already used for the Ca-Cu based power plant, which have been given in Table A1 in the Appendix.

Finally, as a matter of comparison, a reference state-of-the-art NGCC power plant without CO₂ capture has been also considered. The steam cycle is based on a three-pressure level HRSG with reheat, with evaporation pressures of 130/28/4 bar and live steam SH/RH temperatures of 565°C.

4. Mass and energy balance results

The mass and energy balances of the power plant components have been calculated using the in-house GS (Gas-steam cycle) code (GECOS, 2016). The distinctive feature of the code is the built-in model for the cooled gas turbine, suitable for calculating hydrogen-fired cooled gas turbines (Chiesa and Macchi, 2004; Gazzani et al., 2014). The Ca-Cu reactors have been modelled as black box in GS, providing the heat and mass balance resulting from the PHM model, as described in the previous section. The CO₂ compression unit has been calculated with Aspen Plus using the

Peng Robinson equation of state. Both benchmark power plants (i.e. with and without CO₂ capture) have also been calculated with the GS code, with the exception of CO₂ compression and MDEA-based CO₂ separation unit.

For evaluating the performance, the net electric efficiency of the whole power plant and the CO₂ capture rate (CCR) are used. Moreover, the specific primary energy consumption for CO₂ avoided (SPECCA) has been also considered. The SPECCA is calculated according to Eq. (3), where η and η_{ref} correspond to the net electric efficiencies of the power plant with CO₂ capture considered (i.e. the Ca-Cu power plant or the benchmark with CO₂ capture) and the benchmark power plant without CO₂ capture, respectively, and E_{CO_2} are the specific CO₂ emissions per unit of net electricity produced.

$$SPECCA = \frac{\left(\frac{1}{\eta} - \frac{1}{\eta_{ref}}\right)}{E_{CO_2,ref} - E_{CO_2}} \cdot 3600 \quad (3)$$

Table 2 summarises the results obtained for the NGCC benchmark power plants with and without CO₂ capture and for the Ca-Cu process integrated with the combined cycle considering the operating conditions given in Table 1 (referred to as ‘base case’). The electric efficiency of the base case power plant integrated with the Ca-Cu process is 45.29% (LHV-based), 1.8%-points lower than that of the benchmark NGCC power plant with CO₂ capture. The gross electric efficiency (referring to the power of the GT and of the steam turbine) is higher in the Ca-Cu plant with respect to the benchmark plant with CO₂ capture. Therefore, the reason of the lower net efficiency is related to the electric consumption of auxiliaries in the Ca-Cu power plant, which is noticeably larger than that of the benchmark plant. This is especially due to the electric consumption of the N₂ recycle fan from stage B’ to stage B that consumes 68.2 MW_e. Regarding the CO₂ capture ratio, a slightly lower CCR value results for the Ca-Cu process compared to the benchmark plant (89.1 vs. 91.6 %).

Table 2 Performance obtained for the Ca/Cu NGCC plant for the base case, the ‘compact reactor’ case and the ‘optimized case’ analysed and for the benchmark NGCC without and with CO₂ capture

	Benchmark plants		Ca/Cu plants		
	NGCC w/o capture	ATR+MDEA NGCC	Base case	Compact reactor	Optimized case
Reactor module size DxL (m)	--	--	4.5x10	2.25x5	2.25x5
Number of parallel B-B’ reactors	--	--	4	4	5
<i>Power balance (MW_e)</i>					
Gas turbine gross power	273.2	291.3	315.3	315.3	315.3
Gas turbine auxiliaries	-1.09	-1.17	-1.26	-1.26	-1.26
Steam turbine gross power	147.1	119.7	224.2	216.2	206.1
Steam Cycle Pumps power	-1.79	-3.41	-4.28	-4.15	-3.98
Auxiliaries for condenser heat rejection	-1.86	-1.09	-1.89	-1.82	-1.72
Air booster compressor power	--	-10.91	-5.27	-5.27	-3.05

MEA/MDEA plant auxiliaries	--	-5.13	--	--	--
N ₂ -rich gas recycle fan		--	-68.17	-50.12	-29.56
CO ₂ compressor power		-14.73	-17.16	-17.16	-17.16
Auxiliaries for heat rejection other than condenser		-1.06	-0.77	-0.77	-0.77
Natural gas expander		--	9.72	9.72	9.72
Other auxiliaries		-0.19	-0.03	-0.03	-0.03
Net power output	415.6	373.3	450.5	460.7	473.6
<i>General performance indicators</i>					
Thermal input, MW _{LHV}	711.3	792.2	994.6	994.6	994.6
Gross electric efficiency, %LHV	59.09	51.88	54.24	53.44	52.42
Net electric efficiency, %LHV	58.42	47.12	45.29	46.31	47.61
Carbon capture ratio, %	--	91.56	89.14	89.14	89.14
Specific emissions, kg/MWh	350.6	39.6	51.53	50.39	49.01
CO ₂ avoided, %	--	88.7	85.30	85.63	86.02
SPECCA, MJ _{LHV} /kgCO ₂	--	4.75	5.97	5.36	4.64

Based on these results, a sensitivity analysis on the most significant pressure drops of the Ca/Cu process has been carried out. Moreover, these quantities are later used to optimize the plant from an economic point of view, since as general rule the design pressure drops through heat exchangers and reactors can be optimized as trade-off between operating and capital cost. This sensitivity analysis is not performed on the pressure drops of all the reactors and heat exchangers, but it is limited to the pressure drops that have a direct influence on the design of the Ca/Cu reactors and of the heat exchangers with the highest capital cost. Concretely, the impact of the pressure drop through the heat exchangers placed on the N₂-rich gas recycled and at stage B inlet (H and G in Figure 1, respectively) and through the fixed bed reactors have been analysed.

