

# CO<sub>2</sub>-free coal-fired power generation by partial oxy-fuel and post-combustion CO<sub>2</sub> 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<sub>2</sub> capture is gaining interest, since such a hybrid configuration could allow to reduce the size and enhance the performance of post-combustion CO<sub>2</sub> 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<sub>2</sub> capture section.

This work presents the results of a comparative techno-economic analysis of a 1000 MW<sub>th</sub> partial oxy-combustion plant based on an ultra-supercritical pulverized coal combustion power plant integrated with a post-combustion CO<sub>2</sub> 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.

28

29

30 **Keywords:** Carbon capture and storage; Partial oxy-combustion; CO<sub>2</sub> capture, Techno-economic  
31 analysis

32

### 33 **Acronyms**

34 ABS, absorption column; ASU, air separation unit; BEC, bare erected cost; BF, baghouse filters; CCS, carbon capture and storage; CCTS, carbon  
35 capture, transport and storage; CCU, carbon capture and utilization; COE, cost of electricity; CPU, CO<sub>2</sub> capture and purification unit; DES,  
36 regeneration (desorption) column; EOR, enhanced oil recovery; FGC, flue gas cleanup; FGD, flue gas desulphurization; HPT, high pressure turbine;  
37 HTX, heat exchanger; IPT, intermediate pressure turbine; LCOE, levelized cost of electricity; LHV, lower heating value; LPT, low pressure turbine;  
38 MEA, monoethanolamine; NETL, U.S. National Energy Technology Laboratory; RH, re-heater; SCR, selective catalytic reduction; SH, super-heater;  
39 USC, ultra-supercritical pulverized coal combustion; VAT, value added tax; VHPT, very high pressure turbine.

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## 43 **1. Introduction**

44 The increase of the atmospheric CO<sub>2</sub> concentration has led to several environmental issues, notably  
45 an increase in global temperatures commonly referred to as global warming [1-3]. In this context,  
46 carbon capture and storage (CCS) and carbon capture and utilization (CCU) technologies must play  
47 a key role for its mitigation [4,5].

48 In general, CO<sub>2</sub> capture technologies can be classified according to three main approaches: (1) post-  
49 combustion, (2) pre-combustion, (3) oxy-fuel combustion [4,5]. In the post-combustion approach,  
50 fossil fuels are burned (as in conventional power plants) and then the CO<sub>2</sub> is captured from the flue  
51 gas. In the pre-combustion approach, the fossil fuel is gasified and the produced syngas is treated in  
52 a water-gas shift reactor to convert CO and water vapour into H<sub>2</sub> and CO<sub>2</sub> [6,7]. The latter is  
53 captured, while the hydrogen-rich syngas feeds a combined cycle plant for power generation. The  
54 oxy-fuel approach utilizes pure or nearly pure oxygen for combustion, such that primarily CO<sub>2</sub> and  
55 H<sub>2</sub>O are produced by the process [5,8]. All these approaches are characterized by very high energy  
56 penalties: the plant net efficiency could be reduced of about 8-12 percentage points in case of post-

57 combustion processes (mainly due to solvent regeneration) [9,10], and of 7-10 percentage points in  
58 case of pre-combustion approach [10]. Based on the state-of-the-art of a supercritical pulverized  
59 coal power plant, the efficiency losses related to oxy-fuel combustion are in the range of 9-13  
60 percentage points [11], but it is likely that they can be reduced to 7-11 percentage points by means  
61 of processes optimization and heat integration [12]. So, oxy-fuel approach promises to become  
62 more and more interesting for future applications [13].

63 Overall, the very high cost of CCS technologies and the lack of experience in industrial-scale units  
64 are the key issues that are limiting the commercial application of the technologies. Therefore, today,  
65 the only full size CCS application in the world is represented by the Boundary Dam Carbon Capture  
66 Project in Estevan town (Saskatchewan, Canada), where the captured CO<sub>2</sub> is transported by pipeline  
67 (for 66 km) and injected for enhanced oil recovery (EOR) at the SaskPower's Weyburn oil field  
68 [14-16].

69 The main drawback for the large-scale deployment of oxy-combustion is the high energy  
70 consumption for pure O<sub>2</sub> production in the air separation unit (ASU), which causes a significant  
71 energy penalty [17].

72 One of the proposed solutions for short-term commercial applications is a compromise between  
73 post- and oxy-combustion approaches, a hybrid configuration commonly called partial oxy-fuel or  
74 partial oxy-combustion [18]. Primary fuel is burned in an oxygen-enriched environment in order to  
75 reduce the dilution of flue gas by nitrogen, thus enhancing the CO<sub>2</sub> concentration. The ASU for  
76 oxygen separation is smaller (which means a lower incidence in terms of capital cost and energy  
77 penalty) than the same equipment required by the oxy-combustion and the flue gas recirculation  
78 requires minor modifications on conventional boilers; in parallel, thanks to the less dilution by  
79 nitrogen, the volume of flue gas to be treated is significantly lower and CO<sub>2</sub> partial pressure is  
80 higher than in conventional post-combustion processes [19].

81 One of the first studies on the application of partial oxy-combustion for the retrofit of power plants  
82 has been published in 2009 by Doukelis et al. [18] and presents the so-called ECO-Scrub scheme as

83 a good compromise between post-combustion capture and oxy-fuel. One of the key issues regarding  
84 the optimization of a partial oxy-combustion process is related to the definition of the optimal O<sub>2</sub>  
85 concentration in the enriched air. The specific effect of O<sub>2</sub> enrichment in amine-based chemical  
86 absorption has been studied by Lawal et al. [20,21], whereas Vega et al. [19,22,23] have presented  
87 an experimental study on monoethanolamine (MEA) degradation in partial-oxy-combustion CO<sub>2</sub>  
88 capture. Other post-combustion CO<sub>2</sub> capture technologies, such as membranes [24,25], calcium  
89 looping [17] and cryogenic separation [26], have been considered for the potential application in  
90 partial oxy-combustion scheme. Unfortunately, a lack of publications on the effect of oxygen  
91 concentration on plant efficiency and economic performance in partial oxy-combustion CO<sub>2</sub>-free  
92 coal-fired power generation plants can be observed. Only Huang et al. (2012) [26] present an  
93 interesting techno-economic parametric analysis on hybrid coal-fired power plants (intended as  
94 oxy-fuel unit with a variable air dilution – up to 50% – and based on a cryogenic post-combustion  
95 CO<sub>2</sub> capture system). Finally, the same approach has been used in several applications in the  
96 cement industry, but with different techno-economic performance [27].

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

105 Performance evaluation has been carried out through simulation models based on the Aspen-Plus  
106 and Gate-Cycle commercial tools [30,31]. In particular, Gate-Cycle models are used to simulate the  
107 steam power plant in both air-blown and oxy-fuel arrangements, whereas Aspen-Plus models are

108 used to simulate the conditioning and purification processes of exhaust gas and the air separation  
109 unit (ASU) process.

