

## DESIGN AND COST OPTIMIZATION OF CARBON CAPTURE FOR H-CLASS GAS TURBINE

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### ABSTRACT

Reaching carbon neutrality by 2050 will require the decarbonization of remaining power plants, i.e. CCGTs. CO<sub>2</sub> Capture and Storage will be part of the solutions, amine-based CO<sub>2</sub> capture process being today the most mature technology. However, there is still a need to further reduce the high energy penalty and capture cost of this process to enable its deployment. While a 90% capture efficiency was long considered as the techno-economic optimum, a higher capture rate seems to become the new target. In this study, a CO<sub>2</sub> capture plant (CCP) using 2-ethanolamine was modelled using Aspen plus applied to the flue gas of a new Combined Cycle Gas Turbine (H-Class gas turbine). This study aimed at evaluating several approaches (reboiler duty, flue gas temperature, and height of the absorber) to reach a higher capture rate and at assessing their impact on economics (capital and operational costs). This work showed that high capture rates are technically feasible if permitted by the plant design. However, all these strategies do not have the same impact: while some approaches enables to reach a 95% capture rate with limited impact on economics, an exponential effect is observed when going further 95-98%.

### INTRODUCTION

The reduction of CO<sub>2</sub> emissions is an urgent issue that needs to be addressed as the goal is to reach net-zero emissions by 2050 (Shu et al., 2023). Carbon capture and storage (CCS) is considered as one of the main solutions (McLaughlin et al., 2023; Shu et al., 2023; Zhang et al., 2023). Nowadays, one of the main sources of CO<sub>2</sub> emission is power plants (Ali et al., 2021; Zhang et al., 2023). Among the different technologies, absorption columns using an amine solution is the most mature process for capturing CO<sub>2</sub> from diluted flue gas (4-13% vol CO<sub>2</sub>) (Ali et al., 2021; Zhang et al., 2023). Today, opportunities in reducing capture

costs are expected with the economy of scales and learning by doing with the multiplication of projects (Kearns, 2021) While, only one large-scale CO<sub>2</sub> capture plant with a CO<sub>2</sub> capacity of 1 Mtpa is still in operation, an acceleration of CCS projects driven by the climate targets and price of CO<sub>2</sub> is witnessed for power application mainly in the United States, in the United Kingdom, and Asia with several ongoing FEED studies (Global status of CCS, 2022). With the increasing share of renewable energy in the future electricity mix, conventional power assets will still be required to ensure grid flexibility and reliability (Ceccarelli et al., 2014). To back-up renewables, future CCGTs will need to operate with high flexibility with an increased number of start-ups, shut-downs, and fast load transients (Blondeau and Mertens, 2019). Even if the increased efficiency of the new Combined Cycles Gas Turbines (H-Class) will contribute to reducing the CO<sub>2</sub> emissions, that will not be sufficient to reach carbon neutrality. Among the different decarbonization strategies currently evaluated as the use of green fuels (H<sub>2</sub>, biomethane (Ceccarelli et al., 2014; Wu et al., 2021), Carbon capture storage (CCS) applied to new-built combined cycle gas turbines (CCGT) (Sterkhov et al., 2021) is part of the different scenarios. However, key challenges remain besides the high energy penalty and capture costs (Ceccarelli et al., 2014; Qureshi et al., 2021) bringing the challenge of mass transfer, resulting in larger equipment (i.e. flue gas cooler, absorber, heat exchangers) and significant land footprint issues. Moreover, flue gas from natural-gas-fired Power plants contains a high O<sub>2</sub> concentration (10-12% vol) which could pose a risk for solvents sensitive to oxidative degradation and thus related issues (higher solvent consumption, emissions, corrosion, waste handling, foaming). This may result in higher Capex and increased solvent consumption per ton of CO<sub>2</sub> captured compared to coal-fired power plants (Yun et al., 2021; Beiron et al., 2022). In addition, the CCGT-CCS plant

should be able to follow the CCGT flexible operating profile with varying operating loads and frequent start-ups/shut-downs while minimizing residual CO<sub>2</sub> emissions during transient phases.