Concerning the pressure drop along the fixed bed reactors, the reactor dimensions chosen for the base case (given in Table 1) result in a pressure drop of 2 bar and 3 bar in each reactor operating in stage B and B' respectively, calculated according to the Ergun law in the 1-D pseudo-homogenous reactor model (Martini et al., 2016). Such pressure drop, together with the pressure drop in the heat exchanger H placed downstream stage B' of the Ca/Cu process, make the pressure difference along the recycle fan almost 7 bar, which leads to an electric consumption of 68.2 MWe (Table 2). Therefore, it has been analysed the possibility of working with a 'compact reactor' Ca/Cu process, where each reactor is divided into four sub-reactors with a diameter of 2.25 m and a length of 5 m (i.e. keeping the original L/D ratio), sharing the same set of valves and leading to reduced pressure drops (Figure 10). Since the reactor length in this configuration is halved with respect to that of the base case while the overall cross-section remains unchanged, the pressure drop along the reactor in each Ca/Cu stage halves compared to the base case. For carrying out this analysis, the outlet pressure of stage B has been fixed to 19 bar, which is the pressure assumed for the base case. In this way, when reducing the pressure drop across the reactor, inlet pressure to stage B is consequently reduced, resulting also in a higher pressure drop along heat exchanger G (i.e. outlet pressure for the booster and recycle fan is kept at 21.3 bar as

indicated in **Error! No se encuentra el origen de la referencia.** for the base case). Performance results obtained for this case have been included in Table 2 referred to as ‘*compact reactors*’ case. As noticed, the electric consumption of the N₂-rich gas recycle fan is greatly reduced to 50.1 MW_e for this case, which causes an increase of the electric efficiency of the whole plant by 1%-point. Moreover, as will be seen in the economic analysis section, such reduction in the reactor dimension allows reducing the corresponding capital and operating costs since the total reactor volume is also reduced.

Figure 10 Compact reactor configuration, where each reactor is divided into 4 smaller sub-reactors sharing the same set of valves.

Figure 11 Net electric efficiency as a function of: (a) the pressure drop across the heat exchanger ‘H’ downstream stage B’ ($P_{out,booster}=21.3$ bar, Δp in heat exchanger upstream stage B = 1.3 bar); (b) the discharge pressure of the air booster compressor (total Δp in HX-H downstream stage B’ = 1.6 bar)

The effect of decreasing the outlet pressure of the air booster from the value of 21.3 bar considered by far is also evaluated (Figure 11b). Similarly to the previous analysis, this change has been implemented for the ‘compact reactor’ case, keeping the pressure at stage B outlet equal to 19 bar. In this case it is studied the possibility of reducing the booster outlet pressure to 20.3 bar and so work with a limited pressure drop in heat exchanger G of 0.3 bar. Such reduction in the booster outlet pressure results also in a reduction of the compression ratio of the N_2 -rich gas recycled fan, which should work with the same outlet pressure of the booster. According to these assumptions, both air booster and fan electric consumptions decrease by 20% and 17% respectively, when booster outlet pressure reduces to 20.3 bar, which makes the net electric efficiency reach 46.9 %.

Based on the optimization of the cost of electricity discussed in the following section, an optimum pressure drop of 1.3 bar in heat exchanger H has been calculated. Moreover, working with a booster outlet pressure of 20.3 bar allows reducing the investment and operating costs of the plant. Considering such set of optimised conditions, the possibility of having 3 or 5 reactors working in stages B and B’ instead of 4 as in the ‘base case’ and ‘compact reactor’ case is finally evaluated. Increasing the number of reactors in B-B’ to 5 allows reducing the pressure drop in stages B and B’, since the gas velocity in each sub-reactor operating in these stages would decrease, while reducing also the booster outlet pressure (Figure 12a). As a result, air booster and fan electric consumptions decrease by 42% and 41% respectively, with respect to the ‘compact reactor’ case, which results in a net electric efficiency of 47.6 % (Figure 12b) that is 0.5%-points higher than that of the benchmark power plant with CO_2 capture. Despite such increase in the efficiency, the higher number of reactors working in stages B/B’ will have an impact on the economics of the power plant, and it is therefore evaluated from the economic point of view in the following

section. Performance results obtained for this case have been included as ‘*optimized case*’ in Table 2; **Error! No se encuentra el origen de la referencia..** Properties of the streams shown in Figure 1 of this optimised case are reported in the appendix in Table A2.

Figure 12 (a) Pressure at the air booster compressor outlet and pressure increase across the recirculating fan and (b) net electric efficiency as a function of the number of reactors in stage B and B' (total Δp in heat exchanger ‘H’ downstream stage B’ equal to 1.3 bar)

5. Economic analysis

5.1. Methodology

The direct cost or capital cost of each component is evaluated through any of these routes: (i) an exponential cost scaling technique like that indicated in Eq. (4), which is used to calculate the capital cost of a piece of equipment starting from a reference capital cost (C_0) for a reference size (S_0), a scaling parameter (SP) and an exponent (n); (ii) other cost functions available in the literature; (iii) data supplied from industrial partners of ASCENT project (ASCENT, 2014); or (iv) data from commercial software (AspenTech, 2016).

$$C = C_0 \cdot \left(\frac{SP}{SP_0} \right)^n \quad (4)$$

The assumptions related to the capital cost estimation methodology are given in Table B1. Installation factors (IF) are derived from the same sources where the material and labour cost are taken (DOE/NETL, 2010; Franco et al., 2011; Rath, 2010). For some pieces of equipment, IF are not reported in the table since they are included in the provided cost data. All installed costs are referred to 2017. The sum of the installed cost of all the pieces of equipment corresponds to the *Total Direct Plant Cost* (TDPC).

For calculating the TDPC of the Ca/Cu reactors, the sum of the reactor vessel cost and the cost of the switching valves is considered. Each reactor vessel is considered to be made of steel with a design pressure of 26 bar. To withstand the high temperature reached inside the vessels, a 300 mm

thick refractory material covers the internal surface of the reactor. This allows to have a suitable temperature (i.e. around 70°C) on the outer surface of the steel vessel. For the initial fill of the reactors, the assumed materials costs are: 5 €/kg for the Ca-based sorbent, 20 €/kg for the Cu-based carrier and 50 €/kg for the Ni-based catalyst. The lifetime of each material has been considered 5 years. This means that the operation cost of the Ca/Cu system includes a complete material replacement every 5 year. Moreover, each reactor has 5 switching valves on the outlet pipe and 5 switching valves on the inlet pipe. An installation cost for the complete Ca/Cu system equal to 50% of the sum of the vessels and valves cost has been assumed.