110

## 111 **2. Plant configurations**

112 As the main aim of this study is to make a techno-economic comparison between post-combustion,  
113 full and partial oxy-combustion approaches, the study considers, for each plant configuration, the  
114 same coal chemical power input of 1000 MW and the same USC power generation unit, equipped  
115 with a conventional flue gas cleanup (FGC) section and a low temperature CO<sub>2</sub> removal section,  
116 based on a chemical absorption process with an aqueous solution of MEA. To match CO<sub>2</sub> transport  
117 and storage requirements, the CO<sub>2</sub> removal section is also integrated with a conditioning and  
118 compression section to provide a high pressure (11 MPa) and high purity (CO<sub>2</sub> fraction of 99.7% by  
119 volume) CO<sub>2</sub> flow. Moreover, each plant configuration is considered to be fed with a commercial  
120 coal, whose main characteristics (lower heating value – LHV – proximate and ultimate analysis) are  
121 reported in table 1.

122

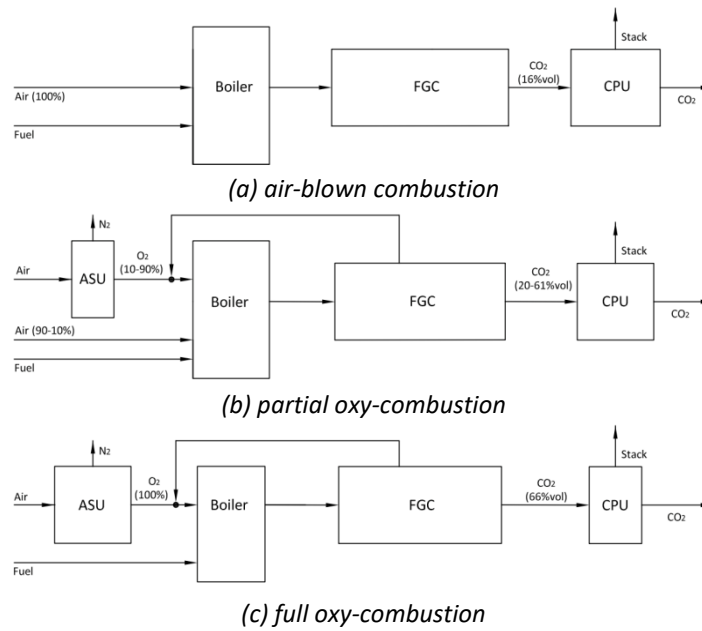
<i>Proximate Analysis (% by weight)</i>	
Fixed carbon	52.70
Volatile matter	25.90
Ash	14.40
Moisture	7.00
<i>Ultimate Analysis (% by weight)</i>	
Total carbon	65.66
Hydrogen	3.64
Sulphur	0.85
Nitrogen	1.61
Oxygen	6.84
Ash	14.40
Moisture	7.00
<i>Heating value (MJ/kg)</i>	
Lower heating value	25.03

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

124

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

126



127 Figure 1. Conceptual scheme of the three configurations.

128

129 2.1. Air-blown configuration and USC steam cycle

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

132

133 2.1.1. Steam cycle

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

141 Due to the presence of the double reheat, the selected configuration includes four steam turbines: a  
 142 very high-pressure turbine (VHPT), a high-pressure turbine (HPT), an intermediate pressure turbine  
 143 (IPT) and a low-pressure turbine (LPT). Figure 2 shows a simplified scheme of the air-blown USC

144 power plant, whereas the main operating parameters assumed for the simulation models are  
 145 reported in tables 2 and 3.

146

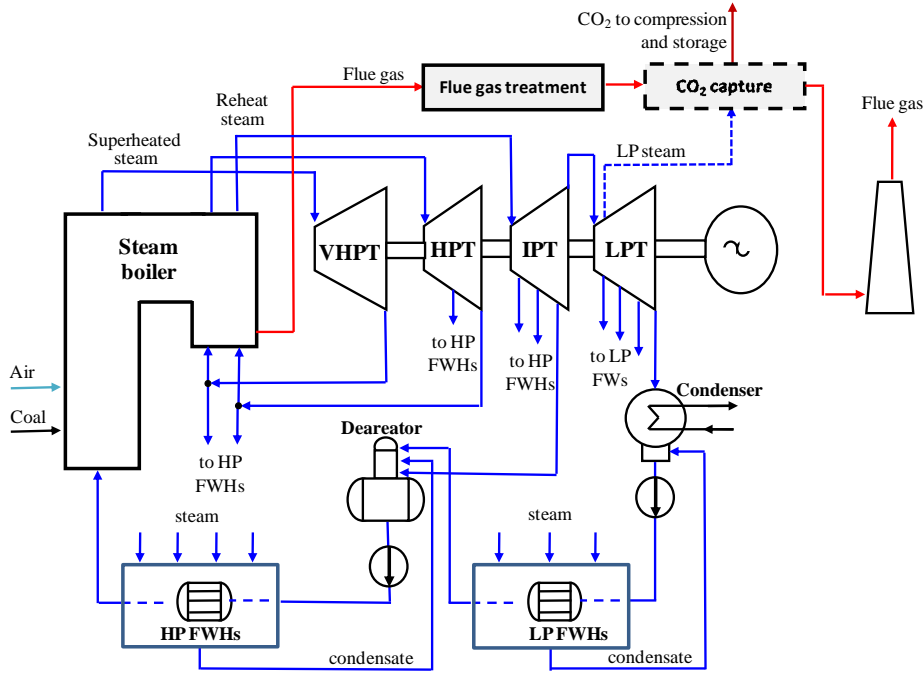


Figure 2. Simplified scheme of the air-blown USC plant.

147

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 $\Delta T$ (°C)	-1.5/1.5

Table 2. Main USC operating parameters.

148

149

	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.

150

151

152 VHPT and HPT expansion ratios (about 0.48 and 0.42, respectively) have been chosen in order to  
153 maximize the efficiency of the double reheat steam cycle [33]. A first steam extraction is performed  
154 at the VHPT output, whereas, in the order, 2, 3, 4 extractions are performed in the HPT, IPT and  
155 LPT respectively. The very high pressure of the first steam extraction (slightly lower than 15 MPa)  
156 allows to increase water temperature upstream of the economizer above 335 °C.  
157 Steam extraction pressures are established, regardless of turbine functional and constructive  
158 constraints, in order to assure a similar temperature rise inside the feedwater heat exchangers.

