Carbon Capture Performance Assessment Applied to Combined Cycle Gas Turbine Under Part-Load Operation

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ABSTRACT

The growing share of renewable energies in our electricity production, together with the still lacking storage capacity, strongly reinforces the need for more flexible electricity production units. In this context, Combined Cycle Gas Turbines (CCGTs) have a role to play, both in the current and future electricity production system due to their high efficiency, high load flexibility, and low CO₂ emissions compared to other conventional thermal power plants. Nevertheless, bearing in mind our current challenges concerning climate change, the CO₂ emissions of these CCGTs need to be reduced drastically. The amine-based absorption Carbon Capture (CC) process is currently the most mature and applicable CC technology. This process is known to require a considerable amount of energy, degrading the plant performance. However, to back-up renewable production, CCGTs will operate most of the time under part-load conditions. The impact of these part-load operations on the CC is still relatively unknown. In this context, this study aims to assess the performance of the CC process applied to a typical CCGT under part load operation using specific simulation models. The CC plant model has been successfully validated against experimental data from a pilot-scale capture facility. Then, the CC plant has been scale-up to the CCGT scale and the process has been optimized for each operating condition. The simulation results show that the specific reboiler duty increases for part-load operation, while the specific cooling requirements decrease. Moreover, the analysis of the yearly CCGT operation highlights a relative increase in CC energy penalty of 21% for an annual CCGT load factor of 0.5, impacting significantly plant performance. The next step will involve reducing this energy penalty.

NOMENCLATURE

Acronymns
CAPEX Capital Expenditure
CC Carbon capture
CCGT Combined Cycle Gas Turbine
EGR Exhaust Gas Recirculation
LCOE Levelized Cost of Energy

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INTRODUCTION

To fulfil the objectives set by Paris Agreements, the signatory countries push the deployment of renewable energy to decarbonise their electricity production. In 2020, renewable energy sources produced 38% of the European electricity production, including 14% and 5% wind and solar production, respectively [1]. To reach 55% reduction in greenhouse emissions by 2030, the wind and solar generation must grow by 100 TWh per year between 2020 and 2030, while the yearly average growth was 38 TWh between 2010 and 2020 [1]. To deal with this increasing share of intermittent electricity production, the use of backup power plants is required to balance the overall electricity production due to the lack of large-scale electricity storage capacity. Combined Cycle Gas Turbines (CCGTs) are good candidates to be used as renewable back-ups due their high efficiency (65%), high operational flexibility and relatively low CO₂ emissions (350 kg/MWhe) compared to other conventional thermal plants [2]. However, to achieve the emission reduction targets in 2030 and carbon neutrality in 2050, even CCGTs need to be decarbonised.

Amines-based post-combustion Carbon Capture (CC) is the most promising technology to retrofit existing power plants into low-carbon power plants [3]. Indeed, this CC technology, which is commercially mature (Technology Readiness Level (TRL) of 9 [4]), can be implemented without modifying the power plant as this is an end-of-pipe technology. However, the low CO₂ content in exhaust gas coming from gas-fired power plants is less favorable from a CC point of view. Moreover, the large flue gas flow rate requires a large capacity capture facility, increasing the Capital Expenditure (CAPEX) of the plant [5]. The performance of a CCGT coupled with a CC plant has been evaluated by several researchers through simulation and modelling. Hetland et al. [6] designed and modelled a post-combustion CC for a 450 MW offshore CCGT. The CC specific energy consumption has been evaluated at 3.77 GJ/tCO₂, reducing by 9 absolute percentage points the plant efficiency. Soltani et al. [7] showed that 3.98 GJ/tCO₂ are required to capture CO₂ from a 600 MW CCGT. Finally, Canepa et al. [5] demonstrated that the CCGT power output was reduced from 240 MW to 209.4 MW by applying CC, while Biliyok et al. [8] observed a decrease of 15% in power output for a 440 MW CCGT.