Indirect Costs are assumed to account for 14% of the *Total Direct Plant Cost* (TDPC) as displayed in **¡Error! No se encuentra el origen de la referencia. B2**. The sum of the direct and the indirect costs is the so-called Engineering, Procurement and Construction Cost (EPC). Owner's Costs and Contingencies, which account for 15 % of the EPC, must be added to account for planning, designing and commissioning the plant. The resulting amount is the *Total Plant Cost* (TPC). Finally, initial fills of chemicals and catalysts in the reactors beds have been added to the TPC (Table B3).

Operation and maintenance (O&M) of the plant requires annualized expenses in terms of labour, consumables (fuel, water and chemicals) and materials (e.g. spare parts and others). The labour cost is essentially divided into operational labour, maintenance labour and administrative & support labour. The average gross hourly salary for operational labour is assumed to be 40.2 €/hour (Franco et al., 2011). A total number of 15 employees per shift has been considered (US DOE, 2010). Concerning the maintenance labour, the yearly expense has been computed according to Table B4 (Demoys, 2010). The shares of labour and materials on the total maintenance cost have been assumed to be 40% and 60%, respectively. Administrative and support labour have been considered equal to 25% of the total labour cost, whereas property taxes and insurance contribute for 2% of the TPC. Finally, variable O&M costs (Table B5) are essentially related to the purchase of NG, having a unit cost of 6.5 €/GJ_{LHV}. Raw water and process water make-up are needed due to evaporation in the cooling towers and blowdown of steam cycle or steam addition to the process gas. To reduce the need of water make-up, the condensate produced in the plant is assumed to be recovered and treated to produce demi-water.

In Table B6 assumptions for the long-run profitability analysis have been reported. All costs are referred to the same year 2017 by means of the CEPCI coefficients CEPCI (n.d.). The availability of the plants has been assumed equal to 7500 hours. However, for the first year of operation the number of hours has been reduced to 5700 due to unforeseen circumstances, typical of newly built plants. An efficiency decay causing net power output reduction of 0.2% per year has also been assumed in the economic calculation.

The economic assessment allows the calculation of the Breakeven Selling Price (BESP) or Levelized Cost of Electricity (LCOE) and the Cost of CO₂ Avoided (CCA). The CCA is calculated as follows:

$$CCA = \frac{LCOE - LCOE_{ref}}{E_{CO_2,ref} - E_{CO_2}} \quad (5)$$

For the calculation of the CCA the reference NGCC plant without CO₂ capture has been considered. The calculated LCOE for this benchmark is 58.25 €/MWh.

5.2. Results

In the performance analysis section, three different Ca/Cu plant cases have been calculated, namely ‘*base case*’, ‘*compact reactor*’ case and ‘*optimized case*’. In this section, the economic analysis for these three cases is included. In the ‘*base case*’, the Ca/Cu system is constituted by 14 reactors (2 for stage A, A’ and C, 4 for stage B and B’), each one 10 m long and with an inner diameter of 4.5 m. In the ‘*compact reactor*’, the same total number of reactors is assumed, but each reactor is divided into 4 vessels, each one 5 m long and with an inner diameter of 2.25 m, sharing the same set of valves (Figure 10). In the ‘*Optimized case*’ 16 reactors are used (one more unit in B and in B’ is considered) with the same number and size of sub-reactors of the ‘*compact reactor*’ case. Main features of the Ca/Cu reactor system of the three cases assessed are summarized in Table 3.

Table 3 Main differences and assumptions in the three cases analysed for the NGCC power plant integrated with the Ca/Cu process

	Base case	Compact reactor	Optimized case
Reactors diameter [m]	4.5	2.25	2.25
Reactors length [m]	10	5	5
Number of parallel reactors in B-B’ [-]	4	4	5
Total number of reactors [-]	14	14	16
Total reactors useful volume [m ³]	2227	1113	1272
Gas side pressure drop in evaporator upstream stage B [bar]	0.3	1.3	0.3
Gas side pressure drop in evaporator-superheater downstream stage B’ [bar]	1.6	1.6	1.3
Pressure drop in stage B [bar]	2	1	0.65
Pressure drop in stage B’ [bar]	3	1.5	1.3
Pressure at the air booster compressor outlet [bar]	21.3	21.3	19.95
Pressure increase across the recirculating fan [bar]	6.9	5.4	3.21

Table 4 includes the cost breakdown for each reactor in the three cases studied for the NGCC power plant integrated with the Ca/Cu process. The steel and the valves represent the most important items from the total reactors cost. It can be also observed that moving from the ‘*base*

case’ to the ‘*compact reactors case*’ the steel and functional material cost is significantly reduced (by about 30%) since the reactors volume is halved. In Table B8 in the Appendix the valves cost for the three final cases is reported.

Table 4 Details of reactors components cost for three cases of the NGCC power plant integrated with the Ca-Cu process

	Base case	Compact reactor	Optimized case
Costs for each reactor [M€]			
<i>Insulation</i>	0.19	0.21	0.21
<i>Steel (including head, shell, nozzle)</i>	3.88	2.51	2.51
<i>Valves</i>	1.32	1.32	1.26
<i>Ca-based sorbent</i>	0.38	0.19	0.19
<i>Cu-based oxygen carrier</i>	1.47	0.73	0.73
<i>Ni-based catalyst</i>	0.98	0.49	0.49
Total material cost of each reactor [M€]	4.07	2.72	2.72
Total cost of each reactor [M€]	5.39	4.04	3.99
Total installed reactors cost (all reactors) [M€]	113.19	84.89	95.68
Cost of functional material per reactor [M€]	2.83	1.41	1.41
Total cost of functional material (all reactors) [M€]	39.60	19.80	22.63

Details of the cost associated to the heat exchangers of the ‘*optimized case*’ are given in Table B9 in the Appendix. As anticipated before, the most expensive shell and tubes heat exchangers correspond to evaporator G, evaporator H and superheater H in Figure 1. For this reason, these heat exchangers have been optimized in terms of the pressure drops considering a trade-off between the economic results and the global plant performance.

Capital cost of the different units of the Ca/Cu-based power plant is detailed in Table 5 for the three cases studied. The main differences among the cases are related to the Ca/Cu reactors and valves, the accessory electric plant (mainly because of the difference in the consumption of auxiliaries) and the shell and tube heat exchangers. Detail of the variable and fixed O&M costs are reported in Table 6, where the main differences are related to the cost for the replacement of the Ca/Cu functional material, which is proportional to the reactors volume.