159

#### 160 2.1.2. Flue gas treatment systems

161 The flue gas exiting from the boiler is sent to a conventional flue gas cleanup (FGC) section. A  
162 high-dust FGC configuration has been assumed, including a selective catalytic reduction (SCR)  
163 denitrification system for NO<sub>x</sub> removal, baghouse filters (BF) for particulate removal and a low  
164 temperature flue gas desulphurization (FGD) system for SO<sub>x</sub> removal.  
165 SCR section causes a flue gas pressure drop in the range of 5-10 kPa, leading to an electrical power  
166 requirement for driving the fans of about 1% of the overall plant generation [34].  
167 Baghouse filters, are installed downstream of the air preheater at 120-180 °C and cause a flue gas  
168 pressure drop of about 1-2 kPa, assuring a removal efficiency higher than 99% [35].  
169 FGD process operates at low temperature with a flue gas pressure drop in the range of 5-10 kPa,  
170 requiring an electrical power of about 1% of the overall plant generation [36]. Globally, such a  
171 section accounts for an overall electrical power consumption of about 9 MW, mainly due to fan  
172 requirements for pressure drop of flue gas. Electrical power accounts for about 2% of the gross  
173 plant power, penalizing the plant efficiency of about one percentage point.

174

#### 175 2.1.3. CO<sub>2</sub> capture and compression



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

179 Such a process allows a CO<sub>2</sub> removal efficiency of 90% [39,40], separating high-purity (92-93% by  
 180 volume) CO<sub>2</sub>, which is sent to the conditioning and compression section.

181 The performance analysis of the CO<sub>2</sub> removal process has been carried out under equilibrium  
 182 conditions, leading to an acceptable approximation [41,42].

183 The model assumes a MEA concentration of 30% (by weight) and a CO<sub>2</sub>/MEA molar ratio of 0.28.

184 The main assumptions and simulation results of the CO<sub>2</sub> removal process are reported in table 4.

185

CO <sub>2</sub> removal efficiency (%)	90.0
Flue gas mass flow at the absorber inlet (kg/s)	410.2
CO <sub>2</sub> 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 <sub>2</sub> 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 <sub>2</sub> /MEA molar ratio at the absorber inlet	0.28
CO <sub>2</sub> -lean solvent temperature at the absorber inlet (°C)	35.0
CO <sub>2</sub> -rich solvent temperature at the absorber outlet (°C)	50.5
CO <sub>2</sub> -rich solvent temperature at the desorber inlet (°C)	90.0
CO <sub>2</sub> -lean solvent temperature at the desorber outlet (°C)	102.7
CO <sub>2</sub> mass flow (kg/s)	85.7
CO <sub>2</sub> molar fraction in stream to CO <sub>2</sub> compressors	0.924
Reboiler specific thermal energy (MJ/kg <sub>CO2</sub> )	3.72

186 Table 4. Main operating parameters and performance of the CO<sub>2</sub> removal section.

187

188 In order to obtain a removal efficiency of 90%, a solvent/gas mass ratio of about 4.5 and a reboiler  
 189 specific thermal energy of 3.75 GJ per ton of removed CO<sub>2</sub> have been calculated. The flue gas from  
 190 the CO<sub>2</sub> capture section is mainly composed by N<sub>2</sub> (about 78%, by volume), while the CO<sub>2</sub>  
 191 concentration decreases from about 14% to about 1.5%. The CO<sub>2</sub>-rich gas from the absorption  
 192 section is compressed to the transport pressure (11 MPa). It has been assumed that the compression  
 193 process takes place up to 8 MPa by three intercooled compressors in series and then through a  
 194 pump. The substantial water condensation leads to an almost pure CO<sub>2</sub> flow (with a molar fraction  
 195 over 99.5%), as required for transport and storage.

196 The CO<sub>2</sub> removal dramatically affects the plant performance. In particular, the thermal power  
197 required by the reboiler to desorb CO<sub>2</sub> is remarkable (about 320 MW) and it is supplied by a low-  
198 pressure (0.39 MPa) steam extraction carried out in the LPT, which notably affects the plant power  
199 output. Another significant energy consumption is the electrical power required by the CO<sub>2</sub>  
200 compression and pumping system (about 30 MW), whereas the power required by the fan of the  
201 decarbonization section is limited to about 3 MW.

202

#### 203 2.1.4. CO<sub>2</sub> transport and storage

204 The high-pressure and almost pure CO<sub>2</sub> stream exiting from the conditioning and compression  
205 section must be transported to the site designed to carbon dioxide storage. Transport of CO<sub>2</sub> has  
206 become a key factor in CCS, fixing CO<sub>2</sub> characteristics in terms of purity and pressure suitable for  
207 transportation. A 25 km long pipeline has been assumed as transport mode to the geological storage  
208 site for the captured carbon dioxide. The injection in saline aquifers has been chosen as the storage  
209 option in this study representing one of the highest storage capacity solution [43].

210

#### 211 2.2. Full oxy-combustion plant configuration

212 The oxy-combustion plant configuration is based on the same steam cycle of the air-blown plant.  
213 The main functional and constructive differences regard the boiler, the oxygen supplied by a  
214 cryogenic ASU and the flue gas management and clean-up. As a matter of fact, oxy-combustion  
215 leads to higher temperatures in comparison to air-blown boiler. Therefore, flue gas recirculation (in  
216 this case about 70%, at a temperature of 310 °C) is carried out to control the flame temperature [44]  
217 and to obtain a boiler heat transfer profile similar to the one in air-blown steam generators [26].  
218 Flue gas contains mainly CO<sub>2</sub> and water vapour and a small amount of un-reacted oxygen and inert  
219 gases. Consequently, just a CO<sub>2</sub> purification unit is required to attain a high purity CO<sub>2</sub> stream,  
220 avoiding the post-combustion CO<sub>2</sub> capture and its strong energy penalty. However, a remarkable

221 energy penalty is related to the ASU for oxygen production and to the CO<sub>2</sub> compression for  
 222 transport and storage.

