CO_2-free coal-fired power generation by partial oxy-fuel and post-combustion CO_2 capture: techno-economic analysis

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Abstract

Among the carbon capture and storage (CCS) technologies suitable for power generation plants, partial oxy-combustion coupled with post combustion CO_2 capture is gaining interest, since such a hybrid configuration could allow to reduce the size and enhance the performance of post-combustion CO_2 capture by operating combustion with air enriched with oxygen and reducing the dilution of flue gas. Moreover, partial oxy-combustion is a potential candidate for the retrofit of existing steam plants because it could be based on an almost conventional boiler and requires a smaller CO_2 capture section.

This work presents the results of a comparative techno-economic analysis of a 1000 MW_{th} partial oxy-combustion plant based on an ultra-supercritical pulverized coal combustion power plant integrated with a post-combustion CO_2 capture system and geological storage in saline aquifer. In particular, plant performance is assessed by using simulation models implemented through Aspen Plus 7.3 and Gate Cycle 5.40 commercial tools, whereas economic performance are evaluated on the basis of the expected annual cash flow. The analysis shows that, for new plants, this hybrid approach is not feasible from the economic point of view and full oxy-combustion potentially remains the most profitable technology even if, in the short-term period, the lack of commercial experience will continue to involve a high financial risk.

Keywords: Carbon capture and storage; Partial oxy-combustion; CO_2 capture, Techno-economic analysis

Acronyms

ABS, absorption column; ASU, air separation unit; BEC, bare erected cost; BF, baghouse filters; CCS, carbon capture and storage; CCTS, carbon capture, transport and storage; CCU, carbon capture and utilization; COE, cost of electricity; CPU, CO_2 capture and purification unit; DES, regeneration (desorption) column; EOR, enhanced oil recovery; FGC, flue gas cleanup; FGD, flue gas desulphurization; HPT, high pressure turbine; HTX, heat exchanger; IPT, intermediate pressure turbine; LCOE, levelized cost of electricity; LHV, lower heating value; LPT, low pressure turbine; MEA, monoethanolamine; NETL, U.S. National Energy Technology Laboratory; RH, re-heater; SCR, selective catalytic reduction; SH, super-heater; USC, ultra-supercrytical pulverized coal combustion; VAT, value added tax; VHPT, very high pressure turbine.

1. Introduction

The increase of the atmospheric CO_2 concentration has led to several environmental issues, notably an increase in global temperatures commonly referred to as global warming [1-3]. In this context, carbon capture and storage (CCS) and carbon capture and utilization (CCU) technologies must play a key role for its mitigation [4,5].

In general, CO_2 capture technologies can be classified according to three main approaches: (1) post-combustion, (2) pre-combustion, (3) oxy-fuel combustion [4,5]. In the post-combustion approach, fossil fuels are burned (as in conventional power plants) and then the CO_2 is captured from the flue gas. In the pre-combustion approach, the fossil fuel is gasified and the produced syngas is treated in a water-gas shift reactor to convert CO and water vapour into H_2 and CO_2 [6,7]. The latter is captured, while the hydrogen-rich syngas feeds a combined cycle plant for power generation. The oxy-fuel approach utilizes pure or nearly pure oxygen for combustion, such that primarily CO_2 and H_2O are produced by the process [5,8]. All these approaches are characterized by very high energy penalties: the plant net efficiency could be reduced of about 8-12 percentage points in case of post-combustion processes (mainly due to solvent regeneration) [9,10], and of 7-10 percentage points in case of pre-combustion approach [10]. Based on the state-of-the-art of a supercritical pulverized coal power plant, the efficiency losses related to oxy-fuel combustion are in the range of 9-13 percentage points [11], but it is likely that they can be reduced to 7-11 percentage points by means of processes optimization and heat integration [12]. So, oxy-fuel approach promises to become more and more interesting for future applications [13].

Overall, the very high cost of CCS technologies and the lack of experience in industrial-scale units are the key issues that are limiting the commercial application of the technologies. Therefore, today, the only full size CCS application in the world is represented by the Boundary Dam Carbon Capture Project in Estevan town (Saskatchewan, Canada), where the captured CO_2 is transported by pipeline (for 66 km) and injected for enhanced oil recovery (EOR) at the SaskPower's Weyburn oil field [14-16].

The main drawback for the large-scale deployment of oxy-combustion is the high energy consumption for pure O_2 production in the air separation unit (ASU), which causes a significant energy penalty [17].

One of the proposed solutions for short-term commercial applications is a compromise between post- and oxy-combustion approaches, a hybrid configuration commonly called partial oxy-fuel or partial oxy-combustion [18]. Primary fuel is burned in an oxygen-enriched environment in order to reduce the dilution of flue gas by nitrogen, thus enhancing the CO_2 concentration. The ASU for oxygen separation is smaller (which means a lower incidence in terms of capital cost and energy penalty) than the same equipment required by the oxy-combustion and the flue gas recirculation requires minor modifications on conventional boilers; in parallel, thanks to the less dilution by nitrogen, the volume of flue gas to be treated is significantly lower and CO_2 partial pressure is higher than in conventional post-combustion processes [19].

One of the first studies on the application of partial oxy-combustion for the retrofit of power plants has been published in 2009 by Doukelis et al. [18] and presents the so-called ECO-Scrub scheme as a good compromise between post-combustion capture and oxy-fuel. One of the key issues regarding the optimization of a partial oxy-combustion process is related to the definition of the optimal O_2 concentration in the enriched air. The specific effect of O_2 enrichment in amine-based chemical absorption has been studied by Lawal et al. [20,21], whereas Vega et al. [19,22,23] have presented an experimental study on monoethanolamine (MEA) degradation in partial-oxy-combustion CO_2 capture. Other post-combustion CO_2 capture technologies, such as membranes [24,25], calcium looping [17] and cryogenic separation [26], have been considered for the potential application in partial oxy-combustion scheme. Unfortunately, a lack of publications on the effect of oxygen concentration on plant efficiency and economic performance in partial oxy-combustion CO_2-free coal-fired power generation plants can be observed. Only Huang et al. (2012) [26] present an interesting techno-economic parametric analysis on hybrid coal-fired power plants (intended as oxy-fuel unit with a variable air dilution – up to 50% – and based on a cryogenic post-combustion CO_2 capture system). Finally, the same approach has been used in several applications in the cement industry, but with different techno-economic performance [27].

This work, starting from a comparative techno-economic assessment between post- and oxy-combustion technologies previously published by the authors [28,29], aims to extend the analysis to partial oxy-combustion in order to evaluate if the technology could be feasible for commercial applications. In particular, with the aim to compare conventional air-blown coal-fired steam power plants with full and partial oxy-combustion units, a detailed techno-economic analysis of an ultra-supercritical (USC) steam power plant equipped with CCS is carried out by varying oxygen concentration in the oxidant agent from about 21% (conventional air-blown combustion) to 95% (full oxy-fuel).

Performance evaluation has been carried out through simulation models based on the Aspen-Plus and Gate-Cycle commercial tools [30,31]. In particular, Gate-Cycle models are used to simulate the steam power plant in both air-blown and oxy-fuel arrangements, whereas Aspen-Plus models are used to simulate the conditioning and purification processes of exhaust gas and the air separation unit (ASU) process.

2. Plant configurations

As the main aim of this study is to make a techno-economic comparison between post-combustion, full and partial oxy-combustion approaches, the study considers, for each plant configuration, the same coal chemical power input of 1000 MW and the same USC power generation unit, equipped with a conventional flue gas cleanup (FGC) section and a low temperature CO_2 removal section, based on a chemical absorption process with an aqueous solution of MEA. To match CO_2 transport and storage requirements, the CO_2 removal section is also integrated with a conditioning and compression section to provide a high pressure (11 MPa) and high purity (CO_2 fraction of 99.7% by volume) CO_2 flow. Moreover, each plant configuration is considered to be fed with a commercial coal, whose main characteristics (lower heating value – LHV – proximate and ultimate analysis) are reported in table 1.

Proximate Analysis (% by weight)	
Fixed carbon	52.70
Volatile matter	25.90
Ash	14.40
Moisture	7.00
Ultimate Analysis (% by weight)	
Total carbon	65.66
Hydrogen	3.64
Sulphur	0.85
Nitrogen	1.61
Oxygen	6.84
Ash	14.40
Moisture	7.00
Heating value (MJ/kg)	
Lower heating value	25.03

Table 1. Reference coal properties (as received basis).

A conceptual scheme of each configuration is reported in figure 1.

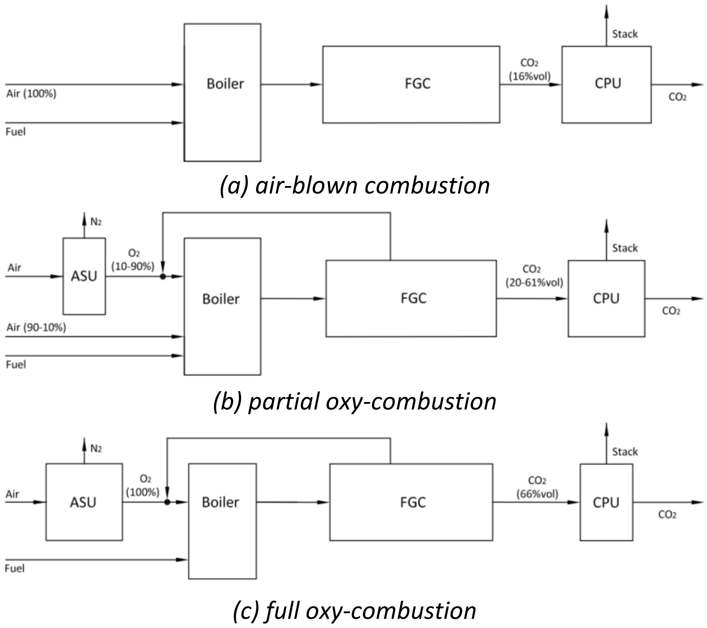


Figure 1. Conceptual scheme of the three configurations.