Nevertheless, most of the studies in literature focus on the CC performance assessment for base load CCGT operation. However, with the deeper penetration of renewable energies, CCGTs will increasingly be used as peak units. Therefore, CCGTs will run less frequently on base load, increasing the CCGT part-load operation time. Moreover, the number of transient phases, such as start-ups, shut-downs and fast load changes, will drastically increase. Hence, this could impact the CC, which is commonly designed for full-load CCGT operation, as this operating mode produces the highest flue mass flow rate. Some studies have assessed the performance viability of an CCGT plant with an integrated carbon capture at part-loads. Jordal et al. [9] demonstrated that the steam supply from the power plant to the CC unit is sufficient for solvent regeneration to maintain a capture efficiency of 90% at full load, partial load and during load changes. Rezazadeh et al. [10] showed the CC steady-state performance is viable at part-loads down to 60%. Alcaráz et al. [11] has investigated the viability of a CCGT with CC and EGR (Exhaust Gas Recirculation) and showed that EGR benefit on the plant efficiency is reduced at part-load compared to full-load CCGT operation. Adams et al. [12] observed an energy penalty of 8.64% at full load and 7.93% at 60% load when CC is applied to a CCGT. However, there is a limited amount of information available on the real impact of off-design operation on the profitability of the CC plant when CC is applied. However, the real impact of partial loads on carbon capture is still relatively unknown highlighting the strong need to assess the CC performance under real operating conditions.

This study aims to evaluate, through steady-state thermodynamic cycle modelling, the performance of the CC process applied to a typical CCGT power plant under various operating conditions, i.e. at full and partial loads. Moreover, the impact of realistic CCGT operating conditions over a year on CO₂ emissions and on the profitability of the plant will be assessed, providing useful insight for a thorough economic analysis. To this end, an absorption-based capture facility, using monoethanolamine (MEA) as solvent, is designed on the basis of full load CCGT operation. The integration between the CO₂ capture and the CCGT usually requires some modifications in the original CCGT cycle. Indeed, the heat for the CC process can be supplied by steam extracted from the steam cycle and Exhaust Gas Recirculation (EGR) can be applied on the gas turbine cycle to increase the CO₂ content in flue gas and reduce its mass flow rate. However, in this work, we consider the CC plant as an end-of-pipe unit without modifying the CCGT cycle for the purpose of retrofitting existing power plants, allowing to focus only on the CC perfor-
mance. An optimised heat integration between both processes and advanced cycles using EGR, as well as a study of the impact of transient CCGT operations on the CC through dynamic models, will be considered in future works.

The first section of the paper covers the description of the simulation approach, including CCGT and CC modelling, as well as the validation of the CC model on a pilot-scale capture plant. The next section includes the description of the scale-up methodology of the CC plant to a large-scale CCGT application, as well as the discussion on the process optimisation. The results of the simulation are presented, giving an overview of the impact of CCGT partial-load operation on the CC performance. Different realistic CCGT operational yearly cycles are also considered, allowing for the analysing and comparison of the CC plant as well as global CCGT performance. Finally, a conclusion summarizing the main findings of the study is presented, as well as the perspectives for future works.

SIMULATION APPROACH

This section describes the modelling of the CCGT and CC plants. The validation of the CC process with experimental data from pilot-scale plant is also presented.

Combined Cycle Gas Turbine modelling

A typical 580 MW CCGT power plant has been modelled using the Thermoflow software [13]. The plant consists in a typical H-class gas turbine and three pressure level-steam with reheat. The CCGT has been simulated under full and part load conditions, varying from 100% to 30% by step of 10%. These models have been used to obtain flue gas properties for each CCGT load.

Carbon capture modelling and validation

The CC process, considered in this study and illustrated in Fig. 1, is a conventional amine-based absorption-regeneration process composed of two packed columns: an absorber and a stripper. The hot flue gases coming from the CCGT plant are cooled down to 40 °C before being fed to the bottom of the absorption column (1), while the liquid solvent (lean solvent), an aqueous solution of 30 wt.% of MEA, is fed at the top. In the absorber, the downward liquid solvent, by chemically interacting with the upward gas phase, absorbs the $CO_2$ from flue gas and exits at the bottom with a higher $CO_2$ content (rich solvent), while the flue gas is released at the top with a lower $CO_2$ content depending on the capture rate (2). The absorption process typically occurs at 40 °C and at atmospheric pressure. Then, the rich solvent is pumped (3) to the stripper after being preheated into a counter-flow heat exchanger (4) by the lean solvent exiting the stripper column at higher temperature. The rich solvent is injected at the stripper top, while steam is produced in the reboiler at the stripper bottom (6). Into the stripper (5), the $CO_2$ is removed from the downward solvent by chemically interacting with the upward steam. Then, the vapor stream at the top of the stripper is condensed to separate $CO_2$ from water (7). The conventional regeneration process is performed at high temperature (105-110°C) and under a pressure of 1.2 bar. The captured $CO_2$ is generally compressed to be transported and geologically stored or valorized in another process (e.g. conversion to synthetic fuels). The regenerated solvent exits the desorber bottom, is pumped through the counter-flow heat exchanger (8) and is cooled to 40 °C (9) before entering the absorber top. Moreover, at the top of the absorption and regeneration columns, a water-wash process is usually installed to remove amine-solvent residues carried by the flue gas or the $CO_2$ stream. This washing process has not been considered in this work due to its limited influence on the absorption-regeneration performance.

