Absorption-based carbon capture energy penalty reduction for micro gas turbine applications: investigation of advanced carbon capture configurations[#]

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ABSTRACT

Applying post-combustion Carbon Capture (CC) offers a solution to reduce CO_2 emissions of micro Gas Turbines (mGTs). However, the conventional monoethanolamine-based (MEA) absorption-regeneration process was never found economically feasible for these small scales (5-500 kW_e) due to the high energy consumption of this CC process, degrading the plant performance. Improving the CC configuration can help to reduce the CC energy penalty. To this end, two advanced configurations are investigated, the Rich Solvent Recycle (RSR) and the Rich Solvent Split (RSS) configurations, and applied to a typical mGT, namely the Turbec T100, through thermodynamic cycle modeling in Aspen Plus. The optimal operating parameters were first determined for each configuration to minimize the energy consumption. The results showed that by applying RSR and RSS, the CC reboiler duty could be reduced by 2% and 0.9%, respectively, compared to the conventional one. Moreover, the increase in mGT electrical efficiency is limited to 0.04% for both processes, demonstrating that such configurations do not provide significant benefits for mGT applications.

Keywords: Micro Gas Turbine, Absorption based carbon capture, Rich Solvent Recycle, Rich Solvent Split, Thermodynamic cycle simulations

NOMENCLATURE

Abbreviations	
A	Heat exchanger area (m^2)
CAPEX	Capital Expenditure
CC	Carbon Capture
CHP	Combined Heat and Power
EGR	Exhaust Gas Recirculation
L/G	Liquid to gas ratio (kg/kg)
m	Heat transfer power-law exponent
MEA	Monoethanolamine
mGT	micro Gas Turbine

RSR	Rich Solvent Recycle
RSS	Rich Solvent Split
SRD	Specific Reboiler Duty (MJ/kg_{CO_2})
U	Global heat transfer coefficient (W/m^2K)
Symbols	
η	Electrical efficiency
$\dot{m}$	Mass flow rate (kg/s)

1. INTRODUCTION

In the context of the massive deployment of renewable energies, small-scale power production in decentralized energy systems, possibly in Combined Heat and Power (CHP) applications like micro Gas Turbines (mGTs), offers a high potential to provide the necessary flexibility to the electricity grid. Nevertheless, although burning natural gas results in lower CO_2 emissions than other fossil fuels like coal or oil, reducing the carbon footprint of mGTs is essential to move towards a carbon-clean power production. In this perspective, applying post-combustion Carbon Capture (CC) offers a solution to remove CO_2 from mGT flue gases.

The amine(s)-based absorption-regeneration process is widely used and commercially mature for removing CO_2 from flue gases [1]. Monoethanolamine (MEA) is the benchmark solvent given its high reactivity with CO_2 and its low price [2]. However, the MEA-based process is characterized by a high energy requirement for absorbent regeneration due to the high decomposition enthalpy of carbamate formed by MEA and CO_2 (95 kJ/molCO_2) [3]. Therefore, applying CC significantly reduces the plant efficiency. Moreover, the energy penalty is high for gas turbines, and more specifically for mGTs, due to the low CO_2 concentration in their exhaust gases (1.5 vol.%). Indeed, mGTs operate with a high excess of air to limit the turbine inlet temperature for blade resistance constraints [4]. Applying Exhaust Gas Recirculation (EGR) is a way to increase the CO_2 content

in flue gases (up to 4.5 vol.%) to decrease this energy penalty [5]. Giorgetti et al. [6] investigated the energy impact of coupling an mGT and a CC in series. The results showed that the CC reboiler duty is 4.43 MJ/kgCO$_2$, which reduces the mGT electrical efficiency by 7.9 percentage points. Mansouri et al. [7] showed a decrease in electrical efficiency from 31.2% to 23% by adding CC to an mGT. Therefore, this reboiler duty must be reduced to make carbon-clean mGT more efficient and thus economically profitable.