While a 90% capture efficiency at stable load conditions was long considered the optimum capture rate, a higher capture rate seems to be the new target for future Carbon capture demonstration projects under development in UK (Gibbins and Lucquiaud, 2022). While today there are no legal requirements for a minimal capture rate, a higher capture rate can be seen as an effective approach to reach carbon neutral power production rather than purchasing Carbon offset credits as well as a way to compensate for the reduced capture efficiency during transient phases and start-up of CCGT but also during peak electricity demand when there is a need to maximize power production.

High capture levels are technically feasible if permitted by the plant design and solvent selection. Different parameters can influence the overall CO<sub>2</sub> capture rate, either by increasing the solvent absorption capacity or by enhancing CO<sub>2</sub> desorption (Michailos and Gibbins, 2022). However, all these strategies do not have the same impact on capture costs (Brandl et al., 2021).

The objectives of this study are to assess the impact on Levelized Cost Of Electricity (LCOE) through modeling, when implementing a CO<sub>2</sub> capture plant to a CCGT H-class turbine and to evaluate the impact of the different approaches to reach a higher capture rate on capture costs (Capex and Opex) and LCOE. The different strategies evaluated include variation of flue gas temperature, reboiler duty, and height of the absorber and their respective impact on capture cost.

## METHOD

### Carbon capture plant design

A CO<sub>2</sub> capture plant applied to the flue gas of a CCGT Class H has been simulated using Aspen V12.1. It consists of three sections, a flue gas cooling (FGC) tower often referred to as a scrubber, a CO<sub>2</sub> capture section, and compression of CO<sub>2</sub> for transport and storage. (Figure 1) The CCP has been designed as one single train configuration (1\*100% scrubber – absorber – stripper) treating 100 % total flue gas and operating at full and base load conditions. Other strategies could be considered with a multi-train approach but these are not covered as part of this study.

The specifications of the flue gas characteristics from the CCGT used for the simulation are summarized in Table 1. For this high-level study, the impact of impurities present in the flue gas (NOx) has not been considered. Table 2 presents the general information on the design parameters of the Capture plant. It has to be mentioned that for better comparison and confidentiality reasons, the data are presented in relative compared to the base case (CCGT with 90% CO<sub>2</sub> capture).

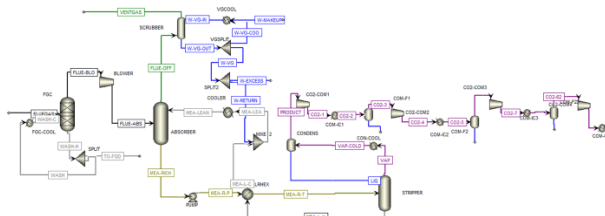


Figure 1. Flowsheet of the CO<sub>2</sub> capture plant for a CCGT class H-turbine simulated using Aspen plus V12.1 using 30 wt% MEA (2-ethanolamine) and 90% capture rate as base case

Table 1. Inlet flue gas composition

Parameter	Units	Value
<b>Flue gas flow rate</b>	kg/s	1054
<b>Flue gas pressure</b>	mbar g	0
<b>Flue gas composition</b>		
<b>N<sub>2</sub></b>	%vol	73.8
<b>O<sub>2</sub></b>	%vol	10.6
<b>CO<sub>2</sub></b>	%vol	4.8
<b>H<sub>2</sub>O</b>	%vol	9.9
<b>Ar</b>	%vol	0.9

### Modeling of flue gas conditioning section

A flue gas pre-treatment is typically required to cool down the gas temperature before absorber and remove impurities (SO<sub>x</sub>, NO<sub>x</sub>, acid gases) from the flue gas: at absorber inlet, the flue gas should be saturated at approximately 30 to 50°C. A higher temperature will reduce the capture rate and affect the overall water balance.

A scrubber also called Direct Contact Cooler (modelled using a RedFrac block) was considered in this study to cool down and saturate the flue gas at the defined temperature. No addition of caustic was foreseen considering the very low expected SO<sub>x</sub> concentration in the flue gas from the natural gas-fired Power Plant (< 0.1 mg/Nm<sup>3</sup> at 15% O<sub>2</sub> ref).

The design specifications of the scrubber and blower are summarized in Table 2. To compensate the pressure loss from the DCC and absorber columns, a blower is installed after the FGC column, which raises the pressure by 0.1 bar. Due to the work energy of the blower, the flue gas temperature rises by 10°C. The extra water from flue gas is removed from the scrubber to be processed in the Waste Water Treatment plant.