2.1. Air-blown configuration and USC steam cycle

The reference air-blown plant configuration considered in this paper is a typical medium-size USC power plant.

2.1.1. Steam cycle

According to the current state-of-the-art, the plant is based on a superheated and double reheat steam cycle with ten regenerative steam extractions. The double reheat requires higher capital costs, due to a higher complexity of the boiler and of the expansion train and to a more complex ducting system. On the other hand, it allows for a substantial increase of plant efficiency (in the order of 1 percentage point) in comparison to single reheat [32]. Moreover, double reheat leads to a higher steam quality at the outlet of the low-pressure turbine, thus increasing isentropic efficiency of the last stages.

Due to the presence of the double reheat, the selected configuration includes four steam turbines: a very high-pressure turbine (VHPT), a high-pressure turbine (HPT), an intermediate pressure turbine (IPT) and a low-pressure turbine (LPT). Figure 2 shows a simplified scheme of the air-blown USC power plant, whereas the main operating parameters assumed for the simulation models are reported in tables 2 and 3.

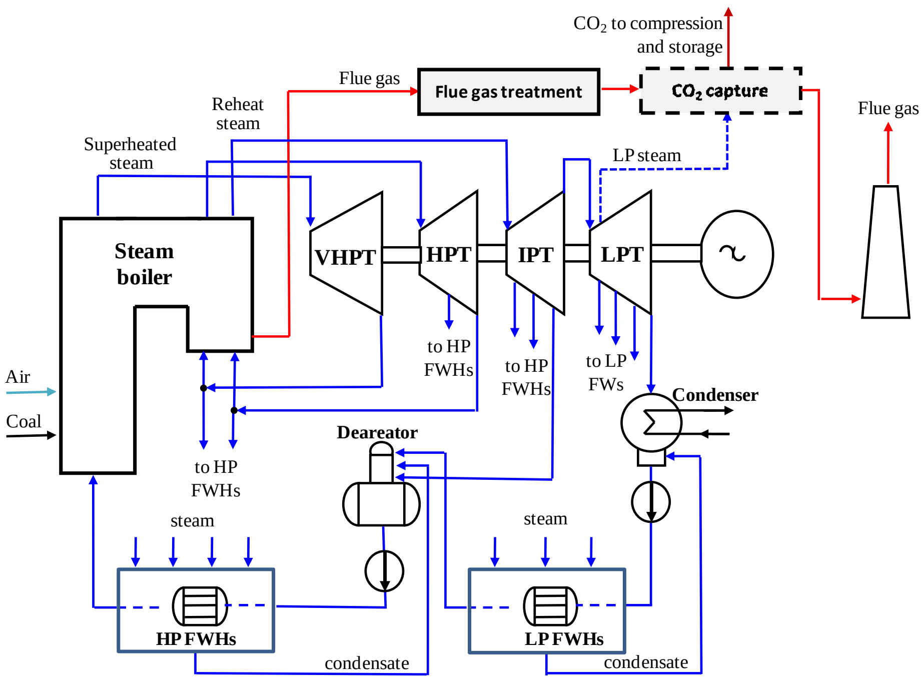


Figure 2. Simplified scheme of the air-blown USC plant.

Coal chemical power input (MW)	1000
SH/RH1/RH2 steam temperatures (°C)	600/620/620
SH/RH1/RH2 steam pressures (MPa)	30.0/13.5/5.4
Cycle maximum pressure (boiler feedwater pump) (MPa)	33.5
Cycle minimum pressure (condenser) (kPa)	4.2
Deaerator pressure (MPa)	0.8
Electric generator efficiency	0.99
BOP loss as steam turbine power fraction	0.03
High/low pressure heat exchangers minimum ΔT (°C)	-1.5/1.5

Table 2. Main USC operating parameters.

	VHPT	*HPT*	*IPT*	*LPT*
Inlet pressure (MPa)	30.0	13.5	5.4	0.5
Outlet pressure (MPa)	14.3	5.7	0.5	0.0042
Steam extractions	1	2	3	4
Turbine isentropic efficiency	0.92	0.94	0.94	0.89

Table 3. Main steam turbines operating parameters.

VHPT and HPT expansion ratios (about 0.48 and 0.42, respectively) have been chosen in order to maximize the efficiency of the double reheat steam cycle [33]. A first steam extraction is performed at the VHPT output, whereas, in the order, 2, 3, 4 extractions are performed in the HPT, IPT and LPT respectively. The very high pressure of the first steam extraction (slightly lower than 15 MPa) allows to increase water temperature upstream of the economizer above 335 °C.

Steam extraction pressures are established, regardless of turbine functional and constructive constraints, in order to assure a similar temperature rise inside the feedwater heat exchangers.

2.1.2. Flue gas treatment systems

The flue gas exiting from the boiler is sent to a conventional flue gas cleanup (FGC) section. A high-dust FGC configuration has been assumed, including a selective catalytic reduction (SCR) denitrification system for NO_x removal, baghouse filters (BF) for particulate removal and a low temperature flue gas desulphurization (FGD) system for SO_x removal.

SCR section causes a flue gas pressure drop in the range of 5-10 kPa, leading to an electrical power requirement for driving the fans of about 1% of the overall plant generation [34].

Baghouse filters, are installed downstream of the air preheater at 120-180 °C and cause a flue gas pressure drop of about 1-2 kPa, assuring a removal efficiency higher than 99% [35].

FGD process operates at low temperature with a flue gas pressure drop in the range of 5-10 kPa, requiring an electrical power of about 1% of the overall plant generation [36]. Globally, such a section accounts for an overall electrical power consumption of about 9 MW, mainly due to fan requirements for pressure drop of flue gas. Electrical power accounts for about 2% of the gross plant power, penalizing the plant efficiency of about one percentage point.

2.1.3. CO_2 capture and compression

The study considers a conventional chemical absorption process operating at atmospheric pressure with MEA; as a matter of facts, despite of its high energy requirements, it is currently one of the most proven and widespread solvents [37,38].

Such a process allows a CO_2 removal efficiency of 90% [39,40], separating high-purity (92-93% by volume) CO_2, which is sent to the conditioning and compression section.

The performance analysis of the CO_2 removal process has been carried out under equilibrium conditions, leading to an acceptable approximation [41,42].

The model assumes a MEA concentration of 30% (by weight) and a CO_2/MEA molar ratio of 0.28.

The main assumptions and simulation results of the CO_2 removal process are reported in table 4.

CO_2 removal efficiency (%)	90.0
Flue gas mass flow at the absorber inlet (kg/s)	410.2
CO_2 molar fraction in flue gas at the absorber inlet	0.154
Solvent/gas mass ratio	4.53
Flue gas mass flow at the absorber outlet (kg/s)	347.4
CO_2 molar fraction in flue gas at the absorber outlet	0.017
Flue gas temperature at the absorber outlet (°C)	58.6
MEA concentration at the absorber inlet (%)	30
CO_2/MEA molar ratio at the absorber inlet	0.28
CO_2-lean solvent temperature at the absorber inlet (°C)	35.0
CO_2-rich solvent temperature at the absorber outlet (°C)	50.5
CO_2-rich solvent temperature at the desorber inlet (°C)	90.0
CO_2-lean solvent temperature at the desorber outlet (°C)	102.7
CO_2 mass flow (kg/s)	85.7
CO_2 molar fraction in stream to CO_2 compressors	0.924
Reboiler specific thermal energy (MJ/kg_{CO2})	3.72

Table 4. Main operating parameters and performance of the CO_2 removal section.

In order to obtain a removal efficiency of 90%, a solvent/gas mass ratio of about 4.5 and a reboiler specific thermal energy of 3.75 GJ per ton of removed CO_2 have been calculated. The flue gas from the CO_2 capture section is mainly composed by N_2 (about 78%, by volume), while the CO_2 concentration decreases from about 14% to about 1.5%. The CO_2-rich gas from the absorption section is compressed to the transport pressure (11 MPa). It has been assumed that the compression process takes place up to 8 MPa by three intercooled compressors in series and then through a pump. The substantial water condensation leads to an almost pure CO_2 flow (with a molar fraction over 99.5%), as required for transport and storage.

The CO_2 removal dramatically affects the plant performance. In particular, the thermal power required by the reboiler to desorb CO_2 is remarkable (about 320 MW) and it is supplied by a low-pressure (0.39 MPa) steam extraction carried out in the LPT, which notably affects the plant power output. Another significant energy consumption is the electrical power required by the CO_2 compression and pumping system (about 30 MW), whereas the power required by the fan of the decarbonization section is limited to about 3 MW.