Table 5 Capital cost breakdown for three cases of the NGCC power plant integrated with the Ca-Cu process

Cost item		Base case		Compact reactor		Optimized case	
		M€	€/kWe	M€	€/kWe	M€	€/kWe
Feedwater & Miscellaneous BOP Systems		51.76	115	50.80	110	49.58	105
CO ₂ Compression & Drying		19.04	42	19.04	41	19.04	40
NG Pre-reforming		0.58	1.3	0.58	1.3	0.58	1.2
NG Desulphurization		0.50	1.1	0.50	1.1	0.50	1.1
Gas Turbine/Accessories		64.23	143	64.23	139	64.23	136
HRSG, Ducting & Stack		44.81	99	44.81	97	44.81	95
Steam Turbine Generator		61.33	136	60.16	131	58.66	124
Condenser & Auxiliaries		5.20	12	5.17	11	5.14	11
Cooling Water Systems		20.51	46	20.13	44	19.65	41
Accessory Electric Plant		51.49	114	46.61	101	39.70	84
Recirculating fan		2.46	5.5	2.17	4.7	2.00	4.2
NG expander		2.14	4.7	2.14	4.6	2.14	4.5
Air booster compressor		1.28	2.8	1.28	2.8	0.84	1.8
Shell and tube Heat Exchangers		26.38	59	21.03	46	24.41	52
CO ₂ rich-gas HRSG		5.76	13	5.76	12	5.76	12
H ₂ recycle compressor		0.10	0.2	0.10	0.2	0.10	0.2
Reactors and valves		113.19	251	84.89	184	95.68	202
	TDPC	470.76	1046	429.40	931	432.81	915
Yard improvement		7.06	16	6.44	14	6.49	14
Service facilities		9.42	21	8.59	19	8.66	18
Engineering & consultancy		21.18	47	19.32	42	19.48	41
Building		18.83	42	17.18	37	17.31	37
Miscellaneous		9.42	21	8.59	19	8.66	18
	EPC	536.67	1193	489.52	1062	493.41	1043
Other Owner's costs		80.50	179	73.43	159	74.01	156
	TPC	617.17	1372	562.95	1221	567.42	1199
Hydrodesulphurization catalyst		0.36	0.8	0.36	0.8	0.36	0.8
ZnO absorbent		1.04	2.3	0.99	2.2	0.99	2.1
Stage A pre-reformer catalyst		0.63	1.4	0.62	1.3	0.62	1.3
Stage C pre-reformer catalyst		0.79	1.8	0.77	1.7	0.77	1.6
Ca/Cu reactors material		39.60	88	19.80	43	22.63	48
	TPC+I.F	659.59	1466	585.48	1270	592.78	1253

Table 6 Calculated variable and fixed O&M costs for three cases of the NGCC power plant integrated with the Ca-Cu process

		Base case		Compact reactor		Optimized case	
		M€/year	€/MWh	M€/year	€/MWh	M€/year	€/MWh
Variable O&M							
Natural gas		179.29	53.07	179.29	51.89	179.29	50.48
Raw water		1.95	0.58	1.90	0.55	1.83	0.52
Process water		11.41	3.38	11.41	3.30	11.41	3.21
	Total	192.65	57.02	192.60	55.74	192.53	54.21
Fixed O&M							
Operating labor		5.29	1.57	5.29	1.53	5.29	1.49
Property taxes and insurance		12.34	3.65	11.26	3.26	11.35	3.20
Maintenance labor		4.02	1.19	3.76	1.09	3.75	1.06
Maintenance materials		6.03	1.79	5.63	1.63	5.63	1.58
Administrative & support labor		2.33	0.69	2.26	0.65	2.26	0.64
Chemicals and replacements							
Hydrodesulphurization catalyst		0.04	0.01	0.04	0.01	0.04	0.01
ZnO absorbent		0.48	0.14	0.47	0.14	0.47	0.13
Stage A pre-reformer catalyst		0.16	0.05	0.15	0.04	0.15	0.04
Stage C pre-reformer catalyst		0.20	0.06	0.19	0.06	0.19	0.05
SCR catalyst		0.08	0.02	0.08	0.02	0.08	0.02
Ammonia (19% NH ₃)		0.15	0.04	0.15	0.04	0.15	0.04
Ca/Cu reactors material		7.92	2.34	3.96	1.15	4.53	1.27
	Total	39.04	11.56	33.24	9.62	33.88	9.54

Table 7 summarizes the cost of the electricity breakdown for the NGCC power plant integrated with the Ca/Cu process for the three cases analysed in detail in this work, together with the costs of the benchmark ATR+MDEA NGCC power plant. Breakdown of Capital and operating costs of the benchmark plant are reported in Tables B10 and B11 in Appendix B. A reduction in the LCOE of around 8 €/MWh is possible when moving from the Ca/Cu ‘*base case*’ to the ‘*optimized case*’. The reasons of this reduction are mainly related to the decrease of the reactors volume and hence of vessels and functional material cost, and the higher efficiency related to the reduction of the pressure drops in stage B-B’ loop. As discussed in the plant performance section, decreasing pressure drops in both the main heat exchangers and in the reactors (by reducing the length and increasing their number) resulted convenient from the point of view of electric plant efficiency since the electric consumption of the N₂ recycle fan and the air booster compressor reduced. This improvement in the net electric efficiency translates in a reduction of both the cost for NG and of the specific capital cost of the whole plant (thanks to the increase of the net power output). Considering the optimized case conditions, the NGCC integrated with the Ca/Cu process has an electricity cost of 82.60 €/MWh, which is 2.2 €/MWh below the benchmark power plant with CO₂ capture. As to the CCA, the Ca/Cu based power plant has a CCA of 80.75 €/tCO₂ that is 4.6 €/tCO₂ lower than that of the benchmark power plant (85.38 €/tCO₂).