Table 2. Design information of the CO<sub>2</sub> capture plant

Equipment	Specifications	Value
<b>Scrubber</b>	Cooling substance	Pure water (20°C inlet)
	Packing	Sulzer 250Y
	Height of packing	3 m
	Column diameter	22 m
<b>Blower</b>	Type	Isentropic
	Outlet pressure	1.13 bar
	Isentropic efficiency	85%
<b>Absorber</b>	Packing	Sulzer 250Y
	Height of packing	17 m
	Column diameter	22 m
	Lean solvent flowrate	3328 kg/s
	Solvent	MEA 30 w%
	Solvent inlet temperature	45°C
	Gas inlet temperature	47°C
<b>Stripper</b>	Packing	Sulzer 250Y
	Height of packing	3.2 m
	Column diameter	9 m

### CO<sub>2</sub> absorption section

The CO<sub>2</sub> capture part consists of an absorber column, heat exchanger stripper column, and pumps. Both absorption and stripper columns are modeled using a RadFrac block and using a rated-based model for calculations. Aqueous 30 wt% Monoethanolamine (MEA) was used as the benchmark solvent. MEA has a high absorption rate, low solvent cost, and low molecular weight which makes it a good candidate for CO<sub>2</sub> absorption, (Mofarahi et al., 2008). However, it has some disadvantages that cannot be neglected, such as low CO<sub>2</sub> loading, the high energy requirement for regeneration, losses due to high vapor pressure, high viscosity, and a lack of chemical stability (oxidative degradation, formation of Heat Stable Salts by reaction with NO<sub>x</sub> and SO<sub>x</sub>, thermal degradation) which leads to operational and environmental issues (fouling, foaming, emission, wastes handling, corrosion). These shortfalls can be partially avoided by choosing optimized conditions in the process.

The reactions and the kinetic data of the MEA and CO<sub>2</sub> are entered separately for the absorber and stripper columns, as described below. The reactions are governed by power-law expressions whose kinetic coefficients are given in Table 3. On the other hand, equilibrium constants for equilibrium reactions are calculated from the standard Gibbs free energy change.

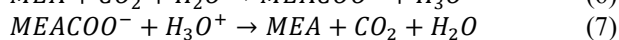
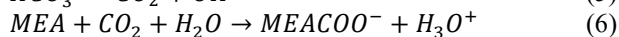
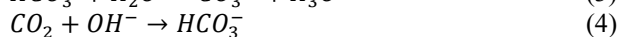
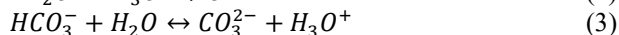
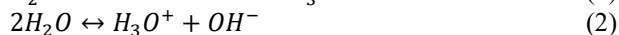
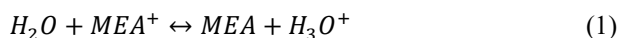


Table 3. Kinetic data of the reactions (Garcia et al., 2017; Ramezani et al., 2021)

Reaction No.	Reaction type	K	E (cal/mol)
4	Kinetics	4.32x10 <sup>13</sup>	13249
5	Kinetics	2.38x10 <sup>17</sup>	29451
6	Kinetics	9.77x10 <sup>10</sup>	9855.8
7	Kinetics	3.23x10 <sup>19</sup>	15655

A lean-rich cross-heat exchanger is used to pre-heat the rich solvent stream entering the stripper column by using heat from the hot lean solvent stream coming from the stripper itself. As the absorption process is favored by low temperature, a cooler is added to further decrease the lean solvent temperature.

Due to its vapor pressure and mechanical losses (droplets), some MEA is released into the vent gas. Therefore, a water wash section is implemented to recover MEA and to cool down the treated gas. The same diameter is used for the water wash section as for the absorber, as in reality the water wash would be a section of the absorber column, while in Aspen, due to numerical complexity, it is simulated as another RadFrac column.

Solvent and water makeup are required to close the liquid inventory in the CO<sub>2</sub> capture loop due to losses in the treated flue gas and CO<sub>2</sub> product.

### Compression

After the CO<sub>2</sub> is stripped out of the rich solvent, it is compressed to 30 bar and 40°C. The CO<sub>2</sub> at the stripper overhead is first cooled down with condensates that are sent back to the top of the stripper column. CO<sub>2</sub> is then compressed with interstage coolers. After each compression step, a cooler and a flash drum is implemented to separate the condensed water.