2.1.4. CO_2 transport and storage

The high-pressure and almost pure CO_2 stream exiting from the conditioning and compression section must be transported to the site designed to carbon dioxide storage. Transport of CO_2 has become a key factor in CCS, fixing CO_2 characteristics in terms of purity and pressure suitable for transportation. A 25 km long pipeline has been assumed as transport mode to the geological storage site for the captured carbon dioxide. The injection in saline aquifers has been chosen as the storage option in this study representing one of the highest storage capacity solution [43].

2.2. Full oxy-combustion plant configuration

The oxy-combustion plant configuration is based on the same steam cycle of the air-blown plant. The main functional and constructive differences regard the boiler, the oxygen supplied by a cryogenic ASU and the flue gas management and clean-up. As a matter of fact, oxy-combustion leads to higher temperatures in comparison to air-blown boiler. Therefore, flue gas recirculation (in this case about 70%, at a temperature of 310 °C) is carried out to control the flame temperature [44] and to obtain a boiler heat transfer profile similar to the one in air-blown steam generators [26]. Flue gas contains mainly CO_2 and water vapour and a small amount of un-reacted oxygen and inert gases. Consequently, just a CO_2 purification unit is required to attain a high purity CO_2 stream, avoiding the post-combustion CO_2 capture and its strong energy penalty. However, a remarkable energy penalty is related to the ASU for oxygen production and to the CO_2 compression for transport and storage.

A simplified scheme of the full-oxy configuration is reported in figure 3.

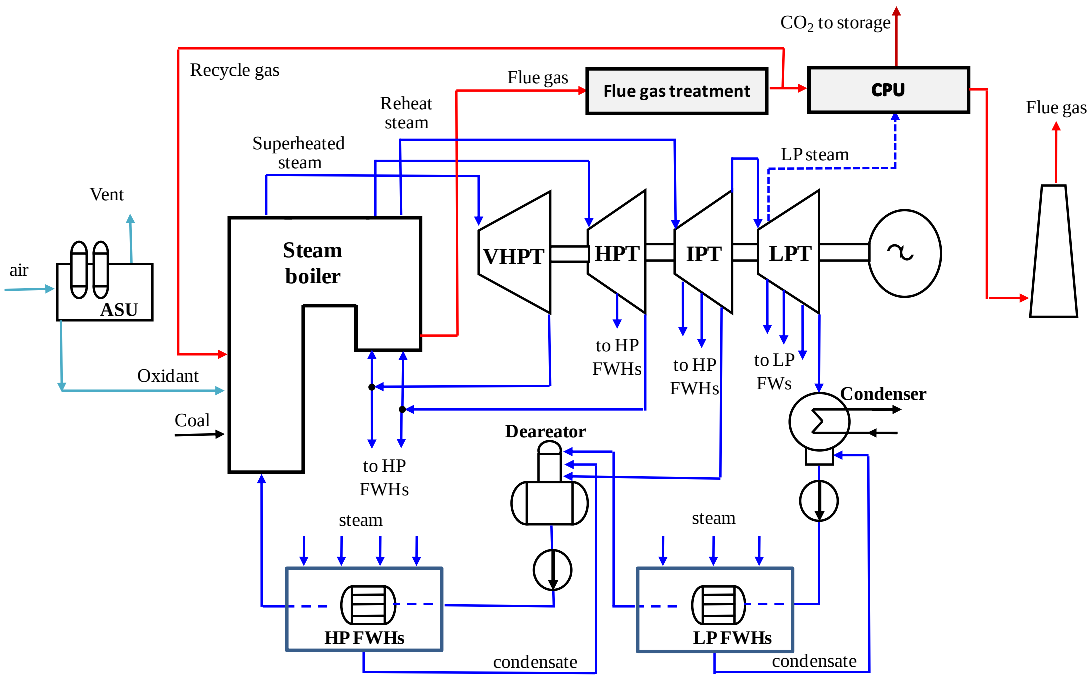


Figure 3. Simplified scheme of the full-oxy configuration.

The main operating parameters of the full-oxy configuration are reported in table 5.

Oxydant mass flow (kg/s)	85.26
$O_2/N_2/Ar$ molar fractions in oxydant	0.95/0.02/0.03
O_2 specific separation energy (kWh/t_{O2})	200.0
Flue gas recycle rate	0.684
Recycle gas mas flow (kg/s)	257.2
Recycle gas temperature (°C)	307.9

Table 5. Main operating parameters of the full-oxy configuration.

The power unit is equipped with a flue gas cleanup system similar to that used in the air-blown USC configuration, including SCR, BF and FGD systems. The high concentration of CO_2 in flue gas influences both $DeSO_x$ and $DeNO_x$ systems, but most of the studies assume that they can operate with better performance than in conventional steam plants [44]. Clean gas is mainly composed by CO_2 (about 66% by volume) and water vapour (about 26%), with small amounts of N_2

(3%), O_2 (2.5) and Ar (1.5%). The CO_2 could be easily separated by water condensation, but a CO_2 capture and purification unit (CPU) is still required to reduce the amount of oxygen and other incondensable gases and match the CO_2 purity requirements for transportation and storage [45]. In such a unit, the CO_2-rich gas is firstly cooled and compressed up to about 2.5 MPa with the condensation of a large amount of water. The CO_2-rich gas is cooled to -40 °C with the condensation of the largest portion of CO_2 and the separation of a considerable amount of incondensable gases. Then, the high-purity CO_2 stream is heated and sent to the second section of the compression train where the almost pure CO_2 gas is pressurized to transport and storage conditions (about 11 MPa). Conversely, the separated incondensable gases (N_2, O_2, Ar and residual CO_2) expand in a turbine to recover energy.

A larger CO_2 removal efficiency than that obtained with the post-combustion section has been calculated (about 94%) with a CO_2 purity of 96.4%. CO_2-rich gas is still composed by a smaller amount of N_2 (1.3%), O_2 (1.4%) and Ar (0.8%). The whole power requirement of the intercooled compression train is considerably higher than the one associated with the post-combustion section, due to the freezing unit (even if the compressors needs less energy due to the lower temperature of the treated stream). However, most of the CPU energy absorption is required for compression of CO_2, while a smaller amount is required for the separation of impurities.

2.3. Partial oxy-combustion configuration

The partial oxy-combustion configuration is a compromise between air-blown and full-oxy ones. Conceptually, an enrichment in oxygen of the combustion air involves a reduction of flue gas dilution by nitrogen. So, in the partial-oxy configuration, the boiler is fed with a mixture of atmospheric air (with an O_2 molar fraction of 0.206) and oxygen-rich gas (with a purity of 95%) produced by the ASU. Flue gas is characterized by a lower mass flow and by a higher CO_2 concentration in comparison with the air-blown configuration and it is treated by a similar (except for the size) high-dust FGC system. The configuration of the post-combustion CO_2 capture unit is the same considered in the air-blown case, but the higher CO_2 concentration involves better performance and a lower equipment size. Finally, the same compression system of the air-blown configuration has been considered for the partial-oxy approach.

3. Parametric analysis

A performance analysis has been carried out to assess the influence of air enrichment on plant performance and CO_2 removal, conditioning and compression processes. The increase of O_2 concentration in the oxidant involves a lower oxidant mass flow required for combustion, as shown in figure 4.

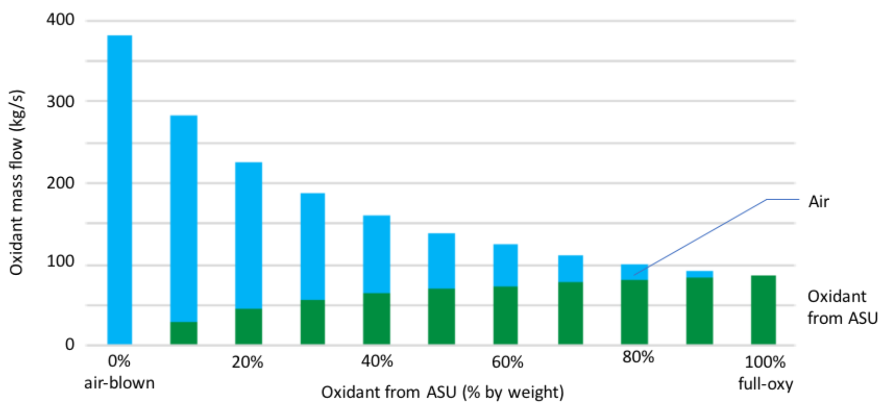


Figure 4. Whole oxidant mass flow as a function of oxidant from the ASU.

The air-blown plant configuration requires an air mass flow slightly higher than 380 kg/s, while the oxidant mass flow is reduced to about 85 kg/s with the full-oxy configuration. A 10% partial-oxy leads to an oxidant mass flow reduction of about 100 kg/s compared to the air-blown case. The mass flow of oxidant from ASU increases significantly for low air enrichment ratios (28.3 kg/s at 10% enrichment and 45 kg/s with 20% enrichment). The increase of ASU mass flow is moderate for major values, up to a maximum ASU production of about 85 kg/s for the full-oxy configuration. The reduction in the oxidant mass flow leads to a sensible decrease of the flue gas mass flow. A lower mass flow to be treated in the subsequent conditioning systems leads to a substantial reduction of the power requirements of the FGC section. Figure 5 shows how air enrichment, with the corresponding reduction of oxidant mass flow, involves a significant decrease of flue gas mass flow.