Table 7 Cost of electricity breakdown for the NGCC power plant integrated with the Ca-Cu process

	Benchmark plants		Ca/Cu plants		
	NGCC w/o capture	ATR+MDEA NGCC	Base case	Compact reactor	Optimized case
Capital cost [€/MWh]	11.16	19.84	21.97	19.15	18.85
Natural gas [€/MWh]	41.13	52.08	53.07	51.89	50.48
Other variable O&M [€/MWh]	0.52	2.43	3.95	3.85	3.73
Fixed O&M [€/MWh]	5.44	10.45	11.56	9.62	9.54
Cost of electricity (LCOE) [€/MWh]	58.25	84.80	90.55	84.51	82.60
Cost of CO₂ Avoided (CCA) [€/tCO₂]	-	85.38	108.01	87.48	80.75

In Figure 13 the Cost of CO₂ avoided and the LCOE are represented as a function of the installed cost of the Ca/Cu system. This cost accounts only for the reactors material (steel and insulation) and valves. The same initial fill cost and material replacement cost of the ‘*optimized case*’ has been assumed in this analysis. From the figure it can be concluded that the reactors material costs and the valves cost of the Ca/Cu process should be lower than around 130 M€ to have both the LCOE and the CCA below the reference ATR+MDEA plant.

Figure 13 CCA and LCOE for the NGCC integrated with the Ca/Cu process as a function of the reactors investment cost (reactors and valves). The horizontal dotted lines represent the CCA and LCOE of the benchmark plant (ATR+MDEA). The vertical dotted line represents the installed cost of the ‘optimized case’

Regarding the initial fill costs of the reactors of the Ca/Cu process, Figure 14 shows the variation of the CO₂ avoided cost and the LCOE of the Ca/Cu system when modifying the initial fill cost of the Ca/Cu reactors. For this analysis, it has been assumed that the reactors investment cost (related to the reactors and the valves) correspond to those of the ‘*optimized case*’. As can be noticed from this analysis, an initial fill cost lower than around 30 M€ makes the Ca/Cu plant economically better in terms of CCA and LCOE with respect to the ATR+MDEA plant.

Figure 14 CCA and LCOE for the NGCC integrated with the Ca-Cu process as a function of the reactors initial fill cost. The horizontal dotted lines represent the CCA and LCOE of the benchmark plant (ATR+MDEA). The vertical dotted line represents the initial fill cost of the “optimized case”

The economic analysis carried out has demonstrated that the NGCC plant integrated with the Ca/Cu process can be economically competitive with respect to the benchmark NGCC integrated with the ATR+MDEA. A decrease in the LCOE and the CCA values shown in this analysis could be possible by means of a decrease in the capital cost of the Ca/Cu process. Reactors, valves and initial fill strongly affect the economy of the plant and the decrease of the costs associated to these items can decrease the TDPC, making the plant even more competitive with respect to the benchmark technology.

Conclusions

In this work, a techno-economic analysis of a natural gas combined cycle integrated with the Ca/Cu process has been carried out. Balances of the whole plant have been calculated and the cost of electricity and of the CO₂ avoided have been estimated. Based on the analyses carried out, it can be concluded that:

- From an economic point of view, it is important to optimise the Ca/Cu reactor size and the pressure drop in heat exchangers with the highest capital cost in order to achieve competitive efficiency and economic performance.
- Vessels of reduced size for the Ca/Cu process are important to limit pressure drop, which will reduce the electric consumption associated to the auxiliaries needed in the Ca/Cu process.

Moreover, reduced Ca/Cu reactors volume needs lower amount of functional material and therefore lower initial fill and replacement cost.

- An optimized case has been calculated, considering reduced vessel size and optimised pressure drops in the crucial heat exchangers. For this case, net electric efficiency results 0.5% points higher than the benchmark power plant based on an ATR and MDEA process for CO₂ capture. An electricity cost of 82.60 €/MWh has been estimated, which is 2.2 €/MWh less than the benchmark. Regarding the CCA, the Ca/Cu based power plant has a cost of CO₂ avoided of 80.7 €/t_{CO2}, 4.6 €/t_{CO2} lower than that of the benchmark power plant integrated with the ATR+MDEA.

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Appendix A. Assumptions and results of heat and mass balance calculations

Table A1 Main assumptions used for calculating the balances of the NGCC integrated with the Ca/Cu process

<i>Natural Gas pre-treatment</i>	
Natural gas composition and LHV	CH ₄ 89.00, C ₂ H ₆ 7.00, C ₃ H ₈ 1.00, C ₄ H ₁₀ 0.11, CO ₂ 2.00, N ₂ 0.89 LHV 46.482 MJ/kg, HHV 51.454 MJ/kg
Operating temperature of the desulphurization unit [°C]	365
Inlet temperature at the HP/LP adiabatic pre-reformers [°C]	490
S/C molar ratio at HP adiabatic pre-reformer inlet before stage A-A' [-]	5
S/C molar ratio at HP adiabatic pre-reformer inlet before stage C [-]	1
Pressure at the adiabatic pre-reformers inlet [bar]	26.4
Pressure drop in the adiabatic pre-reformers [% of inlet pressure]	3
<i>Gas turbine (GT)</i>	
Mass flow rate of the flue gas at GT outlet [kg/s]	665
Compressor pressure ratio [-]	18
Compressor polytropic efficiency [%]	92.5
Turbine Inlet Temperature [°C]	1360
Turbine expansion stages cooled/uncooled [-]	3/1
Isentropic efficiency of the expansion stages [%]	89.5/90.8/92.1/91.6*
Organic efficiency of compressor/turbine [%]	99.6
Electric generator efficiency [%]	98.5
<i>Steam cycle</i>	
Live steam temperature SH/RH [°C]	565/565
Evaporating pressure [bar]	130.0
RH inlet pressure [bar]	27.0
Pressure losses in SH/RH tubes [% of inlet pressure]	5.1
SH/RH steam pressure losses at turbine inlet valve [% inlet pressure]	2
Feedwater pump outlet pressure [bar]	150
Pressure loss in the deaerator [bar]	2.0
Deaerator pressure [bar]	3.6
Condensing pressure [bar]	0.048
Number of LP/HP feedwater heaters (including deaerator) [-]	2/2
HP turbine isentropic efficiency [%]	94.7
IP turbine isentropic efficiency [%]	92.7
LP turbine isentropic efficiency [%]	91.3
Turbine mechanical efficiency [%]	99.6
Electric generator efficiency [%]	98.5
Feedwater pump hydraulic efficiency [%]	85.0
Mechanical-electric efficiency of the feedwater pump [%]	94.0
<i>Auxiliaries</i>	
Booster compressor polytropic efficiency [%]	88.1
Booster compressor organic efficiency [%]	94.0
Booster compressor outlet pressure [bar]	21.3
NG expander isentropic polytropic efficiency [%]	82.1
NG expander organic efficiency [%]	94.0
Recycle N ₂ -rich gas fan polytropic efficiency [%]	89.8
Recycle N ₂ -rich gas fan organic efficiency [%]	96.0
Pressure at NG expander outlet [bar]	2