### Assumptions made for economic calculations

The equipment costs and utilities consumption of the CCS plant are calculated using Aspen Economics with process data extracted from Aspen plus. All assumptions and economic indicators used for economic evaluation are summarized in Table 4.

The Capex is determined using the equipment cost that was estimated by Aspen economics, with a Lang factor equal to 4.05, according to Equation 8:

$$\text{Capex} = \text{lang factor} * \text{total equipment costs} \quad (8)$$

The operational costs consist of both fixed and variable Opex. The variable charges include the costs of the chemicals (mainly solvent), utilities (electricity, steam and cooling water), and waste treatment. The fixed costs include costs related to maintenance, taxes and insurance, investment costs (7% included), and depreciation (20 years) (A. Chauvel, 2003). The main utilities for the CO<sub>2</sub> capture plant are energy required for the regeneration of the solvent in the stripper as Low Pressure steam, electrical power (mainly for pumps, blower and CO<sub>2</sub> compression) and solvent consumption. A solvent consumption rate of 1 kg/t<sub>CO2</sub> has been used to compensate solvent losses by evaporation, mechanical losses and degradation. According to the IEA GHG (2019), the solvent consumption ranged from 0.1 to 1.6 kg/t<sub>CO2</sub> during pilot testing campaigns using proprietary solvents. The specific solvent consumption will depend on several parameters such as solvent properties (volatility, stability), flue gas composition (% O<sub>2</sub> and impurities), and process design.

To determine the LCOE when implementing a CO<sub>2</sub> capture plant to a CCGT, the extra power and steam consumption from the CCP was translated into a loss of CCGT net efficiency. A lifetime of 20 years was assumed with 4500 and 6500 hours of operation per year for respectively the CCGT and the CCGT-CCS: it is assumed that in the future the CCGT equipped with a CCP will be operated in a preferential manner to support grid stabilization.

Table 4. Key assumptions and economic indicators

<b>Capex</b>	Class 5 (+/- 50%)
<b>Plant Life time</b>	20 years
<b>Yearly operating hours</b>	
CCGT w/o CCS	4500 h/year
CCGT with CCS	6500 h/year
<b>Opex</b> (Intergovernmental Panel on Climate Change, 2014)	
LP Steam	15 €/MWh
Electricity	100 €/MWh
Solvent (MEA)	1 €/kg
Demin water	2.5 €/m <sup>3</sup>
Cooling water	0.02 €/m <sup>3</sup>
Waste water treatment	2 €/t
Waste incineration	400 €/t
CO <sub>2</sub> transport and storage	40 €/t <sub>CO2</sub>
CO <sub>2</sub> ETS	70 €/t <sub>CO2</sub>
Fuel cost	30 €/MWh
<b>Weighted Average Cost of Capital</b>	<b>8%</b>

## RESULTS AND DISCUSSION

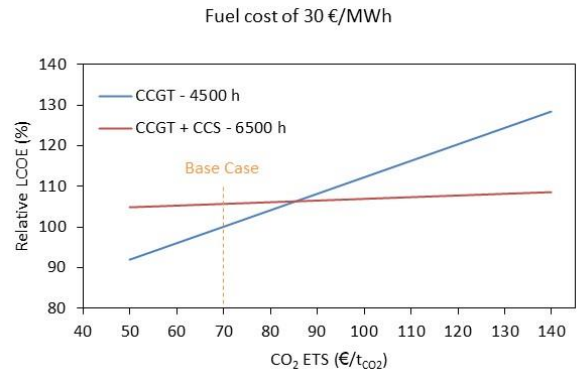
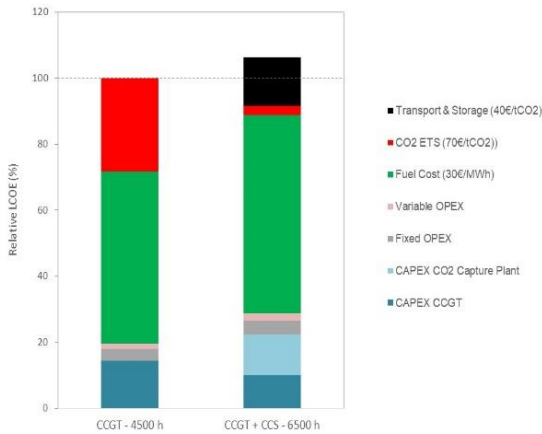
### Base case