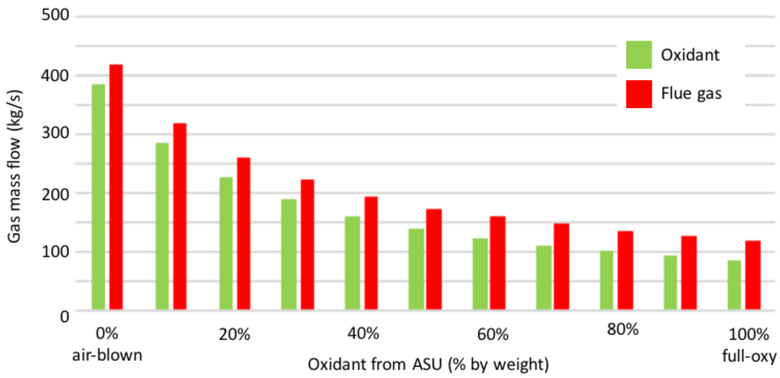
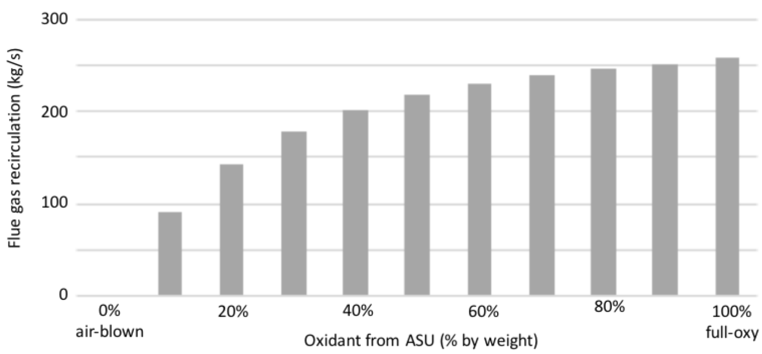


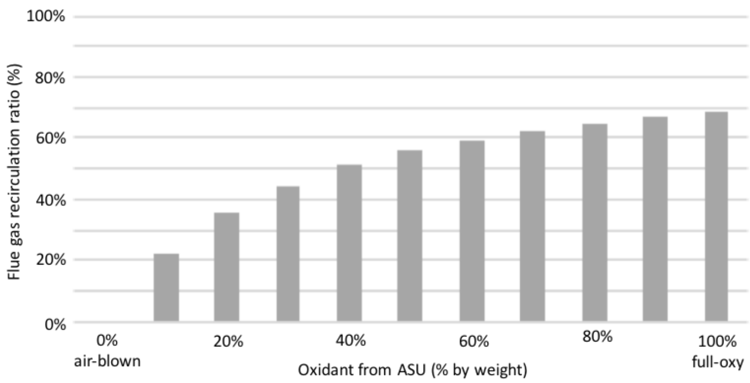
Figure 5. Reduction of the flue gas mass flow with air enrichment.

The air-blown plant configuration produces a flue gas mass flow slightly higher than 415 kg/s, while it is reduced to about 120 kg/s with the full-oxy configuration. The decrease of the flue gas mass flow is very pronounced at the lower values of the air enrichment: a 10% enrichment reduces flue gas mass flow to about 315 kg/s, 24,1% less than in the air-blown case.

The increase of oxygen content in the oxidant requires a greater gas recirculation to the boiler in order to control flame temperature. The mass flow of recirculated gas is calculated by imposing a constant maximum temperature inside the combustion chamber and is shown in figure 6.



(a) mass flow.



(b) recirculation ratio.

Figure 6. Flue gas recirculation and recirculation ratio.

A gas mass flow of about 260 kg/s is recirculated to the boiler in the full-oxy configuration. A 10% enrichment requires about 90 kg/s of gas recirculation, while a recirculated mass flow greater than 200 kg/s is required starting from a 40% enrichment. As a matter of fact, a higher fraction of recirculated flue gas corresponds to a greater mass flow of flue gas recirculated.

Figure 7 reports the mass flow of the main components of the flue gas at the reboiler exit, as a function of percentage of oxidant from ASU.

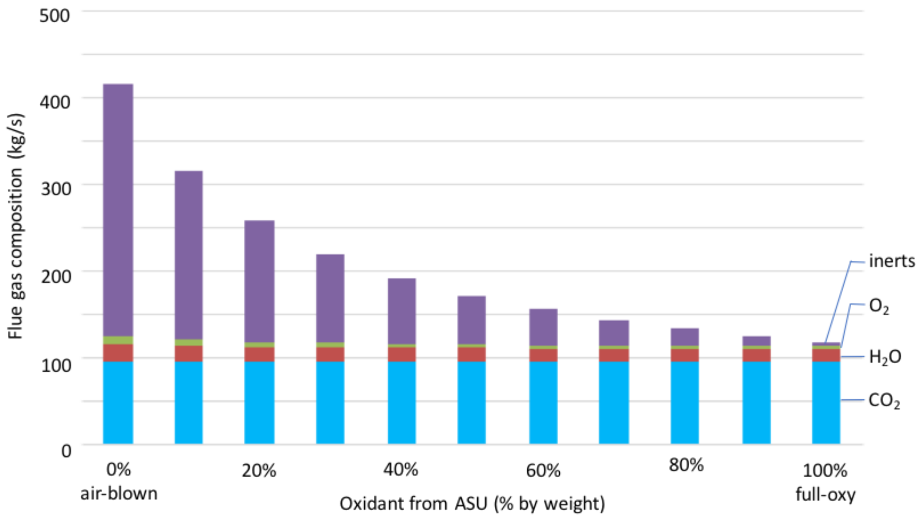


Figure 7. Composition of the flue gas.

The CO_2 content remains constant, only depending on coal feeding, but, due to the reduction of the gas flow, its concentration increases from 15.5% to 65.9% (by volume) as shown on figure 8, where the CO_2 molar fraction in the flue gas is reported as a function of the percentage of oxidant from ASU. The mass flow of inert gas (nitrogen and argon) is largely reduced increasing the air enrichment. Also, a slight reduction of water vapour and residual oxygen can be observed increasing air enrichment.

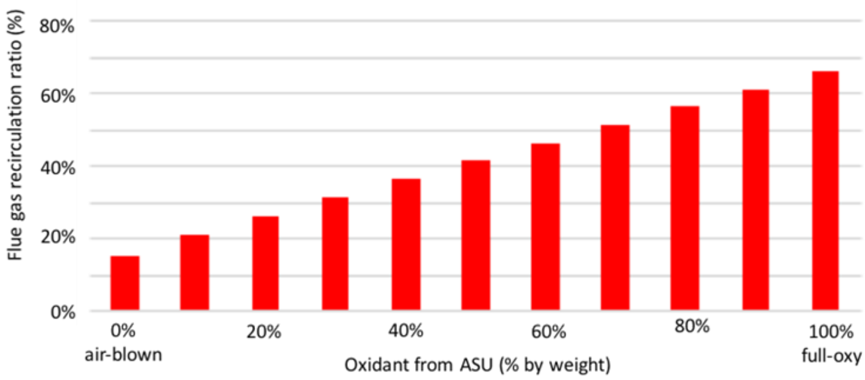


Figure 8. CO_2 concentration in flue gas.

A more concentrated flue gas improves solvent regeneration, slightly reducing thermal energy required in the reboiler, from a maximum value of about 3.75 GJ per ton of CO_2 removed (air-blown combustion) to a minimum value of about 3.50 GJ/t_{CO2} with a 90% enrichment. Despite a modest reduction of the specific thermal energy required by the reboiler, partial oxy-combustion enhances CO_2 removal process. In fact, the treatment of a flue gas with a more concentrated CO_2 greatly reduces the MEA degradation process [22].

4. Performance comparison

Table 6 summarizes the parametric performance assessment carried out through the simulation models with reference to the plant configurations previously described.

The reference (without CCS) air-blown plant shows a steam cycle output of about 500 MW. Auxiliaries (air fans, cooling water pumps, etc.) power absorptions, mechanical and generator losses reduce power output to about 475 MW. Considering the FGC section consumption, finally results a net power output slightly higher than 465 MW, leading to a net efficiency of 46.60%. The integration with the CO_2 removal section reduces the gross output of about 75 MW, mainly due to the large steam extraction from the steam turbine for solvent regeneration. This remarkable penalty combined with the power requirements of the CO_2 capture and compression section causes a noteworthy power output reduction slightly lower than 110 MW. Globally, CCS system reduces plant efficiency of 10.7 percentage points to 35.90%.

Full-oxy plant configuration shows a gross power output sensibly higher than the air-blown configuration with CCS (478.2 MW vs. 400.4 MW), due to the absence of steam extraction for solvent regeneration. However, the noteworthy power absorption of the ASU (more than 60 MW) and the high power requirement of the CPU unit lead to a net power output of about 360 MW and a net efficiency of 36.1%, very close to the air-blown case. On the other hand, such a configuration leads to CO_2 specific emissions (about 55 g/kWh) lower than those of air-blown CO_2-free one (about 95 g/kWh), thanks to a higher CO_2 removal efficiency (about 94.0%).