Next to the selection of a more appropriate solvent [8], the reduction of the CC energy consumption can be achieved by using more advanced process configurations. Le Moullec et al. [9] provide an in-depth review of possible improvements for the amine(s)-based carbon capture process. Some of these improvements, such as InterCooled Absorber (ICA), Rich Solvent Recycle (RSS), or Rich Solvent Split (RSS), require only minor modifications of the process, enabling potential performance improvement without significantly increasing the CAPEX of the plant. Several studies demonstrated the positive effect of using advanced configurations for removing CO$_2$ from large-scale coal-fired [2,3], gas-fired [10–12], and cement plants [13] flue gases. However, the impact of using advanced configurations for removing CO$_2$ from mGT exhaust gas (with a lower CO$_2$ content) has not been investigated.

Due to the lack of studies dealing with the CC process improvement for mGT application, a preliminary investigation is needed to determine whether advanced configurations are efficient for treating flue gases with low CO$_2$ content. To this end, this work focuses on the implementation of two advanced CC process configurations applied to a typical mGT, the Turbec T100, through thermodynamic simulation with the Aspen Plus software v9. The considered configurations are the Rich Solvent Recycle (RSS), and the Rich Solvent Split (RSS).

The paper is structured as follows: first, the simulation approach of the CC plant is presented. Next, both process modifications are described. The related operating parameters to each process are then optimized to minimize the energy consumption followed by the comparison of CC performance.

2. SIMULATION APPROACH

A conventional 30 wt.% MEA absorption-based CC process, was considered, including two columns: absorber and stripper (Fig. 1). The CC plant has been modeled upon the Pilot-scale Advanced Capture Technology (PACT) facilities at the UK Carbon Capture and Storage Research Centre (UKCCSRC). The model has been developed in Aspen Plus v9.
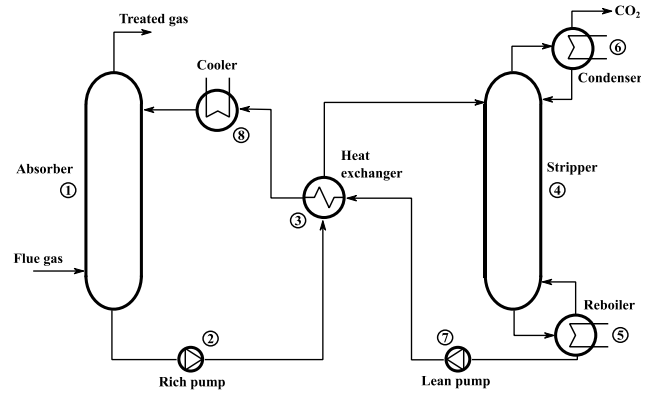

Fig. 1 The CC plant is a conventional amine(s)-based absorption regeneration process including absorber and stripper columns and heat exchangers.

The thermodynamic models used are the Electrolyte Non-Random Two Liquid (ENRTL) model for electrolyte properties in liquid phase and the PC-SAFT equation of state for vapor phase. The absorber and the stripper columns were simulated using the Aspen Plus RadFrac model and the rate-based approach was used. The Hanley IMPT model has been applied for the mass transfer and the interfacial area calculations. The absorption and stripping columns have been divided into 30 calculation stages in the simulation.

Regarding the internal lean/rich heat exchanger modelling, the exchanger area A has been set and the global heat transfer coefficient U is calculated using the following power law (Eq. 1) to consider the impact of the solvent flow rate on the heat exchange performance:

$$\frac{U}{U_{ref}} = \left(\frac{\dot{m}}{\dot{m}_{ref}}\right)^m \qquad (1)$$

with U_{ref} the global heat transfer coefficient for a reference mass flow $\dot{m}_{ref}$, corresponding to a validated operating point. Moreover, corrugated plates with a 30°/30° chevron angle, together with a global counter-flow arrangement, have been assumed. The power-law exponent m has been determined based on the correlation of Khan et al. [14] for this plate configuration (m=0.7424).

The numerical model has been validated against experimental data from Agbonghae et al. [15] for two specific operating points, defined by their Liquid to Gas (L/G), which is the ratio between solvent and flue gas mass flow rates. The comparison between experimental and numerical results is provided in Table 1 for L/G ratios of 1.86 kg/kg and 3.77 kg/kg, respectively. We can observe that the experimental data and our numerical results are in good agreement for both operating points.