The Base Case plant was designed according to the technical specifications described in the previous section with a capture rate of 90% and has been used as a reference to benchmark the different strategies to increase the capture rate. The economic evaluation is summarized in Table A.1. Implementing a CO<sub>2</sub> capture plant results in large Capex (491 M€) due to large columns (flue gas scrubber/absorber). The Opex is dominated by high energy requirements: low-pressure steam for solvent regeneration and electrical power for blowers, pumps, and CO<sub>2</sub> compression.

Figure 2 compares the Levelized Cost of Electricity (LCOE) for a CCGT class-H without and with CCS. As shown in Figure 2a, the implementation of CCS doubles the Capex share in electricity production costs. Moreover, the electrical power consumption and low-pressure steam requirements of the CO<sub>2</sub> capture plant reduce the CCGT net efficiency. This translates into a higher fuel cost per MWh. Finally, the cost related to the remaining CO<sub>2</sub> emissions will be dictated by both the future CO<sub>2</sub> ETS price and the CO<sub>2</sub> capture efficiency. Today, it is still unclear what would be the future cost of CO<sub>2</sub> capture for CCGTs as high uncertainties remain on several key influencing parameters, i.e. ETS CO<sub>2</sub> costs, natural gas price, number of operating hours of future CCGTs but also transport and storage costs as project specific. Figure 2.b illustrates the influence of several parameters on the capture cost and thus on LCOE.

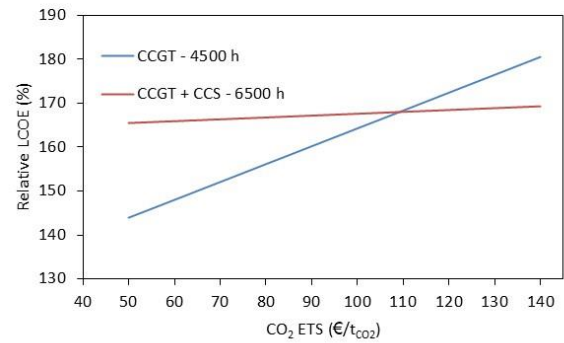
In particular, in the current uncertain geopolitical context, two parameters, i.e. the fuel cost and CO<sub>2</sub> ETS, have a considerable impact on LCOE and drive the economic viability of CCS applied to CCGT. As illustrated in Figure 3, the fuel costs and CO<sub>2</sub> ETS price drive the break-even point where CCGT-CCS can become competitive. The authors emphasize that the graphs should be read with caution as the exact threshold values are specific to the economic assumptions considered in the present study.

a)



c)

Fuel cost of 60 €/MWh



b)

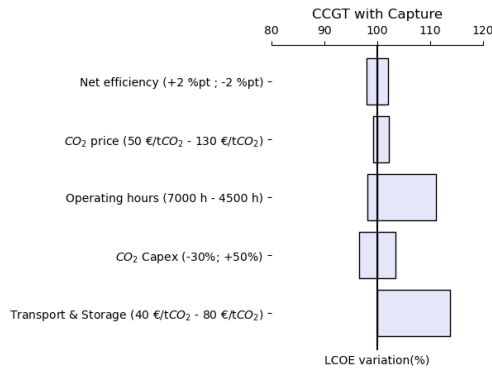
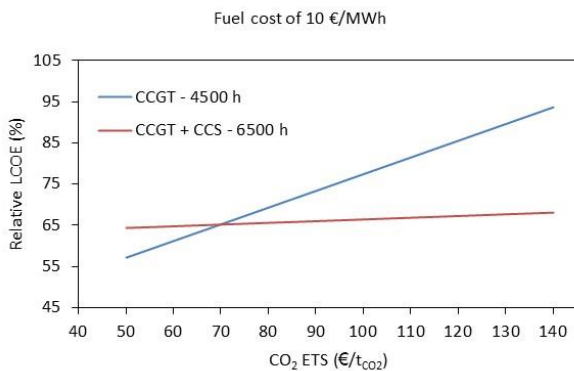


Figure 2. a) Levelized Cost Of Electricity for CCGT-Class H without and with CCS. B) Sensitivity analysis – effect of different parameters on LCOE for the case of CCGT+CCS (index 100 is the result of the base case condition).

a)



b)

Figure 3. Relative variation of the LCOE for a CCGT without and with CCS in function of CO<sub>2</sub> ETS price and fuel cost (a) 10€/MWh – (b) 30€/MWh and (c) 60€/MWh. All the values have been normalized based on the results from the base case study at 90% capture rate.