Partial oxy-combustion configurations present a gross power output comparable to that of the air-blown CO_2-free one (in the range 400-405 MW), but the net power output is dramatically reduced by the presence of the ASU. A net power output of about 340 MW has been calculated for a 10% enrichment, while the net power output is reduced to about 315 MW for a 90% enrichment. The lower power output associated to partial-oxy configurations leads to a slight increase (in the range of 100-110 g/kWh) of CO_2 specific emissions in comparison to air-blown configuration.

For comparative purposes, an annual availability of 7,600 hours has been arbitrarily assumed in this paper for all the considered configurations, despite oxy-fuel technology is still not commercially mature and the introduction of post-combustion CCS system could reduce the plant availability, due to the current poor experience in industrial-scale units.

Configuration	**ref. (no CCS)**	**air-blown**	**partial-oxy**	**partial-oxy**	**partial-oxy**	**partial-oxy**
Oxidant from ASU (% by weight)	0%	0%	10%	20%	30%	40%
O_2 concentr. in oxidant (% vol.)	20.56%	20.56%	27.29%	34.16%	41.18%	48.36%
Coal chemical power input (MW)	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0
- Steam turbines (MW)	*515.7*	*437.3*	*436.2*	*434.8*	*435.8*	*437.4*
- Pumps (MW)	*16.1*	*16.1*	*16.0*	*16.0*	*16.0*	*16.1*
Steam cycle output (MW)	499.6	420.9	420.2	418.8	419.8	421.3
- Aux. absorptions and mechanical losses (MW)	*19.4*	*16.1*	*16.3*	*16.4*	*16.6*	*16.7*
- Generator losses (MW)	*5.2*	*4.4*	*4.4*	*4.4*	*4.4*	*4.4*
Gross power output (MW)	475.0	400.4	399.5	398.0	398.8	400.2
- CGT section absorptions (MW)	*9.0*	*9.0*	*6.7*	*5.3*	*4.5*	*3.8*
- ASU (MW)	-	-	*20.4*	*32.4*	*40.4*	*46.0*
- CO_2 capture and compression (MW)	-	*32.4*	*31.7*	*31.3*	*31.1*	*30.9*
Net power output (MW)	466.0	359.0	340.7	329.0	322.8	319.5
Net efficiency (%)	46.60	35.90	34.07	32.90	32.28	31.95
Plant availability (h/year)	7600	7600	7600	7600	7600	7600
Energy production (GWh/year)	3541.6	2728.4	2589.3	2500.4	2453.3	2428.2
CO_2 emissions (Mt/year)	2.60	0.260	0.260	0.260	0.260	0.260
CO_2 specific emissions (g/kWh)	734.1	95.3	100.4	104.0	106.0	107.1

Table 6a. Overall performance of air-blown, full-oxy and partial-oxy plant configurations.

Configuration	**partial-oxy**	**partial-oxy**	**partial-oxy**	**partial-oxy**	**partial-oxy**	**full-oxy**
Oxidant from ASU (% by weight)	50%	60%	70%	80%	90%	100%
O_2 concentr. in oxidant (% vol.)	55.70%	63.20%	70.88%	78.73%	86.77%	95.00%
Coal chemical power input (MW)	1000.0	1000.0	1000.0	1000.0	1000.0	1000.0
- Steam turbines (MW)	*439.0*	*440.6*	*442.1*	*443.5*	*444.9*	*519.8*
- Pumps (MW)	*16.1*	*16.1*	*16.1*	*16.1*	*16.2*	*16.3*
Steam cycle output (MW)	422.9	424.5	426.0	427.4	428.7	503.5
- Aux. absorptions and mechanical losses (MW)	*16.8*	*16.9*	*17.0*	*17.0*	*17.1*	*20.1*
- Generator losses (MW)	*4.4*	*4.4*	*4.4*	*4.4*	*4.4*	*5.2*
Gross power output (MW)	401.7	403.2	404.6	406.0	407.2	478.2
- CGT section absorptions (MW)	*3.4*	*3.0*	*2.7*	*2.5*	*2.3*	*2.1*
- ASU (MW)	*50.2*	*53.4*	*56.0*	*58.1*	*59.9*	*61.4*
- CO_2 capture and compression (MW)	*30.8*	*30.7*	*30.6*	*30.5*	*30.4*	*53.4*
Net power output (MW)	317.3	316.1	315.3	314.9	314.6	361.3
Net efficiency (%)	31.73	31.61	31.53	31.49	31.46	36.13
Plant availability (h/year)	7600	7600	7600	7600	7600	7600
Energy production (GWh/year)	2411.5	2402.4	2396.3	2393.2	2391.0	2745.9
CO_2 emissions (Mt/year)	0.260	0.260	0.260	0.260	0.260	0.156
CO_2 specific emissions (g/kWh)	107.8	108.2	108.5	108.6	108.7	56.8

Table 6b. Overall performance of air-blown, full-oxy and partial-oxy plant configurations.

5. Cost evaluation

The economic and financial assessment of the whole CCS project at different oxygen concentrations in the oxidant agent has been carried out on the basis of the levelized cost of electricity (LCOE) and other economic indicators. The study has been carried out by using a detailed economic model and considering the year 2017 as the starting year of the project. This assumption allows to compare the economic results on partial oxy-combustion configurations with the results on post- and oxy-combustion, previously published by the authors [29].

5.1. Project's milestones and financial assumptions

The economic analysis is based on several key assumptions. First of all, the investment is distributed in the four years of the construction phase (24%, 39%, 32% and 5%), starting from the year 2017 [46], and the whole operating life of the project is assumed 25 years (2021 to 2045). The study is based on the realistic assumption that 80% of the investment for plant construction is supported by the banks through the opening of a senior debt (with a financing fee of 2.5% and a constant annual interest rate of 6.14% in 10 years), whereas the remaining 20% is directly provided by the owner company. A value added tax (VAT) of 22% is assumed for both capital and operating costs [47]. An amortization rate of 10% has been assumed for both the power generation and the CCS systems, whereas a rate of 14% is considered for the material handling system [47]. The model also considers a yearly extra investment during the operation of the plant [46].

Finally, the calculation of the present values is based on an assumed annual discount rate of 8% [26].

5.2. Capital and operating costs estimation

Capital costs of each component are assessed on the basis of industrial data recently published by the U.S. National Energy Technology Laboratory (NETL) [48,49], following the same approach widely described in Pettinau et al. 2017 [29]. In addition, the following assumptions have been taken for the boilers components: (i) the cost of the air-blown boiler is the same as reported in [29], with an extra cost of 50 €/kW [50] to consider the second reheat; (ii) the cost of the full-oxy boiler has been calculated from the air-blown one, with an extra cost of 7% [51] to consider the different operating conditions; (iii) the costs of the boilers for partial-oxy configurations are calculated through a linear variation between air-blown and full-oxy configurations, on the basis of oxygen enrichment. Moreover, the full-oxy configuration considers a cryogenic CO_2 separation system, whose cost (including CO_2 compression) has been calculated as 10.3% of the bare erected cost [26].

Capital cost of all the considered plant configurations are summarized in table 7.

Configuration	ref. (no CCS)	air-blown	partial-oxy	partial-oxy	partial-oxy	partial-oxy
Oxidant from ASU (% by weight)	0%	0%	10%	20%	30%	40%
O_2 concentr. in oxidant (% vol.)	20.56%	20.56%	27.29%	34.16%	41.18%	48.36%
Coal and sorbents handling	27,727.47	27,727.47	27,727.47	27,727.47	27,727.47	27,727.47
Coal & sorbents prep. and feed	13,247.26	13,247.26	13,247.26	13,247.26	13,247.26	13,247.26
Feedwater and balance of plant	58,791.58	58,791.58	58,791.58	58,791.58	58,791.58	58,791.58
Air sep. unit and accessories	0.00	0.00	134,032.61	176,980.01	201,814.28	218,273.97
Boiler and accessories	244,696.10	244,696.10	246,408.97	248,121.85	249,834.72	251,547.59
Gas cleanup and piping	109,478.37	109,478.37	106,194.02	102,909.67	99,625.31	96,340.96
CO_2 removal system	0.00	241,374.03	204,942.97	181,469.89	164,846.07	152,359.66
CO_2 compression and drying	0.00	49,996.52	49,996.52	49,996.52	49,996.52	49,996.52
CO_2 transport	0.00	26,691.55	26,691.55	26,691.55	26,691.55	26,691.55
CO_2 injection infrastructure	0.00	322,772.72	322,772.72	322,772.72	322,772.72	322,772.72
Ducting and stack	24,267.41	24,267.41	20,604.68	18,244.73	16,573.39	15,318.03
Steam turbine generator	109,041.51	109,041.51	108,896.83	108,634.70	108,652.65	108,760.30
Cooling water system	37,497.59	37,497.59	37,447.84	37,357.69	37,363.87	37,400.89
Ash & spent sorbent handling	10,522.41	10,522.41	10,311.96	10,101.51	9,891.06	9,680.61
Other auxiliaries	114,280.06	114,280.06	115,172.02	116,063.97	116,955.93	117,847.88
Bare erected cost (BEC)	749,549.75	1,390,384.57	1,483,238.98	1,499,111.11	1,504,784.37	1,506,756.98
Engineering and commissioning	74,954.98	139,038.46	148,323.90	149,911.11	150,478.44	150,675.70
Contingencies	98,225.44	225,283.45	231,250.08	230,267.69	228,977.33	227,768.99
Total plant cost (TPC)	922,730.16	1,754,706.48	1,862,812.96	1,879,289.91	1,884,240.13	1,885,201.67
Financing fees	20,806.54	39,566.75	42,004.44	42,375.98	42,487.60	42,509.28
Interests	148,235.63	281,892.13	299,259.37	301,906.38	302,701.63	302,856.10
Total as-spent cost (TASC)	1,091,772.34	2,076,165.36	2,204,076.77	2,223,572.27	2,229,429.36	2,230,567.06
Specific TPC (€/kW net)	1,980.02	4,887.49	5,467.54	5,712.50	5,838.43	5,900.64

Table 7a. Capital costs estimation (in k€).