In this work, the capture efficiency has been set to 90% for all the simulations, as is commonly done in similar cases in the literature. To facilitate convergence

Table 1   CC model validation for 2 specific operating points

	L/G = 1.86 kg/kg		L/G = 3.77 kg/kg	
Inputs	Experiments [16]	Numerical model	Experiments [16]	Numerical model
Flue gas mass flow rate [Nm³/h]	207.3 ± 1.8	207.3	192.1 ± 1.6	192.1
Flue gas temperature [°C]	41.3 ± 0.5	41.3	39.4 ± 0.3	39.4
Flue gas pressure [barg]	0.17 ± 0.02	0.17	0.19 ± 0.01	0.19
CO_2 content in flue gas [vol.%]	4.48 ± 0.11	4.48	4.55 ± 0.11	4.55
CO_2 mass flow in flue gas [kg/h]	18.23 ± 0.16	18.23	17.17 ± 0.14	171.17
Solvent mass flow [kg/h]	515.6 ± 5.4	515.6	964.3 ± 8.5	964.3
MEA concentration [wt.%]	28.2 ± 0.1	28.2	25.6 ± 0.1	25.6
Lean solvent temperature [°C]	39.9 ± 0.9	39.9	40 ± 0.5	40
Lean CO_2 loading [molCO_2/molMEA]	0.246 ± 0.001	0.246	0.153 ± 0.001	0.153
Stripper condenser pressure [barg]	0.2 ± 0.02	0.2	0.2 ± 0.02	0.2
Results				
Rich CO_2 loading [molCO_2/molMEA]	0.409 ± 0.001	0.411 (0.56%)*	0.247 ± 0.001	0.246 (0.4%)*
CO_2 captured [kg/h][a]	16.47 ± 0.4	16.48 (0.07%)*	16.3 ± 0.59	16.09 (1.31%)*
CO_2 capture efficiency [%]	90.35 ± 3	90.38 (0.04%)*	94.93 ± 4.2	93.73 (1.27%)*
Specific reboiler duty [MJ/kg$_{CO2}$]	5.92 ± 0.8	5.81 (1.77%)*	13.27 ± 2.21	15.67 (18.09%)*

* Relative difference between experimental and numerical results
[a] CO_2 capture efficiency calculated on a mass basis

and reach this capture efficiency, several design specifications have been implemented: the CO_2 flow rate in the lean aqueous solution is adjusted to reach an absorption efficiency of 90% and the reboiler duty is adapted to release the targeted amount of captured CO_2.

Note that the thermal energy source for the reboiler has not been defined in this work, since we aim to focus on the impact of CC configurations on the CC process performances themselves. This was done to better understand these impacts, and to avoid any possible interference of the heat source on the performance.

As the input of the CC model, mGT flue gas properties were determined by modeling the advanced mGT+EGR cycle. More details about modeling can be found in [5]. The requested electrical power output was set to the nominal power of 100 kW$_e$. The EGR ratio was adapted to obtain a concentration of oxygen in the combustor inlet of 16 vol.%. This fraction corresponds to the minimal O_2 amount ensuring stable combustion and limiting CO emissions. The obtained mGT flue gas mass flow rate is 0.291 kg/s with a CO_2 content of 4.31 vol.%. The absorber and stripper dimensions were adapted to treat the entire mGT flue gas flow rate [6]. The columns specifications are presented in Table 2.

3. PROCESS MODIFICATIONS

In this section, both considered advanced configurations are described.

3.1 Rich Solvent Recycle

The Rich Solvent Recycle (RSR) configuration, shown in Fig. 2, consists in recycling a fraction of the rich solvent coming from the bottom of the absorber to send it back to an upper level of the column. This modification aims to increase the rich solvent CO_2 loading. To maximize the absorption capacity, and thus reduce the energy consumption of the CC plant.

3.2 Rich Solvent Split

In the Rich Solvent Split (RSS) configuration, shown in Fig. 3, a fraction of the rich solvent coming from the absorber by-pass the internal lean/rich heat exchanger

Table 2   Columns specifications.