The CC plant modelling has been based on the Pilot-scale Advanced Capture Technology (PACT) from the UK Carbon Capture and Storage Research Centre (UKCCSRC) [15]. The PACT columns specifications are reported in Table 1. The model has been developed in Aspen Plus v12. The thermodynamics models used are the Electrolyte Non-Random Two Liquid (EN-RTL) model for electrolyte properties in liquid phase and the PC-SAFT equation of state for vapor phase properties. The model properties were taken from the work of Zhang et al. [16].

![FIGURE 1: The considered CC plant is a conventional amine-based absorption-regeneration process](image)

<table>
<thead>
<tr>
<th>TABLE 1: PACT COLUMNS SPECIFICATIONS [14]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Absorber</td>
</tr>
<tr>
<td>Diameter [m]</td>
</tr>
<tr>
<td>Packed height [m]</td>
</tr>
<tr>
<td>Packed type</td>
</tr>
</tbody>
</table>

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in which these thermodynamic models have been validated for CO$_2$ absorption in aqueous MEA solution against experimental values. Both absorber and stripper are modelled with RadFrac distillation units, which are rigorous simulations of distillation columns. The chemical reactions occurring in the columns are defined as:

1. $$2\text{H}_2\text{O} \leftrightarrow \text{H}_2\text{O}^+ + \text{OH}^-$$
2. $$\text{HCO}_3^- + \text{H}_2\text{O} \leftrightarrow \text{H}_2\text{O}^+ + \text{CO}_2^{2-}$$
3. $$\text{MEA}^+ + \text{H}_2\text{O} \leftrightarrow \text{MEA} + \text{H}_2\text{O}^+$$
4. $$\text{CO}_2 + \text{OH}^- \rightarrow \text{HCO}_3^-$$
5. $$\text{HCO}_3^- \rightarrow \text{CO}_2 + \text{OH}^-$$
6. $$\text{MEA} + \text{CO}_2 + \text{H}_2\text{O} \rightarrow \text{MEACOO}^- + \text{H}_2\text{O}^+$$
7. $$\text{MEACOO}^- + \text{H}_2\text{O}^+ \rightarrow \text{MEA} + \text{CO}_2 + \text{H}_2\text{O}$$

The equations (1) to (3) describe rather fast reactions assumed to be in chemical equilibrium, while equations (4) to (7) describe reactions involving chemical kinetics.

Both absorption and stripping processes have been modelled using the rate-based approach, offering a more accurate resolution than the equilibrium-stage modeling approach [17]. The HanleyIMTP model has been applied for the mass transfer and the interfacial area calculation [18]. This calculated interfacial area is multiplied by an interfacial area factor to determine the required area for mass transfer. This interfacial area factor has been set to match with experimental results for both columns. Finally, the absorption and stripping columns have been divided in 30 calculations stages in the simulations.

The CC model has been validated against experimental values from Agbonghae et al. [14] for two specific operating point, defined by their Liquid to Gas ratio (L/G), which is the ratio between solvent and flue gas mass flow rates. The comparison between experimental and numerical results is presented in Table 2 for L/G of 1.86 kg/kg and 3.77 kg/kg, respectively. We can observe that the experimental data and our numerical results are in general in good agreement for both operating points. However, for the second operating point (L/G=3.77 kg/kg), the calculated Specific Reboiler Duty (SRD), corresponding to the reboiler heat consumption for capturing 1 kg of CO$_2$, is too high compared to the experimental data. This demonstrates that the model overestimates the SRD for higher solvent flow rate.