### Impact of higher capture rate on capture costs

During transient phases and the start-up or shut-down process, the percentage of CO<sub>2</sub> that is captured is potentially lower than during stable full-load operation. To compensate for this reduced capture efficiency, one could increase the overall CO<sub>2</sub> capture rate at stable load, either by increasing the solvent absorption capacity (i.e. CO<sub>2</sub> rich loading) or by enhancing CO<sub>2</sub> desorption (i.e. CO<sub>2</sub> lean loading). Changing the solvent loading has an impact on both Capex and Opex. The effects of reboiler duty, flue gas temperature and absorber height are described in the next paragraphs.

### Variation of the reboiler duty

Higher reboiler duty directly affects the regeneration of the solvent and reduces the CO<sub>2</sub> loading of both lean and rich solvents and as a consequence increases the CO<sub>2</sub> capture rate of the process. As represented in Figure 4, high capture rates up to 98% are achievable with a limited impact on Capex (+5%) explained by the slightly oversized reboiler and CO<sub>2</sub> compressor. On the other hand, the highest steam demand from the CCGT plant results in CCGT efficiency loss of 7.6 % and 7.9% point of efficiency at respectively 95% and 98% capture rates. However, the non-linear profile

of the capture rate with steam input suggests that after a threshold, the relative energy demand increases exponentially to increase the capture rate. Based on current results, increasing the capture rate up to 95% would be achievable with a relatively limited cost (energy demand) increase.

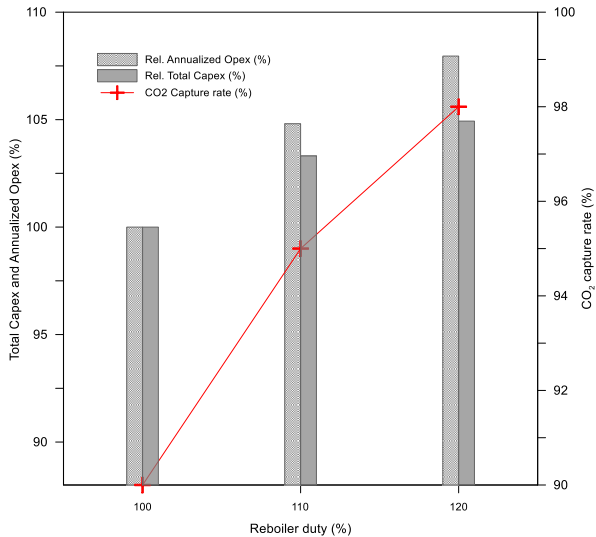


Figure 4. Impact of increased reboiler duty on Capture rate (right axis) and Total Capex / Annualized Opex (left axis)

#### Variation of packing height

Figure 5 shows the effect of the absorber's packing height on the capture rate and costs. Increasing the absorber packing height improves the gas/liquid contact which in turn increases the CO<sub>2</sub> absorption capacity of the solvent. Based on the current design with MEA as a solvent, increasing the absorber packing height did not significantly improve the CO<sub>2</sub> absorption capacity: a maximum increase of +4% on the capture rate was only achieved while the packing height was multiplied by 3. On the other hand, this led to a steep increase in Capex (+21%) due to the bigger amount of packing materials, as packing and internals are one of the main costs of the absorber column. The Opex increase by 11% is explained by the higher fixed operational costs linked to higher investment costs. Thus, based on the studied configuration with 2-ethanolamine (MEA), increasing the packing height proved not to be an attractive solution as it had a significant impact on economics of the plant while providing a limited improvement in plant performance.