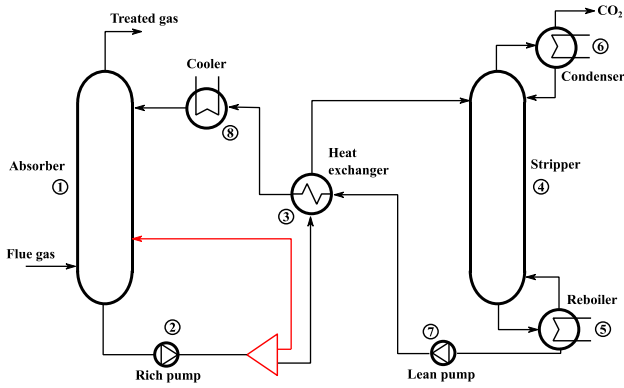|	
Absorber height	6 m
Absorber diameter	0.6 m
Stripper height	6 m
Stripper diameter	0.4 m
Packing type of the column	IMTP
Packing size	38 mm
Absorber pressure	1.01325 bar
Stripper pressure	1.2 bar

Fig. 2 The Rich Solvent Recycle process consists in recycling a fraction of the rich solvent exiting the absorber into an upper level of the column.

and is directly sent to the top of the stripper column. This option aims to improve the heat integration in the process by increasing the heat recovered from the hot lean solvent, as well as reducing the condenser cooling energy.

4. RESULTS AND DISCUSSION

To perform a relevant comparison between the different CC configurations, the key operating parameters related to each process have been optimized to minimize the Specific Reboiler Duty (SRD), representing the reboiler heat consumption for capturing 1 kg of CO_2.

4.1 Conventional configuration

The solvent flow rate, or the L/G ratio, is the main operating parameter for the conventional absorption-regeneration process. Therefore, its influence on the SRD is shown in Fig. 4. The minimum energy consumption is identified for an L/G ratio of 1.2 kg/kg, leading to an SRD of 4.53 MJ/kg_{CO2}. The optimization of the L/G ratio will be performed for all advanced processes.
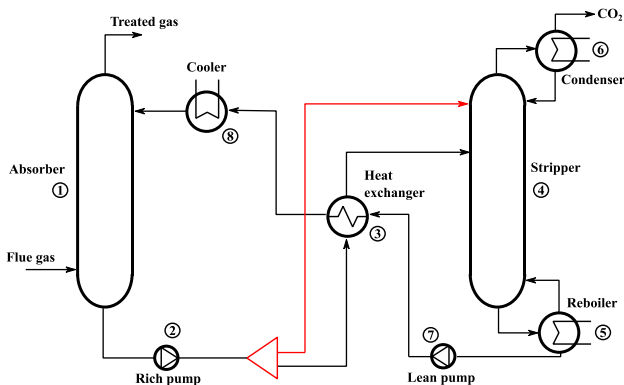

Fig. 3 The Rich Solvent Split configuration consists in sending a fraction of the rich solvent to the top of the stripper without being preheated in the internal lean/rich heat exchanger.
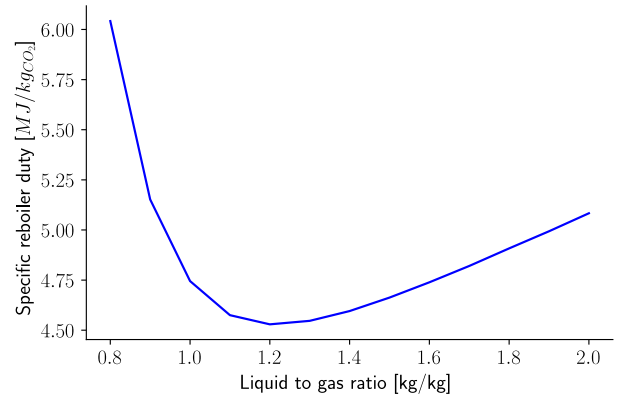

Fig. 4 The solvent flow rate has a strong influence on the reboiler duty for the conventional process.

4.2 Rich Solvent Recycle

The parameters to be optimized in the RSR configuration are the L/G ratio, the fraction of the rich solvent which is sent back to the absorption column, and the injection level of the recycled solvent into the absorber column.

In a first step, different injection levels of the recycled solvent were considered between stage 1 (bottom of the absorber) and stage 30 (top of the absorber). For each considered injection level, the split fraction has been set to 50% and the L/G has been varied as shown in Fig. 5. We can observe that the SRD decreases when the recycled solvent is injected at the bottom of the absorber (between stages 2 and 5). By considering the injection level at stage 2, the L/G ratio minimizing the energy consumption is 1.2 kg/kg.