To overcome this limitation, the counter-flow rich/lean heat exchanger modelling has been improved. Indeed, in the previous model, the temperature difference between the hot inlet and cold outlet streams has been fixed at a constant pinch of 19 °C. However, the solvent flow rate has an influence on the heat exchange performance since it alters the local heat transfer coefficient $h$. In our study, this dependency has to be considered given that the solvent flow rate is expected to vary considerably under CCGT part-load conditions. Therefore, the evolution of the global heat transfer coefficient $U$ as a function of the mass flow $q$ has been modelled in Aspen Plus using the following power law:

$$\frac{U}{U_{ref}} = \left(\frac{m}{m_{ref}}\right)^{m}$$

with $U_{ref}$ the global heat transfer coefficient for a reference mass flow $m_{ref}$. In this case, the reference solvent flow rate corresponds to the first operating point (L/G=1.86 kg/kg). Indeed, the heat exchanger performance is known for this operating point because it has been validated by fixing the cold approach temperature at 19 °C. Therefore, by assuming a global heat transfer coefficient $U_{ref}$ of 2000 W/m$^2$K (typical value for liquid-liquid heat exchanger with aqueous solutions), the required exchanger area $A$ can be calculated. Without knowing the actual plate heat exchanger configuration in the pilot-scale facility, corrugated plates with 30°/30° chevron angle have been envisaged, together with a global counter-flow arrangement. Therefore, the power law exponent $m$ has been determined from the Nusselt correlation for this plate configuration ($m = 0.7424$) [19]. Finally, it can be seen in Table 2 that the model gives a better prediction of the SRD for the second operating point.

**CARBON CAPTURE DESIGN AND OPTIMISATION**

As the CC model has been validated at pilot plant scale, a scale-up of the CC plant is needed to be applied to a typical CCGT power plant. The scale-up approach is described in the following section. Moreover, the operating parameters of the CC process have been optimized to minimize the energy consumption of the CC plant.

**Design of the carbon capture plant**

The main design specifications of the carbon capture are reported in Table 3. As in the pilot plant, IMPT#40 packings are used for the stripper and the absorber columns. The CC plant will be designed to support the highest mass flow rate, i.e when the CCGT operates at full load. The scale-up of the CC plant consists in determining the diameter and the height of both columns.

The column diameter is determined in such a way to avoid flooding in the column. The flooding is a phenomenon during which the upward gas entrains the downward liquid to the top of the column. It occurs when the gas velocity is too high for a given liquid flow rate, leading to an increase in pressure drop [20]. The gas velocity from which flooding occurs is called the flooding velocity. For a given flue gas flow rate $V$, the superficial gas velocity $v_s$, which is the gas velocity calculated as if gas is the...
only fluid flowing in the packing cross sectional area, is directly related to the column diameter $d$ via the following equation:

$$d = \sqrt{\frac{4V}{\pi v_s}} \quad (9)$$

The superficial velocity has to be lower than the flooding velocity. Usually, packed columns are designed to operate at about 60-80% of the flooding velocity [21]. In this work, columns have been designed to operate at 80% of the flooding velocity. The design is based on pressure drop correlation aiming to predict flooding points in a column, which are included in Aspen Plus for several packings [22]. The diameter determination is based on a solvent flow rate corresponding to an L/G ratio of 1.5 kg/kg, knowing that operating at a higher L/G ratio will reduce the safety margin against flooding [20]. Moreover, Aspen Plus determines the column diameter with reference to a specific point in the column, corresponding to the stage where the va-

### TABLE 3: CARBON CAPTURE PLANT DESIGN SPECIFICATIONS.

<table>
<thead>
<tr>
<th>Input</th>
<th>L/G=1.86 kg/kg</th>
<th>L/G=3.77 kg/kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>MEA solvent concentration [wt.%]</td>
<td>30</td>
<td></td>
</tr>
<tr>
<td>$CO_2$ capture level [%]</td>
<td>90</td>
<td></td>
</tr>
<tr>
<td>Absorber 1st stage pressure [bar]</td>
<td>1</td>
<td></td>
</tr>
<tr>
<td>Stripper condenser pressure [bar]</td>
<td>1.2</td>
<td></td>
</tr>
<tr>
<td>Cooling temperature of the flue gas [°C]</td>
<td>40</td>
<td></td>
</tr>
<tr>
<td>Cooling temperature of the lean solvent [°C]</td>
<td>40</td>
<td></td>
</tr>
<tr>
<td>Pumps efficiency</td>
<td>0.8</td>
<td></td>
</tr>
</tbody>
</table>