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

224

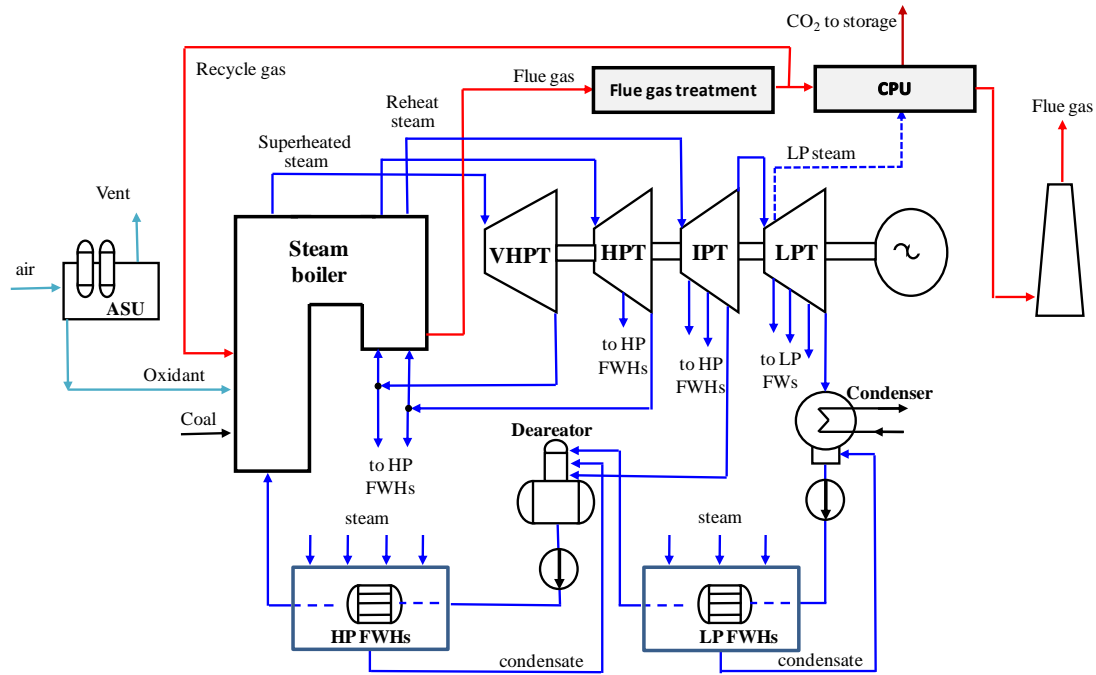


Figure 3. Simplified scheme of the full-oxy configuration.

225

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

227

Oxydant mass flow (kg/s)	85.26
O <sub>2</sub> /N <sub>2</sub> /Ar molar fractions in oxydant	0.95/0.02/0.03
O <sub>2</sub> specific separation energy (kWh/t <sub>O<sub>2</sub></sub> )	200.0
Flue gas recycle rate	0.684
Recycle gas mas flow (kg/s)	257.2
Recycle gas temperature (°C)	307.9

228

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

229

230 The power unit is equipped with a flue gas cleanup system similar to that used in the air-blown

231 USC configuration, including SCR, BF and FGD systems. The high concentration of CO<sub>2</sub> in flue

232 gas influences both DeSO<sub>x</sub> and DeNO<sub>x</sub> systems, but most of the studies assume that they can

233 operate with better performance than in conventional steam plants [44]. Clean gas is mainly

234 composed by CO<sub>2</sub> (about 66% by volume) and water vapour (about 26%), with small amounts of N<sub>2</sub>

235 (3%), O<sub>2</sub> (2.5) and Ar (1.5%). The CO<sub>2</sub> could be easily separated by water condensation, but a CO<sub>2</sub>  
236 capture and purification unit (CPU) is still required to reduce the amount of oxygen and other  
237 incondensable gases and match the CO<sub>2</sub> purity requirements for transportation and storage [45]. In  
238 such a unit, the CO<sub>2</sub>-rich gas is firstly cooled and compressed up to about 2.5 MPa with the  
239 condensation of a large amount of water. The CO<sub>2</sub>-rich gas is cooled to -40 °C with the  
240 condensation of the largest portion of CO<sub>2</sub> and the separation of a considerable amount of  
241 incondensable gases. Then, the high-purity CO<sub>2</sub> stream is heated and sent to the second section of  
242 the compression train where the almost pure CO<sub>2</sub> gas is pressurized to transport and storage  
243 conditions (about 11 MPa). Conversely, the separated incondensable gases (N<sub>2</sub>, O<sub>2</sub>, Ar and residual  
244 CO<sub>2</sub>) expand in a turbine to recover energy.

245 A larger CO<sub>2</sub> removal efficiency than that obtained with the post-combustion section has been  
246 calculated (about 94%) with a CO<sub>2</sub> purity of 96.4%. CO<sub>2</sub>-rich gas is still composed by a smaller  
247 amount of N<sub>2</sub> (1.3%), O<sub>2</sub> (1.4%) and Ar (0.8%). The whole power requirement of the intercooled  
248 compression train is considerably higher than the one associated with the post-combustion section,  
249 due to the freezing unit (even if the compressors needs less energy due to the lower temperature of  
250 the treated stream). However, most of the CPU energy absorption is required for compression of  
251 CO<sub>2</sub>, while a smaller amount is required for the separation of impurities.

252

### 253 2.3. Partial oxy-combustion configuration

254 The partial oxy-combustion configuration is a compromise between air-blown and full-oxy ones.  
255 Conceptually, an enrichment in oxygen of the combustion air involves a reduction of flue gas  
256 dilution by nitrogen. So, in the partial-oxy configuration, the boiler is fed with a mixture of  
257 atmospheric air (with an O<sub>2</sub> molar fraction of 0.206) and oxygen-rich gas (with a purity of 95%)  
258 produced by the ASU. Flue gas is characterized by a lower mass flow and by a higher CO<sub>2</sub>  
259 concentration in comparison with the air-blown configuration and it is treated by a similar (except  
260 for the size) high-dust FGC system. The configuration of the post-combustion CO<sub>2</sub> capture unit is

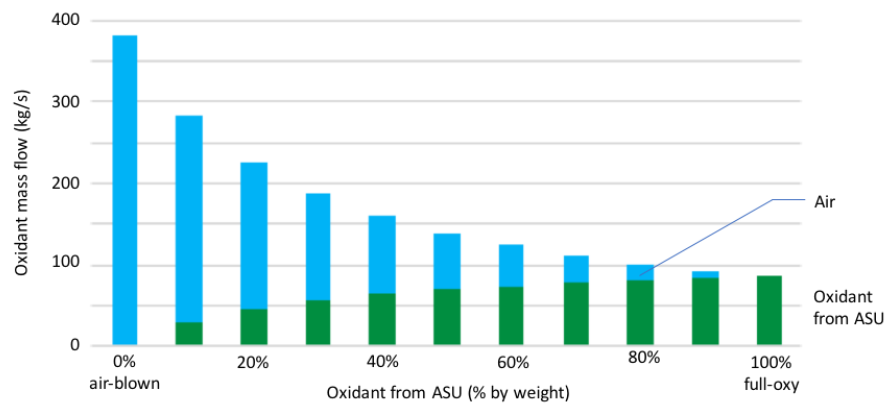
261 the same considered in the air-blown case, but the higher CO<sub>2</sub> concentration involves better  
262 performance and a lower equipment size. Finally, the same compression system of the air-blown  
263 configuration has been considered for the partial-oxy approach.

264

### 265 3. Parametric analysis

266 A performance analysis has been carried out to assess the influence of air enrichment on plant  
267 performance and CO<sub>2</sub> removal, conditioning and compression processes. The increase of O<sub>2</sub>  
268 concentration in the oxidant involves a lower oxidant mass flow required for combustion, as shown  
269 in figure 4.