Then, the fraction of the recycled solvent has been optimized by considering the injection level at stage 2 and an L/G ratio of 1.2 kg/kg (Fig. 6). The split fraction has been varied from 0 to 0.8. It can be observed that as the fraction of solvent sent back to the absorber
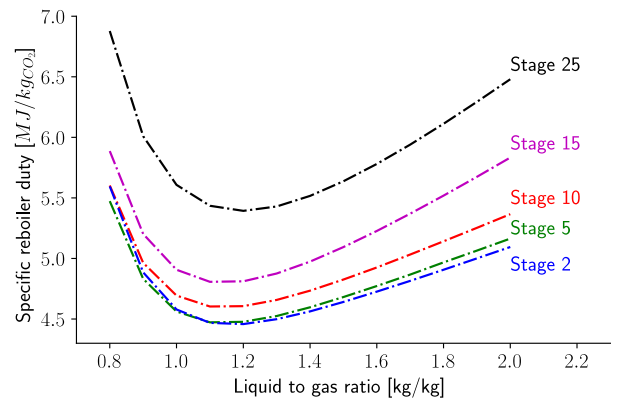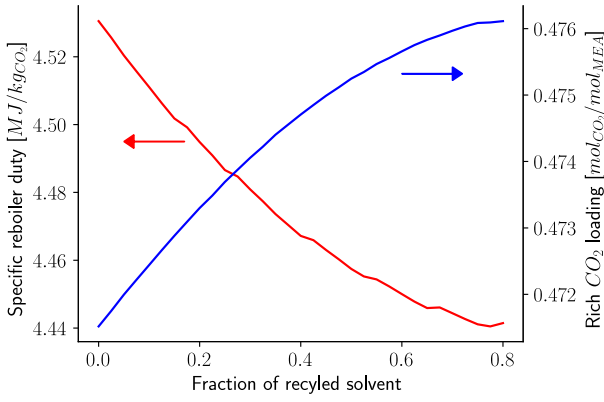

Fig. 5 For a fixed splitting ratio of 50%, injecting the recycled solvent at a low level in the absorber column is more profitable from an energy consumption point of view.

Fig. 6 Increasing the fraction of recycled solvent allows increases the rich solvent CO_2 loading and thus decreases the specific reboiler duty (re-injection at stage 2 and L/G of 1.2 kg/kg).

increases, the rich solvent CO_2 loading increases and, thus, the reboiler duty decreases. Therefore, recycling 80% of the rich solution allows minimizing the energy consumption, leading to an SRD of 4.44 MJ/kg$_{CO2}$. The value of the maximum split fraction depends on the capacity of the absorption column.

4.3 Rich Solvent Split

The parameters to be optimized for the RSS configuration are the L/G ratio, the fraction of the rich solvent which is directly sent to the stripper without being preheated (cold solvent), the injection level of the cold solvent in the stripper, and the injection level of the preheated solution (hot solvent).

Firstly, the effect of the split fraction has been analyzed by considering the injection of the cold solvent at stage 30 (top of the column) and the hot solvent at stage 25. In Fig. 7, we can observe that the optimal split fraction is between 25% and 35% for the different L/G
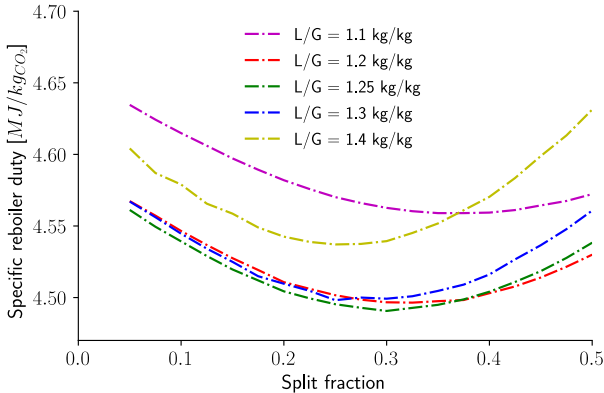
Fig. 7 The optimal splitting fraction is between 25 and 35% for the different L/G ratios (injection of the cold solvent at stage 30 and the hot solvent at stage 25).
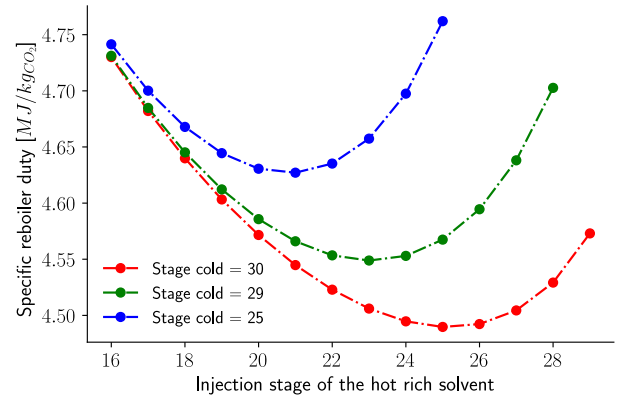
Fig. 8 Injecting the cold solution at the top and the hot solution at stage 25 of the stripper allows to minimize the energy consumption (splitting fraction of 30% and L/G of 1.25 kg/kg).