The superficial velocity has to be lower than the flooding velocity. Usually, packed columns are designed to operate at about 60-80% of the flooding velocity [21]. In this work, columns have been designed to operate at 80% of the flooding velocity. The design is based on pressure drop correlation aiming to predict flooding points in a column, which are included in Aspen Plus for several packings [22]. The diameter determination is based on a solvent flow rate corresponding to an L/G ratio of 1.5 kg/kg, knowing that operating at a higher L/G ratio will reduce the safety margin against flooding [20]. Moreover, Aspen Plus determines the column diameter with reference to a specific point in the column, corresponding to the stage where the va-
por flow is the highest, i.e. the most critical stage. The obtained diameters for the absorber and the stripper columns are respectively 18 m and 8.5 m. Note that one single absorber has been considered in this work. The use of multiple absorbers will be investigated in the future with the aim to increase the CC plant flexibility.

The column height has been determined by performing a sensitivity analysis to evaluate its impact on the CC capture plant energy consumption. The effect of the absorber and stripper heights on the SRD are shown in Fig. 2 for different solvent flow rates. Note that the sensitivity analysis on the absorber height has been performed by keeping the stripper height at 30 m, and the sensitivity analysis on the stripper height has been made by also keeping the absorber height constant at 30 m. Regarding the absorber, we can first observe that decreasing the column height below 15 m leads to a significant increase in SRD for each L/G ratio to keep the CO$_2$ capture rate at 90%. Above 15 m, the SRD decreases linearly with the height. The SRD decrease is more pronounced for low L/G, which means that the higher the solvent flow rate, the lower the influence of the absorber height on the CC performance. Increasing the absorber height from 20 m to 50 m allows to decrease the SRD by 0.2 GJ/tCO$_2$ for an L/G of 0.9 kg/kg, while the SRD reduction is only 0.09 GJ/tCO$_2$ for an L/G of 1.5 kg/kg. As the optimal L/G ratio seems to be between 1.1 and 1.2 kg/kg, and that for these L/G ratios, an increase in absorber height above 30 m does not provide any significant energy consumption reduction anymore, the absorber height has been set at 30 m. Regarding the stripper, the height has a high influence on the SRD for low L/G ratio (0.9 kg/kg), while the impact is very limited for L/G higher than 1.1 kg/kg. Increasing the stripper height from 20 m to 50 m allows to decrease the SRD by 0.27 GJ/tCO$_2$ for L/G=0.9, while no SRD reduction is observed for L/G=1.5 kg/kg. Therefore, similar to the absorber, a stripper height of 30 m has been selected. Note that the choice of the column height has only been performed from an energy consumption point of view. To further optimise the columns size, an economical analysis can be performed [22], but this is outside the scope of this study.

The counterflow heat exchanger has been modelled similarly as in the pilot-plant model, by taking into account the heat transfer dependency to the solvent flow rate. The reference operating point corresponds to the operation at an L/G ratio of 1 kg/kg. By assuming a global heat transfer coefficient of 2000 W m$^{-2}$ K, the required heat transfer area has been determined.

Some design specifications have been implemented in the numerical model to facilitate the model convergence: the CO$_2$ mass flow rate in the lean aqueous solution is adjusted to reach an absorption efficiency of 90% and the reflux ratio of the stripping column is determined to extract the exact CO$_2$ mass flow rate, previously captured in the absorber. Regarding the reboiler, the heat can be provided by an external boiler or by extracting steam from the CCGT steam cycle. However, we opted to not define the thermal energy source. Indeed, since this work aims to evaluate the performance of the CC itself, we wanted to avoid any possible interference of the external boiler or the steam extraction on the CC performance. The heat integration analysis will be part of future works.