270



271

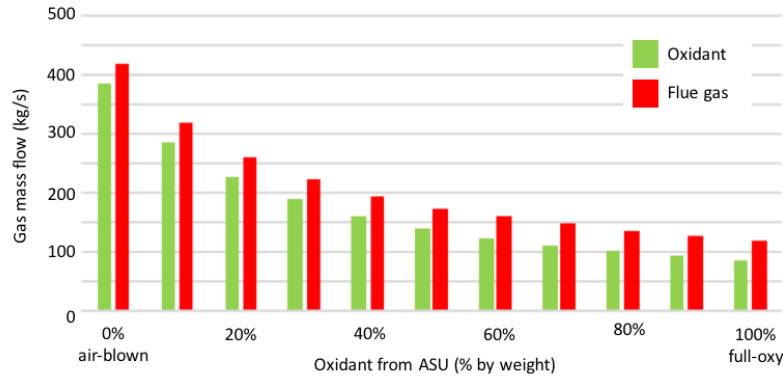
272 Figure 4. Whole oxidant mass flow as a function of oxidant from the ASU.

273

274 The air-blown plant configuration requires an air mass flow slightly higher than 380 kg/s, while the  
275 oxidant mass flow is reduced to about 85 kg/s with the full-oxy configuration. A 10% partial-oxy  
276 leads to an oxidant mass flow reduction of about 100 kg/s compared to the air-blown case. The  
277 mass flow of oxidant from ASU increases significantly for low air enrichment ratios (28.3 kg/s at  
278 10% enrichment and 45 kg/s with 20% enrichment). The increase of ASU mass flow is moderate  
279 for major values, up to a maximum ASU production of about 85 kg/s for the full-oxy configuration.  
280 The reduction in the oxidant mass flow leads to a sensible decrease of the flue gas mass flow. A  
281 lower mass flow to be treated in the subsequent conditioning systems leads to a substantial

282 reduction of the power requirements of the FGC section. Figure 5 shows how air enrichment, with  
 283 the corresponding reduction of oxidant mass flow, involves a significant decrease of flue gas mass  
 284 flow.

285



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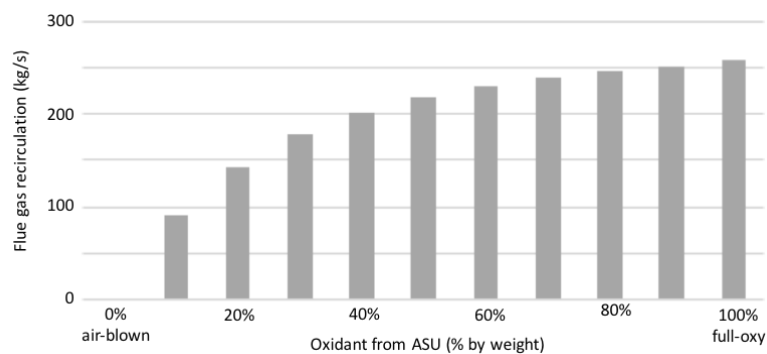
287 Figure 5. Reduction of the flue gas mass flow with air enrichment.

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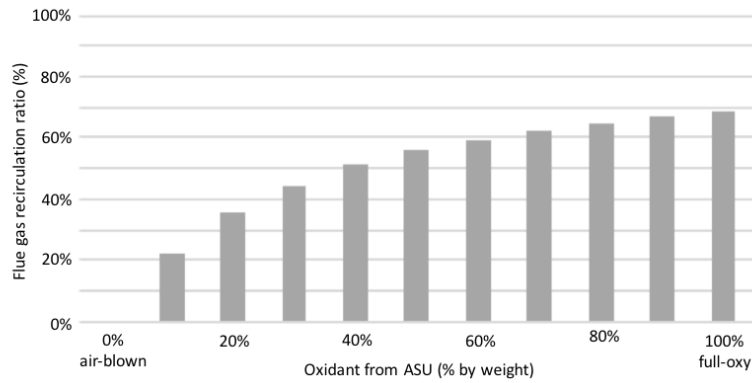
289 The air-blown plant configuration produces a flue gas mass flow slightly higher than 415 kg/s,  
 290 while it is reduced to about 120 kg/s with the full-oxy configuration. The decrease of the flue gas  
 291 mass flow is very pronounced at the lower values of the air enrichment: a 10% enrichment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

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

296



(a) mass flow.



(b) recirculation ratio.

Figure 6. Flue gas recirculation and recirculation ratio.

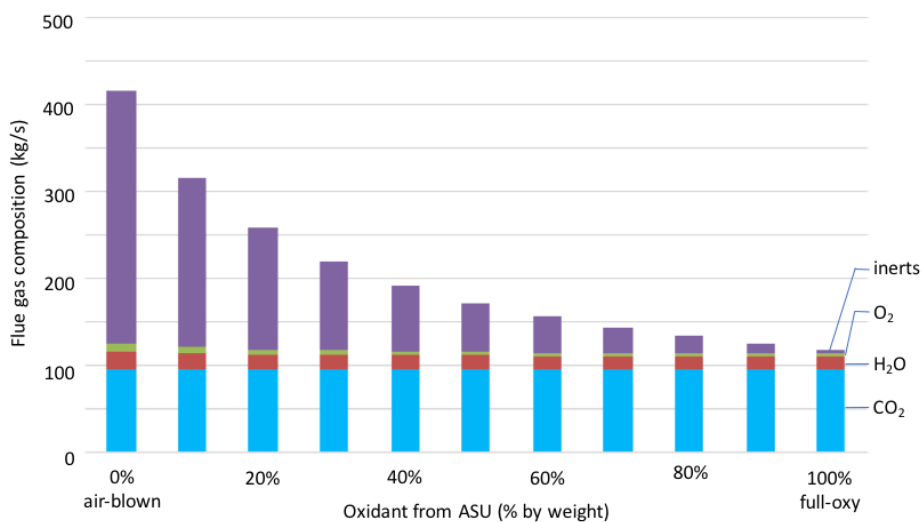
297

298

299 A gas mass flow of about 260 kg/s is recirculated to the boiler in the full-oxy configuration. A 10%  
 300 enrichment requires about 90 kg/s of gas recirculation, while a recirculated mass flow greater than  
 301 200 kg/s is required starting from a 40% enrichment. As a matter of fact, a higher fraction of  
 302 recirculated flue gas corresponds to a greater mass flow of flue gas recirculated.