ratios. The energy consumption is minimized for an L/G ratio of 1.25 kg/kg and a splitting fraction of 30%, resulting in an SRD of 4.49 kg/kg.

Next, the influence of the injection level of the cold and hot solvents has been studied by considering a split fraction of 30% and an L/G ratio of 1.25 kg/kg (Fig. 8). Injecting the cold solvent at stage 30 provides better performance. Moreover, the condenser duty is reduced by 98% when the cold solution is directly injected at the top (condenser). Regarding the hot solvent, the injection at stage 25 provides the best results, allowing to reduce the SRD to 4.49 MJ/kg$_{CO2}$.

4.4 Comparison of process configurations

To assess the benefit of modifying the conventional CC process for mGT application, a comparison of performance between the conventional, the RSR, and the RSS processes has been performed (Table 3). For each process, the optimized operating parameters, determined previously, have been considered (Table 4). The electrical efficiency of the mGT has been calculated by considering the reboiler duty and the pump's electrical consumption of the CC plant. Firstly, we can observe that the RSR configuration results in a lower SRD (4.44 MJ/kg$_{CO2}$), corresponding to a reduction of 2%

Table 3   Performance comparison.

	Conventional	RSR	RSS
SRD (MJ/kg$_{CO2}$)	4.53	4.44	4.49
Rich pump (W)	42	233	44
Lean pump (W)	35	34	36
η$_{mGT+CC}$	22.13	22.17	22.17

Table 4 Optimized parameters for each configuration.

	Conventional	RSR	RSS
L/G (kg/kg)	1.2	1.2	1.25
Split fraction	-	0.8	0.3
Re-injection absorber stage	-	2	-
Hot solution stripper stage	-	-	25
Cold solution stripper stage	-	-	1

compared to the conventional configuration, while the RSS configuration leads to an SRD of 4.49 MJ/kg$_{CO2}$ representing an SRD reduction of 0.9%. Regarding the mGT electrical efficiency, it can be observed that the gain is very limited for both advanced configurations compared to the conventional one (+0.04%). Indeed, despite the lower SRD for the RSR configuration, the rich pump electricity consumption is higher due to the higher solvent flow rate circulating within it, negatively impacting the efficiency. Therefore, although these advanced configurations offer real gains for applications with high CO_2 content in the flue gas (cement plants, coal-fired plants), they do not provide significant improvement of CC performance for an mGT application, characterized by a low CO2 content in the flue gas.

5. CONCLUSIONS

The objective of this work was to investigate the potential reduction in CC energy consumption by slightly modifying the conventional process for mGT application, characterized by a low CO_2 content in flue gases. To this end, two advanced CC configurations, the Rich Solvent Recycle (RSR) and the Rich Solvent Split (RSS), have been considered and modeled using the software Aspen Plus.

The different operating parameters, related to each process, have been first optimized to minimize the specific reboiler duty of the CC plant. Then, the results showed that the RSR and RSS configurations allow to reduce by 2% and 0.9%, respectively, the SRD compared to the conventional process. However, the increase in mGT electrical efficiency is only 0.04% for both processes demonstrating that such configurations do not provide significant benefit for mGT application, even more, the CAPEX of the plant will be slightly higher.

In future works, we will investigate other potential advanced configurations and proceed to the energy integration between the mGT and the conventional CC plant to further decrease the energy penalty.

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