**Carbon capture process optimisation**

As explained before, the properties of the flue gas coming from the CCGT plant are obtained from the Thermoflow model and are presented in Fig. 3 and 4 for the different specific loads. We can observe that the mass flow rate, the CO$_2$ content and the temperature of gas are reduced for part-load operations. The flue gas mass flow rate varies from 700 kg/s to 400 kg/s depending on the CCGT operating mode, while the CO$_2$ content varies from 4.6 to 3.8 % and reach a maximal of 4.7 % at 90 % CCGT load.
The flue gas temperature is 88 °C at full load and can decrease to 74 °C at 30 % CCGT load.

The CC process has been optimised to reduce its energy consumption, based on the reboiler duty, for each operating mode, i.e. for each CCGT load. A key parameter of the process is the solvent flow rate. A sensitivity analysis has been performed to find this optimal L/G ratio, i.e. the solvent flow rate which minimizes the specific reboiler duty. The results are presented in Fig. 5 for 3 different CCGT operating loads: 40 %, 70 % and 100 %. We can observe that, for the 3 loads, the L/G ratio has a strong impact on the CC energy consumption. On the one hand, increasing the L/G ratio requires a high energy input to heat up the solvent flow rate. On the other hand, lower L/G ratios than the optimal one lead to a significant increase in the SRD. Indeed, to operate at similar CO₂ removal rate (90 %), decreasing the solvent flow rate leads to an increase in CO₂ cyclic capacity, which is the difference between the CO₂ loading of the lean and rich solvents. However, ones the L/G ratio becomes very low, the lean solvent CO₂ loading drastically decreases, to compensate the reduction in solvent flow rate by increasing the CO₂ cyclic capacity to maintain the same capture efficiency. Therefore, the solvent circulating in the stripper has a lower CO₂ content, which means that an additional energy input is needed to keep the driving force constant [18]. For the CCGT base load operation, the optimal L/G ratio is around 1.15 kg/kg, leading to a SRD of 3.92 GJ/tCO₂. For lower CCGT load, the curves are shifted to the left, showing a lower optimal L/G ratio, and upwards, corresponding to a higher SRD. Operating at 70 % CCGT load results in an optimal L/G ratio of 1.1 kg/kg, corresponding to a SRD of 3.95 GJ/tCO₂, while for 40 % CCGT load, the optimal L/G ratio is 0.98 kg/kg, leading to a SRD of 4 GJ/tCO₂. The obtained SRD values are consistent with those found in literature. Indeed, SRD values from 3.92 GJ/tCO₂ to 4 GJ/tCO₂ are classical for the MEA 30wt.% CC process with low CO₂ partial pressure in flue gas as in the present case.

RESULTS

In this section, the impact of CCGT part-load operating mode on the CC plant performance is evaluated. Moreover, realistic CCGT-CC yearly operation scenarios, involving different load factors, have been analysed and compared.

Carbon capture performance analysis

The following analysis has been performed by considering the CC operation at the optimal L/G ratio for each CCGT load. Fig. 6 shows the overall specific heating and cooling duties of
the CC process according to the CCGT load. The overall specific heating duty corresponds to the specific reboiler duty, since the only heat input to CC process is through the reboiler ((6) in Fig. 1), while the overall specific cooling duty includes the cooling of flue gases to 40°C before entering the absorber column ((1) in Fig. 1), the condenser duty at the top of the stripper column ((7) in Fig. 1) and the cooling of the lean solvent to 40°C before being fed into the absorber (9) in Fig. 1.

Regarding the heating duty (Fig. 6), it can be seen that the SRD increases when the CCGT load decreases. In Fig. 7, we can observe that the CO₂ loading in the lean and rich solvents does not vary significantly with the CCGT load. Therefore, the increase in SRD is not linked to the change in solvent CO₂ loading.

When looking at the temperature of the streams exiting the counter-flow heat exchanger ((4) in Fig. 1), Fig. 8 reveals that the rich solvent temperature is decreasing when the CCGT load decreases. This can be explained by the reduction in the exchanger effectiveness ε when the load decreases as illustrated in Fig. 9. Therefore, since the temperature of the solvent at the inlet of the stripper is reduced, an additional heat input through the reboiler is required to heat up this solvent in the stripper column, explaining the higher SRD at lower CCGT loads.