Gas pressure drop in heat exchanger G [bar]	0.3
Gas pressure drop in heat exchanger H [% of inlet pressure]	10
<i>Inter-cooled CO₂ compressor</i>	
Number of inter-cooled compression stages [-]	5
Isentropic efficiency [%]	84.0
Electric-mechanical efficiency [%]	94.0
Pump hydraulic efficiency [%]	80.0
Pump electric-mechanical efficiency [%]	94.0
CO ₂ pressure at compressor discharge [bar]	89.1

*Calculated using the GS model for the H₂-rich gas fuelled turbine described by Chiesa and Macchi (2004)

Table A2 Temperature, pressure, flow rate and compositions of the main process streams shown in Figure 1 for the ‘optimised case’ NGCC integrated with the Ca/Cu process

No	T	P	G	Q	G*LHV	Composition [%mol]										
	[°C]	[bar]	[kg/s]	[kmol/s]	[MW]	Ar	CH ₄	CO	CO ₂	H ₂	H ₂ O	N ₂	O ₂	C ₂ H ₆	C ₃ H ₈	C ₄ H ₁₀
1	25.0	28.0	21.40	1.188	994.6		89.00	0.00	2.00	0.00	0.00	0.89		7.00	1.00	0.11
2	25.6	28.0	21.47	1.213	1001.1		87.19	0.01	1.97	2.00	0.00	0.88		6.85	0.98	0.11
3	365.0	27.2	21.47	1.213	1001.1		87.19	0.01	1.97	2.00	0.00	0.88		6.85	0.98	0.11
4	365.0	27.2	8.40	0.475	391.7		87.19	0.01	1.97	2.00	0.00	0.88		6.85	0.98	0.11
5	332.5	27.01	9.09	0.504							100.00					
6	340.0	27.01	17.49	0.979	391.7		42.28	0.01	0.96	0.97	51.52	0.43		3.32	0.48	0.05
7	490.0	26.39	17.49	0.979	391.7		42.28	0.01	0.96	0.97	51.52	0.43		3.32	0.48	0.05
8	450.7	25.60	17.49	1.033	393.4		45.26	0.08	3.47	7.18	43.61	0.40				
9	221.1	2.00	17.49	1.033	393.4		45.26	0.08	3.47	7.18	43.61	0.40				
10	670.0	1.45	17.49	1.033	393.4		45.26	0.08	3.47	7.18	43.61	0.40				
11	699.3	1.10	76.95	2.612					43.94		55.90	0.16				
12	452.1	1.05	37.50	1.273					43.94		55.90	0.16				
13	452.1	1.05	39.45	1.339					43.94		55.90	0.16				
14	90.0	1.02	76.95	2.612					43.94		55.90	0.16				
15	30.0	1.02	76.95	2.612					43.94		55.90	0.16				
16	36.1	150.00	50.64	1.152					99.64			0.36				
17	365.0	27.2	13.07	0.738	609.4		87.19	0.01	1.97	2.00	0.00	0.88	0.00	6.85	0.98	0.11
18	332.5	27.0	70.67	3.923							100.00					
19	317.2	27.0	83.74	4.661	609.4		13.81	0.00	0.31	0.32	84.16	0.14		1.09	0.16	0.02
20	490.0	26.4	83.74	4.661	609.4		13.81	0.00	0.31	0.32	84.16	0.14		1.09	0.16	0.02
21	429.4	25.6	83.74	4.842	620.0		14.03	0.02	2.15	6.38	77.29	0.13				
22	575.0	25.1	83.74	4.842	620.0		14.03	0.02	2.15	6.38	77.29	0.13				
23	784.9	25.0	61.34	5.557	723.6		1.11	2.10	1.75	47.71	47.22	0.12				
24	700.0	25.0	61.34	5.557	723.6		1.11	2.10	1.75	47.71	47.22	0.12				
25	704.4	25.0	54.26	5.374	719.9		1.46	0.36	0.31	50.13	47.62	0.12				

26	549.6	24.3	54.26	5.374	719.9		1.46	0.36	0.31	50.13	47.62	0.12	
27	369.1	23.5	54.26	5.374	719.9		1.46	0.36	0.31	50.13	47.62	0.12	
28	63.5	30.00	0.07	0.025	6.5		2.78	0.69	0.59	95.53	0.18	0.23	
29	369.1	23.5	53.77	5.326	713.4		1.46	0.36	0.31	50.13	47.62	0.12	
30	15.0	1.0	641.17	22.224		0.92			0.03		1.03	77.28	20.73
31	416.5	18.2	545.38	18.903		0.92			0.03		1.03	77.28	20.73
32	416.5	18.18	355.80	12.332		0.92			0.03		1.03	77.28	20.73
33	1436.1	17.63	514.98	20.060		0.78			0.72		27.60	65.62	5.27
34	601.0	1.04	664.97	25.258		0.81			0.58		22.14	68.02	8.45
35	90.0	1.01	664.97	25.258		0.81			0.58		22.14	68.02	8.45
36	416.5	18.2	135.39	4.693		0.92			0.03		1.03	77.28	20.73
37	436.0	20.0	135.39	4.693		0.92			0.03		1.03	77.28	20.73
38	382.8	20.0	915.95	32.431		1.12			0.64		1.26	93.98	3.00
39	340.0	19.7	915.95	32.431		1.12			0.64		1.26	93.98	3.00
40	694.9	19.0	885.97	31.49		1.15			0.74		1.30	96.81	
41	694.9	19.0	780.56	27.739		1.15			0.74		1.30	96.81	
42	546.2	18.0	780.56	27.739		1.15			0.74		1.30	96.81	
43	340.0	16.7	780.56	27.739		1.15			0.74		1.30	96.81	
44	373.7	20.0	780.56	27.739		1.15			0.74		1.30	96.81	
45	694.9	19.0	105.41	3.746		1.15			0.74		1.30	96.81	
46	15.0	1.0	79.76	4.427							100.00		
47	24.7	6.6	116.93	6.490							100.00		
48	140.0	5.6	116.93	6.490							100.00		
49	142.0	150.0	172.23	9.560							100.00		
50	299.0	148.0	172.23	9.560							100.00		
51	288.3	148.0	21.24	1.179							100.00		
52	297.9	148.0	5.13	0.285							100.00		
53	297.9	148.0	188.34	10.454							100.00		
54	331.0	130.0	5.13	0.285							100.00		