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

305



306

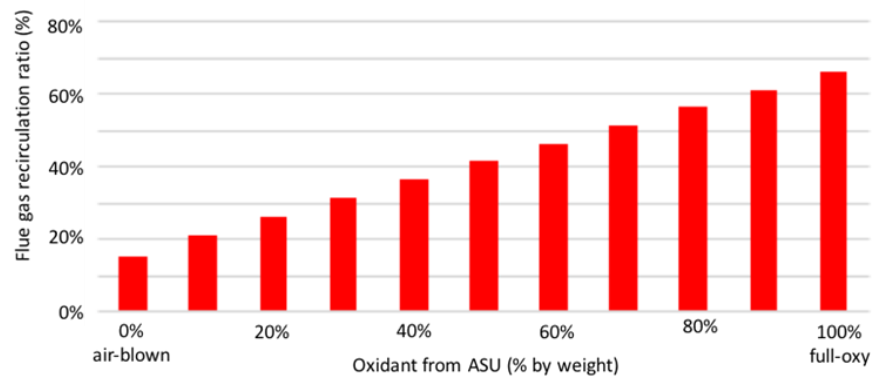
307

308

Figure 7. Composition of the flue gas.

309 The CO<sub>2</sub> content remains constant, only depending on coal feeding, but, due to the reduction of the  
310 gas flow, its concentration increases from 15.5% to 65.9% (by volume) as shown on figure 8, where  
311 the CO<sub>2</sub> molar fraction in the flue gas is reported as a function of the percentage of oxidant from  
312 ASU. The mass flow of inert gas (nitrogen and argon) is largely reduced increasing the air  
313 enrichment. Also, a slight reduction of water vapour and residual oxygen can be observed  
314 increasing air enrichment.

315



316

317 Figure 8. CO<sub>2</sub> concentration in flue gas.

318

319 A more concentrated flue gas improves solvent regeneration, slightly reducing thermal energy  
320 required in the reboiler, from a maximum value of about 3.75 GJ per ton of CO<sub>2</sub> removed (air-  
321 blown combustion) to a minimum value of about 3.50 GJ/t<sub>CO2</sub> with a 90% enrichment. Despite a  
322 modest reduction of the specific thermal energy required by the reboiler, partial oxy-combustion  
323 enhances CO<sub>2</sub> removal process. In fact, the treatment of a flue gas with a more concentrated CO<sub>2</sub>  
324 greatly reduces the MEA degradation process [22].

325

#### 326 4. Performance comparison

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



329 The reference (without CCS) air-blown plant shows a steam cycle output of about 500 MW.  
330 Auxiliaries (air fans, cooling water pumps, etc.) power absorptions, mechanical and generator  
331 losses reduce power output to about 475 MW. Considering the FGC section consumption, finally  
332 results a net power output slightly higher than 465 MW, leading to a net efficiency of 46.60%. The  
333 integration with the CO<sub>2</sub> removal section reduces the gross output of about 75 MW, mainly due to  
334 the large steam extraction from the steam turbine for solvent regeneration. This remarkable penalty  
335 combined with the power requirements of the CO<sub>2</sub> capture and compression section causes a  
336 noteworthy power output reduction slightly lower than 110 MW. Globally, CCS system reduces  
337 plant efficiency of 10.7 percentage points to 35.90%.

338 Full-oxy plant configuration shows a gross power output sensibly higher than the air-blown  
339 configuration with CCS (478.2 MW vs. 400.4 MW), due to the absence of steam extraction for  
340 solvent regeneration. However, the noteworthy power absorption of the ASU (more than 60 MW)  
341 and the high power requirement of the CPU unit lead to a net power output of about 360 MW and a  
342 net efficiency of 36.1%, very close to the air-blown case. On the other hand, such a configuration  
343 leads to CO<sub>2</sub> specific emissions (about 55 g/kWh) lower than those of air-blown CO<sub>2</sub>-free one  
344 (about 95 g/kWh), thanks to a higher CO<sub>2</sub> removal efficiency (about 94.0%).

345 Partial oxy-combustion configurations present a gross power output comparable to that of the air-  
346 blown CO<sub>2</sub>-free one (in the range 400-405 MW), but the net power output is dramatically reduced  
347 by the presence of the ASU. A net power output of about 340 MW has been calculated for a 10%  
348 enrichment, while the net power output is reduced to about 315 MW for a 90% enrichment. The  
349 lower power output associated to partial-oxy configurations leads to a slight increase (in the range  
350 of 100-110 g/kWh) of CO<sub>2</sub> specific emissions in comparison to air-blown configuration.

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

<i>Configuration</i>	<i>ref. (no CCS)</i>	<i>air-blown</i>	<i>partial-oxy</i>	<i>partial-oxy</i>	<i>partial-oxy</i>	<i>partial-oxy</i>
<i>Oxidant from ASU (% by weight)</i>	<b>0%</b>	<b>0%</b>	<b>10%</b>	<b>20%</b>	<b>30%</b>	<b>40%</b>
<i>O<sub>2</sub> concentr. in oxidant (% vol.)</i>	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 <sub>2</sub> 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 <sub>2</sub> emissions (Mt/year)	2.60	0.260	0.260	0.260	0.260	0.260
CO <sub>2</sub> 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.

356

357

<i>Configuration</i>	<i>partial-oxy</i>	<i>partial-oxy</i>	<i>partial-oxy</i>	<i>partial-oxy</i>	<i>partial-oxy</i>	<i>full-oxy</i>
<i>Oxidant from ASU (% by weight)</i>	<b>50%</b>	<b>60%</b>	<b>70%</b>	<b>80%</b>	<b>90%</b>	<b>100%</b>
<i>O<sub>2</sub> concentr. in oxidant (% vol.)</i>	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 <sub>2</sub> 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 <sub>2</sub> emissions (Mt/year)	0.260	0.260	0.260	0.260	0.260	0.156
CO <sub>2</sub> 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.

358

359

360

## 361 5. Cost evaluation

362 The economic and financial assessment of the whole CCS project at different oxygen

363 concentrations in the oxidant agent has been carried out on the basis of the levelized cost of

364 electricity (LCOE) and other economic indicators. The study has been carried out by using a

365 detailed economic model and considering the year 2017 as the starting year of the project. This

366 assumption allows to compare the economic results on partial oxy-combustion configurations with

367 the results on post- and oxy-combustion, previously published by the authors [29].

368

## 369 5.1. Project's milestones and financial assumptions

370 The economic analysis is based on several key assumptions. First of all, the investment is  
371 distributed in the four years of the construction phase (24%, 39%, 32% and 5%), starting from the  
372 year 2017 [46], and the whole operating life of the project is assumed 25 years (2021 to 2045).

373 The study is based on the realistic assumption that 80% of the investment for plant construction is  
374 supported by the banks through the opening of a senior debt (with a financing fee of 2.5% and a  
375 constant annual interest rate of 6.14% in 10 years), whereas the remaining 20% is directly provided  
376 by the owner company. A value added tax (VAT) of 22% is assumed for both capital and operating  
377 costs [47]. An amortization rate of 10% has been assumed for both the power generation and the  
378 CCS systems, whereas a rate of 14% is considered for the material handling system [47]. The model  
379 also considers a yearly extra investment during the operation of the plant [46].