Regarding cooling duties, the flue gas cooling demand is reduced for part-load operations due to the lower flue gas temperature (Fig. 4) and the condenser duty is lower. However, the solvent cooling requirement increases for partial-load operations due to the higher lean solvent temperature at the outlet of the counter-flow exchanger (Fig. 8). Consequently, although the solvent cooling load is higher, the overall specific cooling duty decreases at partial CCGT load as observed in Fig. 6. These cooling requirements have to be considered in the CC performance evaluation. If cooling water is available in the vicinity of the plant, their contribution will be limited to the water pumps consumption. However, if no cooling water is available, their contribution will be more significant because air-coolers have to be used to evacuate these large amount of energy.

In conclusion, this analysis showed that the CC performance is more affected by the off-design operation of the heat exchanger than the absorber and the stripper columns when the CCGT operates at partial load.

**CCGT annual operation analysis**

In this section, a comparison between the performance of the CCGT and CC plants for different scenarios for the CCGT operation over a year is performed. The objective is to evaluate the required energy input in the plant, as well as the CO₂ emissions, when the CCGT operates with and without carbon capture for different annual load factors. To this end, 4 different production scenarios, presented in Table 4, are analysed. The first scenario is the reference case considering 4000 hours full-load CCGT operation over the year. The second and third scenarios have been determined to, respectively, achieve an equivalent yearly electricity production and the same number of operating hours as the reference case, by considering an annual load factor of 0.5. The last scenario considers an equivalent yearly electricity production as the reference case with a load factor of 0.75.

The resulting plant performances for the 4 scenarios are presented in Table 5. The net electrical production of the plant has been determined by subtracting the electrical consumption of the CC plant from the net electrical CCGT production. The CC plant consumption includes the power consumed by the solvent pumps, the blower which pushes flue gas into the absorber at the inlet of the CC plant and the cooling of the CC plant. This last contribution depends on the availability of cooling water in the vicinity of the plant as discussed in the previous section. In
### TABLE 4: CONSIDERED ANNUAL CCGT PRODUCTION SCENARIOS DESCRIPTION

<table>
<thead>
<tr>
<th>CCGT load [%]</th>
<th>30%</th>
<th>40%</th>
<th>50%</th>
<th>60%</th>
<th>70%</th>
<th>80%</th>
<th>90%</th>
<th>100%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Scenario 1</td>
<td>[h]</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>4000</td>
</tr>
<tr>
<td>Scenario 2</td>
<td>[h]</td>
<td>3000</td>
<td>-</td>
<td>3000</td>
<td>2000</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Scenario 3</td>
<td>[h]</td>
<td>1000</td>
<td>1000</td>
<td>1000</td>
<td>1000</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Scenario 4</td>
<td>[h]</td>
<td>-</td>
<td>600</td>
<td>-</td>
<td>1500</td>
<td>2200</td>
<td>-</td>
<td>1000</td>
</tr>
</tbody>
</table>

### TABLE 5: PERFORMANCE COMPARISON ANNUAL CCGT OPERATION SCENARIOS.

<table>
<thead>
<tr>
<th>Scenario</th>
<th>Annual load factor</th>
<th>Net electrical production [GWhe/y]</th>
<th>Energy input without CC [kWh/kWhe]</th>
<th>Energy input with CC [kWh/kWhe]</th>
<th>Increase in energy input due to the CC [kWh/kWhe]</th>
<th>CO₂ emissions without CC [g/kWhe]</th>
<th>CO₂ emissions with CC [g/kWhe]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Scenario 1</td>
<td>1</td>
<td>2109</td>
<td>1.64</td>
<td>2.12</td>
<td>0.48</td>
<td>326</td>
<td>35</td>
</tr>
<tr>
<td>Scenario 2</td>
<td>0.5</td>
<td>2098</td>
<td>1.86</td>
<td>2.44</td>
<td>0.58</td>
<td>370</td>
<td>40</td>
</tr>
<tr>
<td>Scenario 3</td>
<td>0.5</td>
<td>1050</td>
<td>1.86</td>
<td>2.44</td>
<td>0.58</td>
<td>370</td>
<td>40</td>
</tr>
<tr>
<td>Scenario 4</td>
<td>0.75</td>
<td>2105</td>
<td>1.71</td>
<td>2.22</td>
<td>0.51</td>
<td>341</td>
<td>37</td>
</tr>
</tbody>
</table>

#### FIGURE 8: The temperature of the rich solvent at the outlet of the cross heat exchanger decreases when the load is reduced, while the temperature of the lean solvent increases.