55	330.8	130.0	188.34	10.454	100.00
56	330.8	130.0	162.94	9.044	100.00
57	377.9	130.0	162.94	9.04	100.00
58	330.8	130.0	8.63	0.479	100.00
59	565.0	123.4	8.63	0.479	100.00
60	330.8	130.0	25.40	1.410	100.00
61	340.2	130.0	184.84	10.260	100.00
62	565.0	123.4	184.84	10.260	100.00
63	565.0	123.4	193.47	10.739	100.00
64	332.6	27.0	189.93	10.543	100.00
65	565.0	25.6	110.17	6.115	100.00

Appendix B. Assumptions and results of the economic analysis

Table B1 Assumptions for the estimation of the capital costs

Capital Costs	Reference Installed Cost [M€ 2017]	Scale parameter	Reference size	Exponent	Installation cost factor	Ref
Gas production and processing						
NG purification and pre-reforming	Data from Ascent project partners (ASCENT, 2014)					
Ca/Cu reactors	Data from Ascent partners (ASCENT, 2014)					
ATR	Data from Ascent partners (ASCENT, 2014)					
Gas pre-heaters	Direct calculation				2-4	(AspenTech, 2016)
Water saturator	Direct calculation				-	(AspenTech, 2016)
WGS reactors	Data from Ascent project partners (ASCENT, 2014)					
Gas-Heated Pre-reformer	Data from Ascent project partners (ASCENT, 2014)					
CO ₂ removal and compression						
CO ₂ Removal system (MDEA)	79.05	CO ₂ mass flow rate	62.59 kg/s	0.80	-	(Rath, 2010)
CO ₂ Compression & Drying	15.81	CO ₂ compressor power	13.0 MW _e	0.67	1.6	(Franco et al., 2011)
Gas Turbine/Accessories						
Gas Turbine Generator	57.65	Flue gas mass flow rate	650.0 kg/s	0.67	1.1	(Franco et al., 2011)
Gas Turbine Foundations	1.97	Gas turbine power	422.0 MW _e	0.67	1.1	(Fout et al., 2015)
HRSG, Ducting & Stack						
Heat Recovery Steam Generator	Direct calculation				4	(AspenTech, 2016)
Steam Turbine Generator & steam cycle						
Steam TG & Accessories	36.63	Turbine power	200.0 MW _e	0.67	1.16	(Franco et al., 2011)
Turbine Plant Auxiliaries	2.12	Turbine power	200.0 MW _e	0.67	3.29	(Franco et al., 2011)
Condenser & Auxiliaries	5.49	Condenser Duty	373.6 MW _{th}	0.12	1	(US DOE, 2010)
Steam Piping	14.82	HRSG duty	559.71 MW _{th}	0.83	1.41	(Fout et al., 2015)
Steam TG Foundations	2.21	Turbine power	219.0 MW _{th}	0.73	2.65	(Fout et al., 2015)
Cooling water system & Feedwater						
Cooling water system	25.84	Thermal power discharged	470.0 MW _{th}	0.67	1.87	(Franco et al., 2011)

Feedwater & BOP	65.20	Thermal power discharged	470.0 MW _{th}	0.67	1.87	(Franco et al., 2011)
Accessory Electric plant						
GT Generator	3.71	GT electric power	272.1 MW _e	0.67	-	(Franco et al., 2011)
ST Generator	4.76	ST electric power	200.0 MW _e	0.67	-	(Franco et al., 2011)
Station Service Equipment	1.43	Net auxiliary load	11.0 MW _e	0.51	1.1	(Fout et al., 2015)
Switchgear & Motor Control	1.90	Net auxiliary load	11.0 MW _e	0.51	1.2	(Fout et al., 2015)
Conduit & Cable Tray	3.29	Net auxiliary load	11.0 MW _e	0.51	3.9	(Fout et al., 2015)
Wire & Cable	4.27	Net auxiliary load	11.0 MW _e	0.51	1.6	(Fout et al., 2015)
Protective Equipment	2.78	Net auxiliary load	11.0 MW _e	0.58	4.5	(Fout et al., 2015)
Standby Equipment	0.21	Gross total power input	641.0 MW _e	0.47	1.9	(Fout et al., 2015)
Main Power Transformers	10.92	Net electric power input	712.22 MVA	0.62	1.0	(Fout et al., 2015)
Electrical Foundations	0.50	Gross total power input	641.0 MW _e	0.70	3.5	(Fout et al., 2015)

Table B2 Breakdown of Indirect Costs. Table adapted from Franco et al. (2011)

	[% of TDPC]
Yard improvement	1.5 %
Service facilities	2.0 %
Engineering & Consultancy	4.5 %
Building	4.0 %
Miscellaneous	2.0 %

Table B3 Assumptions for the calculation of the reactors initial fills

	Initial fill cost [M€2017]	Lifetime [years]
Hydrogenator catalyst	0.282	10
ZnO absorbent	0.788	2
ATR reforming catalyst (ATR benchmark plant)	1.360	6
High Temperature Shift (ATR benchmark plant)	1.360	6
Low Temperature Shift (ATR benchmark plant)	2.824	6
Heated pre-reformer catalyst (ATR benchmark plant)	0.502	4
Pre-reformer Ca/Cu-Stage A	3.943	4
Pre-reformer Ca/Cu-Stage C	1.611	4

Table B4 Cost assumptions on annual maintenance cost (given as % of TDPC)

Plant section	Maintenance cost [% of TDPC]
Power island, CO ₂ removal & compression	2.50%
Steam Turbine and Accessories	5.00%
Steam Turbine Generator	2.00%
Balance of plant	1.70%
Reactors and other components	1.50%