Configuration	partial-oxy	partial-oxy	partial-oxy	partial-oxy	partial-oxy	full-oxy
Oxidant from ASU (% by weight)	50%	60%	70%	80%	90%	100%
O_2 concentr. in oxidant (% vol.)	55.70%	63.20%	70.88%	78.73%	86.77%	95.00%
Coal and sorbents handling	27,727.47	27,727.47	27,727.47	27,727.47	27,727.47	27,727.47
Coal & sorbents prep. and feed	13,247.26	13,247.26	13,247.26	13,247.26	13,247.26	13,247.26
Feedwater and balance of plant	58,791.58	58,791.58	58,791.58	58,791.58	58,791.58	58,791.58
Air sep. unit and accessories	230,009.51	238,864.36	245,745.41	251,273.72	255,808.04	260,179.95
Boiler and accessories	253,260.46	254,973.34	256,686.21	258,399.08	260,111.96	261,824.83
Gas cleanup and piping	93,056.61	89,772.26	86,487.91	83,203.56	79,919.21	76,634.86
CO_2 removal system	142,560.39	134,673.96	128,139.25	122,645.58	117,938.10	106,195.71
CO_2 compression and drying	49,996.52	49,996.52	49,996.52	49,996.52	49,996.52	49,996.52
CO_2 transport	26,691.55	26,691.55	26,691.55	26,691.55	26,691.55	26,691.55
CO_2 injection infrastructure	322,772.72	322,772.72	322,772.72	322,772.72	322,772.72	322,772.72
Ducting and stack	14,332.82	13,539.93	12,882.94	12,330.61	11,857.33	11,465.30
Steam turbine generator	108,891.31	109,026.36	109,157.16	109,275.48	109,381.35	109,477.54
Cooling water system	37,445.94	37,492.38	37,537.36	37,578.05	37,614.46	37,647.53
Ash & spent sorbent handling	9,470.17	9,259.72	9,049.27	8,838.82	8,628.37	8,417.92
Other auxiliaries	118,739.83	119,631.79	120,523.74	121,415.70	122,307.65	123,199.61
Bare erected cost (BEC)	1,506,994.15	1,506,461.18	1,505,436.35	1,504,187.70	1,502,793.56	1,494,270.33
Engineering and commissioning	150,699.42	150,646.12	150,543.64	150,418.77	150,279.36	149,427.03
Contingencies	226,686.30	225,743.81	224,905.94	224,165.39	223,500.82	218,703.92
Total plant cost (TPC)	1,884,379.87	1,882,851.11	1,880,885.93	1,878,771.86	1,876,573.74	1,862,401.28
Financing fees	42,490.75	42,456.28	42,411.97	42,364.30	42,314.73	41,995.16
Interests	302,724.08	302,478.48	302,162.78	301,823.15	301,470.03	299,193.23
Total as-spent cost (TASC)	2,229,594.70	2,227,785.88	2,225,460.67	2,222,959.31	2,220,358.50	2,203,589.67
Specific TPC (€/kW net)	5,938.55	5,956.67	5,964.90	5,967.41	5,965.16	5,052.60

Table 7b. Capital costs estimation (in k€).

It can be firstly observed that the introduction of the ASU involves a significant increase in BEC, in spite of the small reduction of CO_2 capture section cost.

For comparative purposes, the same assumptions reported in [29] have been used in this work for fuel purchasing, operating and maintenance (reported in table 8), eco-taxes and CO_2 emission allowances (the latter based on a market price of 23 €/t by 2020, according to an assessment published by Thomson Reuters [52]).

Configuration	ref. (no CCS)	air-blown	partial-oxy	partial-oxy	partial-oxy	partial-oxy
Oxidant from ASU (% by weight)	0%	0%	10%	20%	30%	40%
O_2 concentr. in oxidant (% vol.)	20.56%	20.56%	27.29%	34.16%	41.18%	48.36%
Labor	6.69	8.51	8.25	7.99	7.73	7.47
Maintenance materials	7.32	9.23	8.75	8.26	7.78	7.29
Consumables	0.99	1.69	1.29	1.05	0.90	0.78
Waste disposal & by-products	10.38	13.40	11.87	10.98	10.39	9.97
Total O&M	25.38	32.83	30.15	28.28	26.79	25.52

Table 8a. Operating and maintenance costs (in €/MWh).

Configuration	partial-oxy	partial-oxy	partial-oxy	partial-oxy	partial-oxy	full-oxy
Oxidant from ASU (% by weight)	50%	60%	70%	80%	90%	100%
O_2 concentr. in oxidant (% vol.)	55.70%	63.20%	70.88%	78.73%	86.77%	95.00%
Labor	7.21	6.94	6.68	6.42	6.16	5.90
Maintenance materials	6.81	6.33	5.84	5.36	4.87	4.39
Consumables	0.70	0.64	0.59	0.55	0.51	0.48
Waste disposal & by-products	9.66	9.42	9.23	9.07	8.94	8.83
Total O&M	24.38	23.33	22.34	21.40	20.49	19.61

Table 8b. Operating and maintenance costs (in €/MWh).

Finally, the CO_2 compression and transport costs have been assumed equal to 0.75 c€/kg and 2.5 c€/(t km), respectively [53,54], whereas an operating cost of 0.3 €/t has been considered for sequestration in saline aquifer [54].

6. Economic assessment

Table 9 shows a summary of the economic performance of the air-blown, partial-oxy and full-oxy configurations. A detailed definition of all the economic indicators can be found in Pettinau et al. 2017 [29].

Configuration	ref. (no CCS)	air-blown	partial-oxy	partial-oxy	partial-oxy	partial-oxy
Oxidant from ASU (% by weight)	0%	0%	10%	20%	30%	40%
O_2 concentr. in oxidant (% vol.)	20.56%	20.56%	27.29%	34.16%	41.18%	48.36%
Cost of electricity (€/MWh)	104.16	134.51	136.61	137.04	136.55	135.54
LCOE, present values (€/MWh)	40.06	60.39	63.20	64.25	64.56	64.48
CO_2 capture cost (€/t)	n.a.	44.13	40.35	36.98	34.54	32.59

CO_2 capture cost, present (€/t)	n.a.	24.16	23.90	22.77	21.91	21.23
CO_2 avoidance cost (€/t)	n.a.	60.07	54.78	50.06	46.64	43.88
CO_2 avoidance cost, present (€/t)	n.a.	32.89	32.45	30.83	29.59	28.59

Table 9a. Summary of economic performance.

Configuration	partial-oxy	partial-oxy	partial-oxy	partial-oxy	partial-oxy	full-oxy
Oxidant from ASU (% by weight)	50%	60%	70%	80%	90%	100%
O_2 concentr. in oxidant (% vol.)	55.70%	63.20%	70.88%	78.73%	86.77%	95.00%
Cost of electricity (€/MWh)	134.43	133.18	131.89	130.60	129.28	114.88
LCOE, present values (€/MWh)	64.27	63.94	63.57	63.17	62.75	55.21
CO_2 capture cost (€/t)	30.91	29.40	28.00	26.68	25.42	22.81
CO_2 capture cost, present (€/t)	20.64	20.13	19.65	19.20	18.77	17.80
CO_2 avoidance cost (€/t)	41.52	39.41	37.46	35.63	33.88	26.29
CO_2 avoidance cost, present (€/t)	27.73	26.98	26.29	25.64	25.02	20.51

Table 9b. Summary of economic performance.

First of all, it is important to underline that LCOE for air-blown and full-oxy configurations are lower than the corresponding values obtained by the authors in a previous work (63.4 and 62.8 €/MWh, respectively) [29]. The differences are due to the improvement of the steam cycle (a single reheat has been considered in the previous work) and, for full-oxy configuration, also to the higher annual availability of the plant (7,600 h/yr. vs. 7,000 h/yr. considered in the previous work).

The analysis of both cost of electricity (COE) and LCOE shows that, for new plants, full oxy-combustion is the more promising among the considered CO_2-free power generation technologies, allowing for a significant reduction of LCOE with respect to conventional air-blown plants with post-combustion capture (55 €/MWh vs 60 €/MWh). On the other hand, partial oxy-combustion could be competitive, in terms of COE, with respect to air-blown plants (mainly with an oxygen enrichment higher than 40-50%), but it presents a higher COE than the full oxy-combustion technology. Considering that the full oxy-combustion is still quite far from commercial application (due to the relatively low experience on commercial-scale), partial oxy-combustion could be an option for short-term applications.

The comparison between COE and LCOE shows that the increase of capital costs with air enrichment has a higher impact than the decrease of operating costs. In facts, the former has a significant impact in LCOE behaviour (being paid during the first years of the project, their present values remain high), whereas operating costs (paid during the whole operating life) have a minor impact on LCOE. Such a predominant increase of the influence of capital cost can be observed in figure 9, which shows how each cost item impacts the LCOE. It can also be noticed that the impact of the O&M costs of the CCS system significantly decreases with the increase of oxidant from ASU.