380 Finally, the calculation of the present values is based on an assumed annual discount rate of 8%  
381 [26].

382

## 383 5.2. Capital and operating costs estimation

384 Capital costs of each component are assessed on the basis of industrial data recently published by  
385 the U.S. National Energy Technology Laboratory (NETL) [48,49], following the same approach  
386 widely described in Pettinau et al. 2017 [29]. In addition, the following assumptions have been  
387 taken for the boilers components: (i) the cost of the air-blown boiler is the same as reported in [29],  
388 with an extra cost of 50 €/kW [50] to consider the second reheat; (ii) the cost of the full-oxy boiler  
389 has been calculated from the air-blown one, with an extra cost of 7% [51] to consider the different  
390 operating conditions; (iii) the costs of the boilers for partial-oxy configurations are calculated  
391 through a linear variation between air-blown and full-oxy configurations, on the basis of oxygen  
392 enrichment. Moreover, the full-oxy configuration considers a cryogenic CO<sub>2</sub> separation system,  
393 whose cost (including CO<sub>2</sub> compression) has been calculated as 10.3% of the bare erected cost [26].

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

395

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 <sub>2</sub> 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 <sub>2</sub> removal system	0.00	241,374.03	204,942.97	181,469.89	164,846.07	152,359.66
CO <sub>2</sub> compression and drying	0.00	49,996.52	49,996.52	49,996.52	49,996.52	49,996.52
CO <sub>2</sub> transport	0.00	26,691.55	26,691.55	26,691.55	26,691.55	26,691.55
CO <sub>2</sub> 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

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

397

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 <sub>2</sub> 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 <sub>2</sub> removal system	142,560.39	134,673.96	128,139.25	122,645.58	117,938.10	106,195.71
CO <sub>2</sub> compression and drying	49,996.52	49,996.52	49,996.52	49,996.52	49,996.52	49,996.52
CO <sub>2</sub> transport	26,691.55	26,691.55	26,691.55	26,691.55	26,691.55	26,691.55
CO <sub>2</sub> 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

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

399

400 It can be firstly observed that the introduction of the ASU involves a significant increase in BEC, in

401 spite of the small reduction of CO<sub>2</sub> capture section cost.

402 For comparative purposes, the same assumptions reported in [29] have been used in this work for  
 403 fuel purchasing, operating and maintenance (reported in table 8), eco-taxes and CO<sub>2</sub> emission  
 404 allowances (the latter based on a market price of 23 €/t by 2020, according to an assessment  
 405 published by Thomson Reuters [52]).  
 406

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 <sub>2</sub> 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
<b>Total O&amp;M</b>	<b>25.38</b>	<b>32.83</b>	<b>30.15</b>	<b>28.28</b>	<b>26.79</b>	<b>25.52</b>

407 Table 8a. Operating and maintenance costs (in €/MWh).  
 408

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 <sub>2</sub> 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
<b>Total O&amp;M</b>	<b>24.38</b>	<b>23.33</b>	<b>22.34</b>	<b>21.40</b>	<b>20.49</b>	<b>19.61</b>

409 Table 8b. Operating and maintenance costs (in €/MWh).  
 410

411 Finally, the CO<sub>2</sub> compression and transport costs have been assumed equal to 0.75 c€/kg and 2.5  
 412 c€/(t km), respectively [53,54], whereas an operating cost of 0.3 €/t has been considered for  
 413 sequestration in saline aquifer [54].  
 414

## 415 6. Economic assessment

416 Table 9 shows a summary of the economic performance of the air-blown, partial-oxy and full-oxy  
 417 configurations. A detailed definition of all the economic indicators can be found in Pettinau et al.  
 418 2017 [29].  
 419

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 <sub>2</sub> 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 <sub>2</sub> capture cost (€/t)	n.a.	44.13	40.35	36.98	34.54	32.59

CO <sub>2</sub> capture cost, present (€/t)	n.a.	24.16	23.90	22.77	21.91	21.23
CO <sub>2</sub> avoidance cost (€/t)	n.a.	60.07	54.78	50.06	46.64	43.88
CO <sub>2</sub> avoidance cost, present (€/t)	n.a.	32.89	32.45	30.83	29.59	28.59

420 Table 9a. Summary of economic performance.

421

Configuration	partial-oxy 50%	partial-oxy 60%	partial-oxy 70%	partial-oxy 80%	partial-oxy 90%	full-oxy 100%
Oxidant from ASU (% by weight)						
O <sub>2</sub> 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 <sub>2</sub> capture cost (€/t)	30.91	29.40	28.00	26.68	25.42	22.81
CO <sub>2</sub> capture cost, present (€/t)	20.64	20.13	19.65	19.20	18.77	17.80
CO <sub>2</sub> avoidance cost (€/t)	41.52	39.41	37.46	35.63	33.88	26.29
CO <sub>2</sub> avoidance cost, present (€/t)	27.73	26.98	26.29	25.64	25.02	20.51