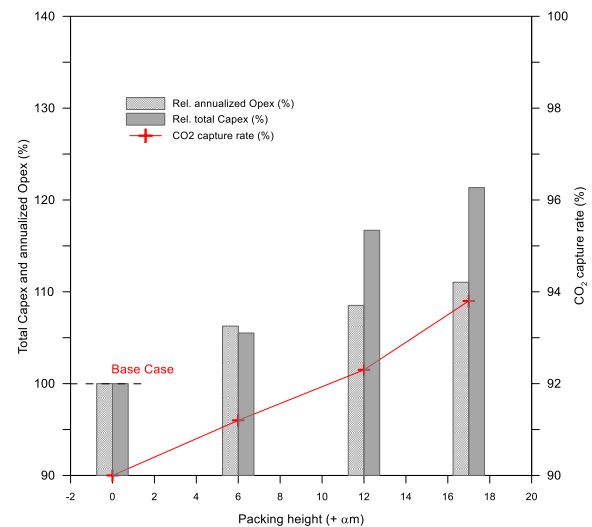


Figure 5. Impact of increased packing height of absorber on Capture rate (right axis) and Capex / Opex (left axis)

#### Variation of absorber inlet temperature

Previous studies have evaluated the effect of temperature on the MEA-CO<sub>2</sub> mass transfer efficiency [24]: these showed that while kinetics are favored when increasing the temperature, above a temperature threshold (40-45°C) the CO<sub>2</sub> absorption efficiency decreases due to the thermodynamics of CO<sub>2</sub> absorption and decrease in CO<sub>2</sub> solubility.

In the present study, the flue gas temperature at absorber inlet was varied from 30 to 50°C to determine the impact on the CO<sub>2</sub> absorption rate in the absorber and thus related capture cost. It is worth mentioning that these results are not directly applicable to other solvents as other physical properties are involved (risk of crystallization, viscosity). As presented in Figure 6, at fixed reboiler duty, reducing the inlet flue gas temperature from 47°C to 30°C enables to increase the capture rate from 90 to 93% with a non-linear profile most probably explained by this opposite effect of temperature for CO<sub>2</sub> absorption: improved kinetics at higher temperature vs improved thermodynamics at reduced temperature. This translates on economics in an optimum of the annualized Opex when the flue gas is introduced at absorber inlet at about 40°C. For an inlet flue gas temperature of 30-35°C, the higher Opex due to the increased cooling duty and use of electricity are not compensated by the improved capture efficiency. On the opposite side, when the flue gas is increased at 47-50°C, the lower power consumption of CO<sub>2</sub> compressor is not compensated by the higher power consumption of the blower, liquid pumps and increased cooling duty in the water wash section due to respectively higher flue gas flow rate and temperature of the treated flue gas upstream the absorber. Thus, based on the current results and MEA solvent, an inlet flue gas temperature at 40°C seems to present the best compromise to reach a higher CO<sub>2</sub> capture rate.

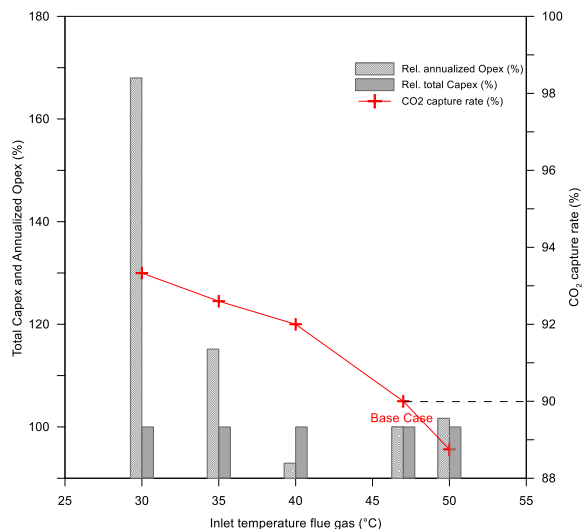


Figure 6. Impact of absorber inlet flue gas temperature on capture rate (right axis) and Capex / Opex (left axis)

### Impact of higher capture rate on LCOE

From the different configurations evaluated in the previous sections, the optimum conditions were compared to the base case: 95% capture rate with increased reboiler duty and 92% based on an absorber inlet flue gas temperature of 40°C. The LCOE for these cases is presented in Figure 7: based on the current assumptions, the present study confirms that a higher capture rate is achievable with a limited impact on cost. In particular at 95%, the highest share of fuel costs due to the highest steam demand is counterbalanced by a lower cost linked to remaining CO<sub>2</sub> emissions. However, the actual difference will be driven by the future costs of CO<sub>2</sub> ETS price, fuel cost and CO<sub>2</sub> transport and storage cost.