Table B5 Consumable cost assumptions

Consumable	Cost	Ref.	Consumable	Cost	Ref.
Natural gas	6.5 €/GJ	(Franco et al., 2011)	SCR catalyst	4760 €/m ³	(Fout et al., 2015)
Raw water	0.35 €/m ³	(Franco et al., 2011)	Ammonia (19% NH ₃)	107 €/m ³	(Fout et al., 2015)
Process demi-water	6.04 €/m ³	(Franco et al., 2011)	MDEA solvent	1890 €/m ³	(US DOE, 2010)

Table B6 Long-run profitability analysis assumptions

Availability	7500 h/y
Operational life	25 y
Construction time	3 y
Capital expenditure during construction: Year -3	40%

Year -2	30%
Year -1	30%
Tax rate	Pre-taxation
Depreciation time	No depreciation
Inflation rate	0 %
Construction inflation	0 %
Interest rate	8 %

Table B7 Gas production and processing capital cost breakdown for the ATR+MDEA NGCC power plant

Cost item	M€	€/kW _e
ATR	6.48	17.35
Gas-heated pre-reformer	17.57	47.07
Gas-gas regenerative HX	0.53	1.42
Syngas cooler-HP boiler 1	3.16	8.46
Syngas cooler-HP boiler 2	1.03	2.77
Syngas cooler-HP economizer	1.36	3.63
Syngas cooler-gas heater	0.09	0.25
Syngas cooler-water heater	0.10	0.26
Regenerative NG pre-heater	0.31	0.82
Water-heated NG pre-heater 1	0.42	1.13
Water-heated NG pre-heater 2	0.03	0.08
Saturator water heater 1	0.23	0.60
Saturator water heater 2	0.30	0.81
HT WGS	1.98	5.30
LT WGS	2.12	5.69
ZnO	0.30	0.80
Hydrogenator	0.14	0.38
Saturator	0.44	1.17
Total	36.58	97.98

Table B8 Detail of the calculated valves cost for three cases of the NGCC power plant integrated with the Ca-Cu process

Valve cost [k€]	Base case	Compact reactor	Optimized case
Valve in A'	105	105	105
Valve out A'	289	289	289
Valve in A	127	127	127
Valve out A	268	268	268
Valve in B	80	83	72
Valve out B	31	31	26
Valve in B'	116	116	100
Valve out B'	167	157	133
Valve in C	78	90	90
Valve out C	54	54	54

Table B9 Main characteristics of heat exchangers of the NGCC integrated with Ca-Cu process for the 'optimized case'

	Heat Duty [MW]	LMTD [°C]	Area* [m ²]	U* [W/ m ² /K]	Weight [t]	Cost [M€]
Shell & tubes heat exchangers						
Evaporator G	42.3	28.0	2843	534	222	5.4
Evaporator H	133.1	66.8	3322	615	131	7.3

LT Superheater H	44.1	159.6	6268	198	97	8.9
HX D	34.4	53.0	2336	280	10	0.6
HX E	29.5	124.9	504	470	7	0.3
Evaporator F	9.8	429.8	75	763	34	1.2
Superheater F	7.3	277.8	82	324	8	1.3
GT HRSG						
HP SH 2	53.4	40.3	5433	244	350	8.8
HP SH 1	82.6	24.5	10867	311	668	8.5
HP RH 2	22.0	36.0	3354	182	182	4.9
HP RH 1	41.1	32.9	6708	186	363	4.9
HP ECO 2	34.8	25.9	3874	346	226	2.4
HP ECO 1	113.3	17.2	20601	320	1267	12.6
LP ECO	53.5	44.1	5031	241	277	2.8
CO₂ rich-gas HRSG						
HX I	21.8	114.1	2042	93	98	2.6
HX J	7.3	58.7	614	202	33	0.5
HX K	19.9	70.6	2653	106	146	1.7
HP EVA	5.6	54.4	230	448	12	0.1
HP ECO 2	0.9	19.8	111	402	6	0.1
HP ECO 1	8.2	21.5	1044	367	55	0.6
LP ECO	3.4	38.2	258	339	14	0.1

**for the HRSG, Area and U are referred to the outer bare tube surface*

Table B10 Capital cost breakdown of the benchmark ATR+MDEA NGCC power plant

Cost item		M€	€/kW _e
Feedwater & Miscellaneous BOP Systems		42.68	114.31
Gas production and processing		36.58	97.98
CO ₂ Removal and Compression		73.91	197.98
Gas Turbine/Accessories*		67.00	179.47
HRSG, Ducting & Stack		49.94	133.77
Steam Turbine Generator		48.10	128.84
Cooling Water Systems		16.91	45.30
Accessory Electric Plant		34.91	93.51
	TDPC	370.03	991.16
Yard improvement		5.55	14.87
Service facilities		7.40	19.82
Engineering & consultancy		16.65	44.60
Building		14.80	39.65
Miscellaneous		7.40	19.82
	EPC	421.83	1129.92
Other Owner's costs		63.28	169.49
	TPC	485.11	1299.41
MDEA solvent		0.27	0.73
ATR reforming catalyst		1.36	3.64
High temperature shift catalyst		1.36	3.64
Low temperature shift catalyst		2.82	7.56
Heated pre-reformer catalyst		0.50	1.34
Hydrodesulphurization catalyst		0.29	0.77
ZnO absorbent		0.81	2.16
	TPC+LF	492.52	1319.27
* GT includes the booster air compressor			

Table B11 O&M costs of the benchmark ATR+MDEA NGCC power plant

Variable O&M		M€/year	€/MWh
Natural gas		145.83	52.08
Raw water		1.46	0.52
Process water		5.33	1.90
	Total	152.63	54.51
Fixed O&M			
Operating labor		5.29	1.89
Property taxes and insurance		9.70	3.47
Maintenance labor		3.47	1.24
Maintenance materials		5.21	1.86
Administrative & support labor		2.19	0.78
Chemicals and replacements			
<i>MDEA solvent</i>		1.71	0.61
<i>ATR reforming catalyst</i>		0.23	0.08
<i>High temperature shift catalyst</i>		0.23	0.08
<i>Low temperature shift catalyst</i>		0.47	0.17
<i>Heated pre-reformer catalyst</i>		0.13	0.04
<i>Hydrodesulphurization catalyst</i>		0.03	0.01
<i>ZnO absorbent</i>		0.39	0.14
<i>SCR catalyst</i>		0.08	0.03
<i>Ammonia (19% NH₃)</i>		0.15	0.05
	Total	29.26	10.45