422 Table 9b. Summary of economic performance.

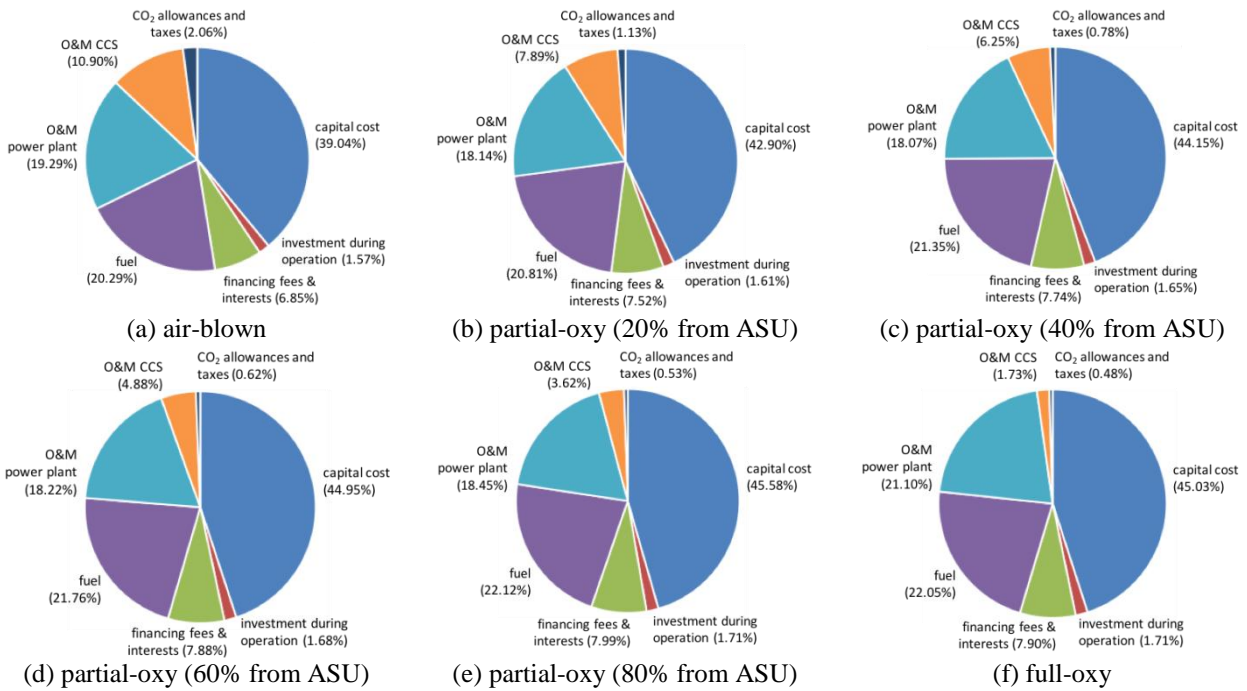
423

424 First of all, it is important to underline that LCOE for air-blown and full-oxy configurations are  
425 lower than the corresponding values obtained by the authors in a previous work (63.4 and 62.8  
426 €/MWh, respectively) [29]. The differences are due to the improvement of the steam cycle (a single  
427 reheat has been considered in the previous work) and, for full-oxy configuration, also to the higher  
428 annual availability of the plant (7,600 h/yr. vs. 7,000 h/yr. considered in the previous work).

429 The analysis of both cost of electricity (COE) and LCOE shows that, for new plants, full oxy-  
430 combustion is the more promising among the considered CO<sub>2</sub>-free power generation technologies,  
431 allowing for a significant reduction of LCOE with respect to conventional air-blown plants with  
432 post-combustion capture (55 €/MWh vs 60 €/MWh). On the other hand, partial oxy-combustion  
433 could be competitive, in terms of COE, with respect to air-blown plants (mainly with an oxygen  
434 enrichment higher than 40-50%), but it presents a higher COE than the full oxy-combustion  
435 technology. Considering that the full oxy-combustion is still quite far from commercial application  
436 (due to the relatively low experience on commercial-scale), partial oxy-combustion could be an  
437 option for short-term applications.

438 The comparison between COE and LCOE shows that the increase of capital costs with air  
439 enrichment has a higher impact than the decrease of operating costs. In facts, the former has a  
440 significant impact in LCOE behaviour (being paid during the first years of the project, their present  
441 values remain high), whereas operating costs (paid during the whole operating life) have a minor

442 impact on LCOE. Such a predominant increase of the influence of capital cost can be observed in  
 443 figure 9, which shows how each cost item impacts the LCOE. It can also be noticed that the impact  
 444 of the O&M costs of the CCS system significantly decreases with the increase of oxidant from  
 445 ASU.  
 446

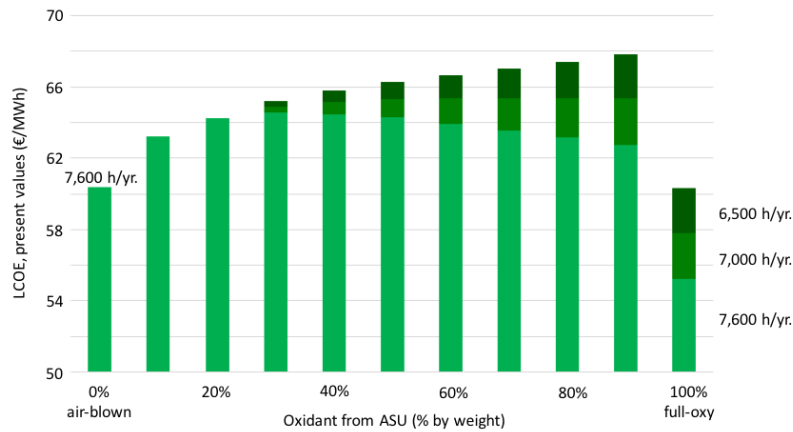


447 Figure 9. Impact of different costs on LCOE.  
 448

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

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

462



463

Figure 10. LCOE at different plant availability.

464

465

## 466 6. Conclusions

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

471 Such an enrichment involves a significant reduction of oxidizing agent flow: the full-oxy  
472 configuration needs almost the same oxygen amount of the air-blown one, which means about 22%  
473 of the whole oxidizing flow. As a consequence, a significant decrease of flue gas flow (from 416  
474 kg/s for air-blown to 119 kg/s for full-oxy) can be observed, due to the less dilution with nitrogen.  
475 The reference (without CCS) air-blown plant configuration shows a net power output of 466 MW,  
476 leading to a net efficiency of 46.6%. The integration with the CO<sub>2</sub> removal section reduces plant  
477 efficiency of 10.7 percentage points to 35.9%. The full-oxy plant configuration shows a net  
478 efficiency of 36.1%, very close to the air-blown case (with CCS).



479 Partial-oxy combustion still requires post-combustion chemical absorption CO<sub>2</sub> capture. In  
480 comparison to air-blown process, the higher concentration of CO<sub>2</sub> in flue gas with oxygen  
481 enrichment reduces energy penalization associated to solvent regeneration, but this reduction does  
482 not compensate for the sensible increase of the ASU energy consumption. Consequently, the plant  
483 net efficiency decreases with air enrichment from 35.9% (air-blown) to 31.5% (90% enrichment).  
484 Levelized cost of electricity is 60.39 €/MWh for the air-blown configuration and increases up to  
485 64.56 €/MWh for an air enrichment of 30% (i.e. 30% by volume of the oxidant agent comes from  
486 the ASU). Then, for high oxygen enrichments, LCOE decreases constantly and its value drops to  
487 55.21 €/MWh for the full-oxy configuration. The latter appears as the most promising technology  
488 for CO<sub>2</sub>-free power generation as soon as the experience at commercial-scale will allow the  
489 optimization of processes and materials.

490 It is important to underline that the reported results are a consequence of two key assumptions: (i)  
491 the same chemical absorption processes have been considered for both air-blown and partial-oxy  
492 configurations and the cryogenic CPU has been considered only for the full-oxy option; (ii) MEA  
493 has been considered as solvent, due to the wide availability of reliable data. A future work will be  
494 devoted to compare chemical absorption and cryogenic capture as decarbonization options for  
495 partial-oxy with high air enrichment and (on the basis of the results of an experimental campaign  
496 currently in progress) the possible advantages of using advanced solvents, such as mixtures of MEA  
497 and piperazine or MEA and potassium carbonate (K<sub>2</sub>CO<sub>3</sub>), both characterized by lower values of  
498 the specific thermal energy for the regeneration process.

499

## 500 **Acknowledgements**

501 This work has been done within the project “Modelling, experimentation and techno-economic  
502 analyses of pre-, post- and oxy-combustion CCS technologies for CO<sub>2</sub> emissions reduction from  
503 thermoelectric power plants” (Tender project, CUP: F78C13000550002), funded by the Regional  
504 Government of Sardinia.

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