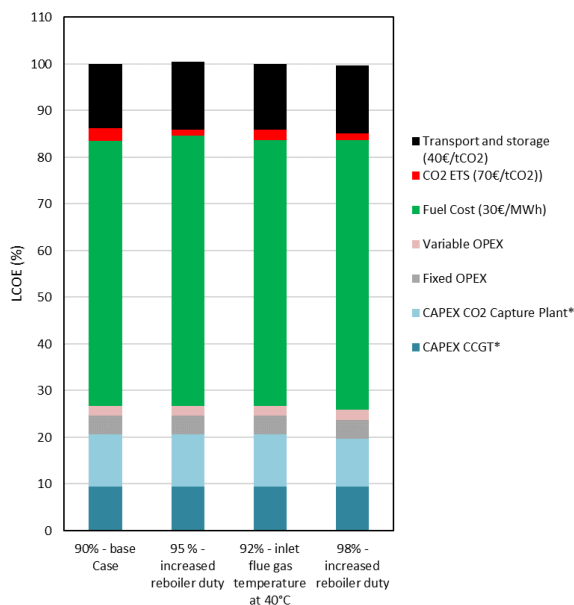


Figure 7. Impact of different strategies to increase the reboiler duty on LCOE

## CONCLUSION

In this work, a capture plant for CO<sub>2</sub> emission reduction from a CCGT power plant with an H-class gas turbine was simulated with Aspen Plus. A reference CO<sub>2</sub> Capture plant (Base Case) was first studied with a capture rate of 90%, and has been evaluated economically. It was found that the Capex in the CCS plant is related to columns, and compression. For Opex the major contributors are reboiler duty, and electricity consumption for CO<sub>2</sub> compression. These were translated in the electricity production costs by doubling the CAPEX share and increased the fuel costs per MWh due to the CCGT efficiency loss.

As described in literature, achieving higher capture rates seems to be a strategy to compensate for reduced CO<sub>2</sub> capture efficiency during transient phases and start-up /shutdown. Different strategies (solvent, advanced design, process optimisation) can be applied to increase the capture rate. This study evaluated the impact of several approaches (reboiler duty, absorber height, flue gas temperature) on the capture rate and related costs (opex and capex). Based on the current assumptions, not all of these strategies were found to be economically favorable. Thus, among the different configurations, increasing reboiler duty by 10% and cooling down the flue gas to 40°C enabled to increase the capture rate respectively to 95% and 92% with relatively limited impact on capture cost. On the other hand, increasing the height of the absorber packing proved not to be an attractive solution as it had a significant impact on economics of the plant while providing a limited improvement in plant performance. In the end, optimising the configuration will require adjusting not just one but several parameters.

Determining the optimal design and operation of the CO<sub>2</sub> Capture plant for minimum cost, will depend on the choice of solvent, site specificities (availability of steam and cooling) but also on external economic drivers (i.e. electricity cost, ETS price, Carbon offsets credit).

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## APPENDIX

Table A.1 Summary of the results.

	Base Case	Reboiler duty		Absorber height			Inlet absorber temperature			
		110%	120%	+6 m	+12 m	+17 m	30°C	35°C	40°C	50°C
CO <sub>2</sub> capture rate	90%	95%	98%	91.2%	92.3%	94%	93.3%	92.6%	92%	88.7%
CO <sub>2</sub> plant capacity (Mtpy)	1.62	1.71	1.77	1.64	1.66	1.7	1.69	1.67	1.66	1.66
Capex (M€)	491	507	514	517	572	595	491	491	491	491
Annualized Opex (M€)	98.8	103.54	106.65	104.98	107.21	109.7	166	113.77	91.82	100.46
Steam consumption (GJ/tCO <sub>2</sub> )	3.53	3.69	3.84	3.48	3.44	3.39	3.40	3.43	3.45	3.58
Power consumption (kW/tCO <sub>2</sub> )	16.75	16.45	16.21	16.4	16.33	16.27	16.27	16.31	16.35	16.55