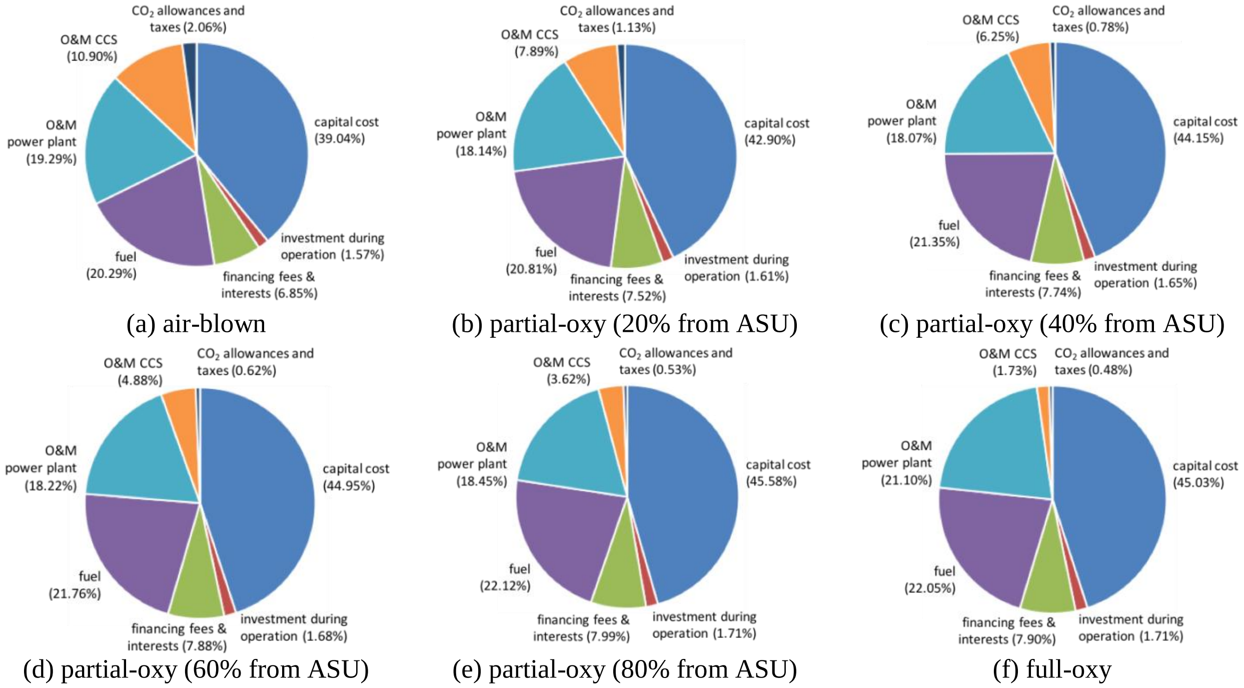


Figure 9. Impact of different costs on LCOE.

As mentioned above, the results here reported have been calculated assuming, for comparative reasons, the same annual availability (7,600 h/yr.) for each plant configuration. This assumption could be quite optimistic for the configurations with significant air enrichment, due to the lack of experience in commercial-scale partial or full oxy-combustion plants. So, a sensitivity analysis has been carried out in order to assess the effect of a potential reduction of plant availability. The analysis considers the following assumptions: (i) for low values of air enrichment (up to 20% of oxidant provided by ASU) plant availability is not influenced and the original value of 7,600 h/yr has been considered; (ii) for the full oxy-combustion configuration, a decrease of plant availability to 7,000 and 6,500 h/yr. has been assumed and a linear variation is considered for the intermediate configurations. As expected, a reduction of the plant annual availability involves an increase of

LCOE, as shown in figure 10. This effect can be observed mainly for the configurations characterized by the strongest reduction of operating hours; it involves that LCOE raises with the air enrichment (without the peak obtained for an air enrichment of 30% in the reference case).

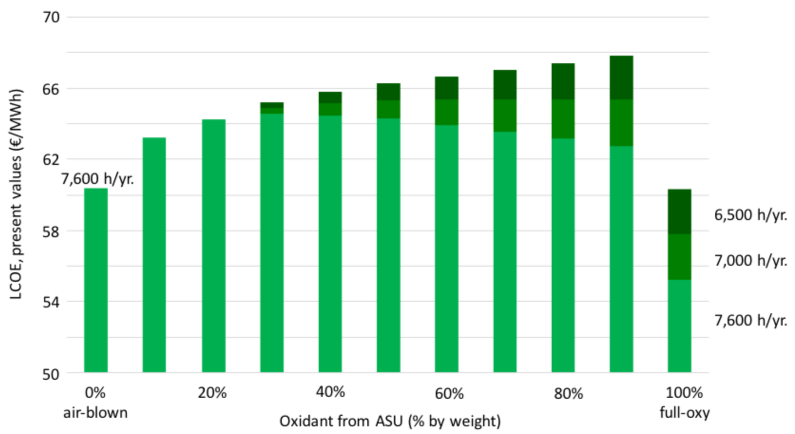


Figure 10. LCOE at different plant availability.

6. Conclusions

In this paper, with the aim to evaluate the feasibility of partial oxy-combustion for commercial applications, a comparative performance analysis – based on simulation models – of an USC power plant equipped with CCS is carried out by varying air enrichment from 0% (conventional air-blown combustion) to 100% (full oxy-combustion).

Such an enrichment involves a significant reduction of oxidizing agent flow: the full-oxy configuration needs almost the same oxygen amount of the air-blown one, which means about 22% of the whole oxidizing flow. As a consequence, a significant decrease of flue gas flow (from 416 kg/s for air-blown to 119 kg/s for full-oxy) can be observed, due to the less dilution with nitrogen.

The reference (without CCS) air-blown plant configuration shows a net power output of 466 MW, leading to a net efficiency of 46.6%. The integration with the CO_2 removal section reduces plant efficiency of 10.7 percentage points to 35.9%. The full-oxy plant configuration shows a net efficiency of 36.1%, very close to the air-blown case (with CCS).

Partial-oxy combustion still requires post-combustion chemical absorption CO_2 capture. In comparison to air-blown process, the higher concentration of CO_2 in flue gas with oxygen enrichment reduces energy penalization associated to solvent regeneration, but this reduction does not compensate for the sensible increase of the ASU energy consumption. Consequently, the plant net efficiency decreases with air enrichment from 35.9% (air-blown) to 31.5% (90% enrichment). Levelized cost of electricity is 60.39 €/MWh for the air-blown configuration and increases up to 64.56 €/MWh for an air enrichment of 30% (i.e. 30% by volume of the oxidant agent comes from the ASU). Then, for high oxygen enrichments, LCOE decreases constantly ant its value drops to 55.21 €/MWh for the full-oxy configuration. The latter appears as the most promising technology for CO_2-free power generation as soon as the experience at commercial-scale will allow the optimization of processes and materials.

It is important to underline that the reported results are a consequence of two key assumptions: (i) the same chemical absorption processes have been considered for both air-blown and partial-oxy configurations and the cryogenic CPU has been considered only for the full-oxy option; (ii) MEA has been considered as solvent, due to the wide availability of reliable data. A future work will be devoted to compare chemical absorption and cryogenic capture as decarbonization options for partial-oxy with high air enrichment and (on the basis of the results of an experimental campaign currently in progress) the possible advantages of using advanced solvents, such as mixtures of MEA and piperazine or MEA and potassium carbonate (K_2CO_3), both characterized by lower values of the specific thermal energy for the regeneration process.

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References

[1] Archer D. Near miss: the importance of the natural atmospheric CO_2 concentration to human historical evolution. Clim Change 2016;138:1–11.

[2] Leung DYC, Caramanna G, Maroto-Valer MM. An overview of current status of carbon dioxide capture and storage technologies. Renew Sustain Energy Rev 2014;39:426–43.

[3] International Energy Agency. Energy and Climate Change. World Energy Outlook Spec Rep. Paris, France, 2015:1–200.

[4] International Energy Agency. World Energy Oulook 2016. Paris, France, 2016.

[5] Dutcher B, Fan M, Russell AG. Amine-based CO_2 capture technology development from the beginning of 2013-A review. ACS Appl Mater Interfaces 2015;7:2137–48.

[6] Casero P, Peña FG, Coca P, Trujillo J. ELCOGAS 14 MWth pre-combustion carbon dioxide capture pilot. Technical & economical achievements. Fuel 2014;116:804–11.

[7] Trapp C, Thomaser T, Van Dijk HAJ, Colonna P. Design optimization of a pre-combustion CO_2 capture plant embedding experimental knowledge. Fuel 2015;157:126–39.

[8] Stelzner B, Weis C, Habisreuther P, Zarzalis N, Trimis D. Super-adiabatic flame temperatures in premixed methane-oxygen flames. Eur Combust Meet 2015:1–6. doi:10.1016/j.fuel.2017.01.025.

[9] Goto K, Yogo K, Higashii T. A review of efficiency penalty in a coal-fired power plant with post-combustion CO_2 capture. Appl Energy 2013;111:710–20.