#### FIGURE 9: The effectiveness of the counter-flow heat exchanger drops when the CCGT load is reduced.

In this analysis, to take into consideration the cooling requirements, the cooling duties were converted into electrical power using a Coefficient of Performance (COP) of 5 as done in the works of Li et al. [23]. Regarding the energy input, it includes the gas turbine fuel chemical energy input, plus the reboiler duty if CC is applied. Given that the heat source for the reboiler has not been defined in this work as described previously, the raw heat provided to the reboiler has been considered in this analysis.

We can observe that the required energy input to produce 1 kWh of electricity is higher for lower annual load factor, considering the operation with and without carbon capture (Table 5). Without CC, the required energy input increases by approximately 12% when the load factor varies from 1 to 0.5 due to the increased CO₂ emissions.
to the decline in CCGT efficiency during part-load conditions. Whereas with CC, the energy consumption increased by 15% when the load factor varies from 1 to 0.5 due to the combined effects of CCGT efficiency reduction and rise in CC energy penalty during part-load conditions. Indeed, the energy penalty induced by the CC increases from 0.48 kWh/kWhe to 0.58 kWh/kWhe when the annual load factor is reduced from 1 to 0.5, representing a 21% relative increase.

Concerning CO₂ emissions (Table 5), the reported values for the case without CC are of the same order of magnitude as the monthly values measured by Blondeau et al. [24] on an existing 300-350 MWhe gas-fired combined cycle power plant (335 kg/MWhe for base-load and 350 kg/MWhe for peak-load). The reduction of the annual load factor induces a slight increase in yearly CO₂ emissions. The application of CC allows to decrease CO₂ specific emissions below 40 g/kWhe.

These results highlight the significant negative impact of the increase in CCGT part-load operation periods on the CC plant performance. Moreover, due to the deeper penetration of renewable energies, CCGTs will increasingly run at partial load over the next year, further reducing their annual load factor. Therefore, the plant profitability, as well as the Levelized Cost of Energy (LCOE), will be strongly negatively affected if CC is applied. Nevertheless, the plant profitability is also impacted by the carbon price, increasing drastically since January 2021 (from 30 €/tCO₂ in January 2021 to 80 €/tCO₂ in December 2021 [1]). Therefore, the considerable reduction in CO₂ emissions by applying CC is advantageous from this point of view. Consequently, an accurate economical analysis, which is outside the scope of this work, has to be performed to assess the economical impact of part-load operations on the CCGT-CC plant.

CONCLUSIONS

In this paper, the application of CC to a typical CCGT has been studied under real operating conditions. Indeed, CCGTs are intended to increasingly run at partial load to back-up the growing renewable production, while the CC plant is designed for nominal CCGT load. Therefore, the impact of part-load CCGT operations on CC performance has been investigated through Aspen Plus simulations. The model of the CC plant has been successfully validated against experimental data from the Pilot-scale Advanced Capture Technology (PACT), highlighting the need to take into account of the heat transfer coefficient dependency to the mass flow rate for the counter-flow heat exchanger modelling. Then, the CC plant has been scale-up to support the entire exhaust mass flow rate from full-load CCGT and the process parameters have been optimised to minimize the energy consumption of the plant.

The simulation results showed that the specific reboiler duty increases, while the specific cooling duties decrease, when the CCGT load is reduced. The higher heating requirement is caused by the lower heat exchanger effectiveness in off-design operation mode, while the absorption and stripping processes do not seem to be affected by part-load operation. Additionally, the analysis of the CCGT operation with CC over a year showed that the energy penalty induced by the CC plant is increased by 21% when an annual CCGT load factor of 0.5 is considered, compared to the full-time CCGT operation scenario at nominal load, impacting considerably the plant profitability.

PERSPECTIVES

The next step will be dedicated to the reduction of the CC energy penalty. To this end, the CC process will be improved by investigating the use of more advanced solvents and process configurations. Moreover, the application of EGR, which is a way to increase the CO₂ content so as to reduce the energy penalty, will be studied. Additionally, the heat integration between the CCGT and the CC plant will be performed to determine the most optimal way to provide the thermal energy to the reboiler.

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REFERENCES