[10] Martelli E, Kreutz T, Carbo M, Consonni S, Jansen D. Shell coal IGCCS with carbon capture: Conventional gas quench vs. innovative configurations. Appl Energy 2011;88:3978–89.

[11] Cormos CC. Oxy-combustion of coal, lignite and biomass: A techno-economic analysis for a large scale Carbon Capture and Storage (CCS) project in Romania. Fuel 2016;169:50–7.

[12] Escudero AI, Espatolero S, Romeo LM. Oxy-combustion power plant integration in an oil refinery to reduce CO_2 emissions. Int J Greenh Gas Control 2016;45:118–29.

[13] Senneca O, Scala F, Chirone R, Salatino P. Relevance of structure, fragmentation and reactivity of coal to combustion and oxy-combustion. Fuel 2017;201:65-80.

[14] Global CCS Institute. Boundary Dam Carbon Capture and Storage Project. Available at https://www.globalccsinstitute.com/projects/boundary-dam-carbon-capture-and-storage-project (accessed 09 February 2017).

[15] Dutta R, Nord LO, Bolland O. Prospects of using equilibrium-based column models in dynamic process simulation of post-combustion CO_2 capture for coal-fired power plant. Fuel 2017;202:85-97.

[16] An C, Yang S, Huang G, Zhao S, Zhang P, Yao Y. Removal of sulfonated humic acid from aqueous phase by modified coal fly ash waste: Equilibrium and kinetic adsorption studies. Fuel 2016;165:264-271.

[17] Ortiz C, Valverde JM, Chacartegui R, Benítez-Guerrero M, Perejón A, Romeo LM. The Oxy-CaL process: A novel CO_2 capture system by integrating partial oxy-combustion with the Calcium-Looping process. Appl Energy 2017;196:1–17.

[18] Doukelis A, Vorrias I, Grammelis P, Kakaras E, Whitehouse M, Riley G. Partial O_2-fired coal power plant with post-combustion CO_2 capture: A retrofitting option for CO_2 capture ready plants. Fuel 2009;88:2428-2436.

[19] Vega F, Navarrete B, Cano M, Portillo E. Development of partial oxy-combustion technology: New solvents applied to CO_2 capture in fossil-fuels power plants. Energy Procedia 2014;63:484–9.

[20] Lawal A, Wang M, Stephenson P. Investigating the dynamic response of CO_2 chemical absorption process in enhanced-O_2 coal power plant with post-combustion CO_2 capture. Energy Procedia 2011;4:1035–42.

[21] Lawal A, Wang M, Stephenson P, Koumpouras G, Yeung H. Dynamic modelling and analysis of post-combustion CO_2 chemical absorption process for coal-fired power plants. Fuel 2010;89:2791-2801.

[22] Vega F, Sanna A, Maroto-Valer MM, Navarrete B, Abad-Correa D. Study of the MEA degradation in a CO_2 capture process based on partial oxy-combustion approach. Int J Greenh Gas Control 2016;54:160–7.

[23] Vega F, Navarrete B, Alonso-Fariñas B, Rodríguez M. Development of partial oxy-combustion technology: Design, commissioning and experimental program in a pilot plant. Energy Procedia 2014;63:6344–8.

[24] Favre E, Bounaceur R, Roizard D. A hybrid process combining oxygen enriched air combustion and membrane separation for post-combustion carbon dioxide capture. Sep Purif Technol 2009;68:30–6.

[25] Davidson R. Hybrid carbon capture systems. IEA Clean Coal Centre, CCC/204, London, United Kingdom 2012.

[26] Huang Y, Wang M, Stephenson P, Rezvani S, McIlveen-Wright D, Minchener A, et al. Hybrid coal-fired power plants with CO_2 capture: A technical and economic evaluation based on computational simulations. Fuel 2012;101:244–53.

[27] Carrasco-Maldonado F, Spörl R, Fleiger K, Hoenig V, Maier J, Scheffknecht G. Oxy-fuel combustion technology for cement production – State of the art research and technology development. Int J Greenhouse Gas

Control 2016;45:189-199.

[28] Tola V, Cau G, Ferrara F, Pettinau A. CO_2 emissions reduction from coal-fired power generation: A techno-economic comparison. J Energy Resour Technol 2016;138:61602.

[29] Pettinau A, Ferrara F, Tola V, Cau G. Techno-economic comparison between different technologies for CO_2-free power generation from coal. Appl Energy 2017;193:426–39.

[30] Aspen technology Inc. Aspen Plus Version V8.8 user guide. Cambridge, MA, USA 2016.

[31] GE Energy. GateCycle 2016. doi:http://site.ge-energy.com/prod_serv/products/oc/ja/downloads/gatecycle.pdf.

[32] Stępczyńska-Drygas K, Łukowicz H, Dykas S. Calculation of an advanced ultra-supercritical power unit with CO_2 capture installation. Energy Convers Manag 2013;74:201–8.

[33] Zhou L, Xu G, Zhao S, Xu C, Yang Y. Parametric analysis and process optimization of steam cycle in double reheat ultra-supercritical power plants. Appl Therm Eng 2016;99:652–60.

[34] Zhang Q, Fan Y, Li W. Numerical simulation and experimental verification of chemical reactions for SCR DeNOx. Front Chem Eng China 2010;4:523–8.

[35] Hudson J, Thaxton L, Ferguson Jr H, Clay N. Design and Construction of of Baghouses for the Shawnee steam plant. Environ Int 1981;6:69–79.

[36] Gómez A, Fueyo N, Tomás A. Detailed modelling of a flue-gas desulfurisation plant. Comput Chem Eng 2007;31:1419–31..

[37] Geuzebroek FH, Schneiders LHJM, Kraaijveld GJC, Feron PHM. Exergy analysis of alkanolamine-based CO_2 removal unit with AspenPlus. Energy 2004;29:1241–8.

[38] Alie C, Backham L, Croiset E, Douglas PL. Simulation of CO_2 capture using MEA scrubbing: A flowsheet decomposition method. Energy Convers Manag 2005;46:475–87.

[39] Abu-Zahra MRM, Schneiders LHJ, Niederer JPM, Feron PHM, Versteeg GF. CO_2 capture from power plants. Part I. A parametric study of the technical performance based on monoethanolamine. Int J Greenh Gas Control 2007;1:37–46.

[40] Abu-Zahra MRM, Niederer JPM, Feron PHM, Versteeg GF. CO_2 capture from power plants. Part II. A parametric study of the economical performance based on mono-ethanolamine. Int J Greenh Gas Control 2007;1:135–42.

[41] Øi LE. Comparison of aspen HYSYS and aspen plus simulation of CO_2 absorption into MEA from atmospheric gas. Energy Procedia 2012;23:360–9.

[42] Yokoyama T. Analysis of reboiler heat duty in MEA process for CO_2 capture using equilibrium-staged model. Sep Purif Technol 2012;94:97–103.

[43] Donda F, Volpi V, Persoglia S, Parushev D. CO_2 storage potential of deep saline aquifers: The case of Italy. Int J Greenh Gas Control 2011;5:327–35.

[44] Toftegaard MB, Brix J, Jensen PA, Glarborg P, Jensen AD. Oxy-fuel combustion of solid fuels. Prog Energy Combust Sci 2010;36:581–625.

[45] Wetenhall B, Race JM, Downie MJ. The Effect of CO_2 Purity on the Development of Pipeline Networks for Carbon Capture and Storage Schemes. Int J Greenh Gas Control 2014;30:197–211.

[46] Pettinau A, Ferrara F, Amorino C, Combustion vs. gasification for a demonstration CCS project in Italy: a techno-economic analysis. Energy 2013;50:160-9.

[47] Tola V, Pettinau A. Power generation plants with carbon capture and storage: A techno-economic comparison between coal combustion and gasification technologies. Applied Energy 2014;113;1461-74.

[48] U.S. Department of Energy, National Energy Technology Laboratory. Cost and performance baseline for fossil energy plants. Volume 1a: Bituminous coal (PC) and natural gas to electricity - Revision 3. Report DOE/NETL-2015/1723, July 2015.

[49] U.S. Department of Energy, National Energy Technology Laboratory. Cost and performance baseline for fossil energy plants. Volume 1b: Bituminous coal (IGCC) to electricity - Revision 2b - Tear dollar update. Report DOE/NETL-2015/1727, July 2015.

[50] Nicol K. Application and development prospects of double-reheat coal-fired power units. IEA Clean Coal Centre, CCC/255, London, United Kingdom, 2015.

[51] Xiong J, Zhao H, Zheng C. Thermoeconomic cost analysis of a 600 MWe oxy-combustion pulverized-coal-fired power plant. Int J Greenhouse Gas Control 2012;9:469-483.

[52] Thomson Reuters. EU carbon price to average €23/t between 2021 and 2030: Thomson Reuters assess the future. 28th August 2014. Available at http://blog.financial.thomsonreuters.com/eu-carbon-price-average-e23t-2021-2030-thomson-reuters-assess-future/ (accessed 30 June 2016).

[53] Metz B, Davidson O, de Coninck H, Loos M, Meyer L. IPCC special report on carbon dioxide capture and storage. Cambridge University Press, Cambridge, 2005.

[54] Hendriks C, Graus W, van Bergen F. Global carbon dioxide storage potential and costs. Ecofys report EEP-02001, Utrecht (The Netherlands), 2004.