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International Journal of Greenhouse Gas Control

Investigation into simulating Selective Exhaust Gas Recirculation and varying pressurized hot water temperature on the performance of a Pilot-scale CO₂ Capture Plant with 40 wt(%) MEA

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ABSTRACT

The concentration of an aqueous solution of amine affects the solvent regeneration energy requirement in the Post-combustion CO₂ Capture (PCC) process. Therefore, this study investigates the performance of the impact of Selective Exhaust Gas Recirculation (S-EGR) under the influence of 40 wt(%) MEA i.e. 10 % above the benchmark Monoethanolamine (MEA) concentration and at 90% CO₂ capture efficiency due to its potential to reduce the Normalized Specific Reboiler Duty (N-SRD) of the CO₂ capture process. The experimental research work was carried out at the United Kingdom Carbon Capture Storage Research Centre - Pilot-scale Advanced CO₂ Capture Technology (UKCCSRC-PACT) National Core Facility, UK. The S-EGR was proposed as a means of CO₂ enhancement at the inlet of the absorber column to expedite the driving force of CO₂ absorption and consequently reduce the N-SRD. CO₂ concentrations from 5.0 to 9.9 vol(%) of CO₂ were studied. Also, this experiment studied the influence of the Pressurized Hot Water (PHW) inlet temperature at the reboiler on the performance of the CO₂ capture process that includes the CO₂ recovery rate, CO₂ loadings and N-SRD of the Solvent-based CO₂ Capture Plant (SCCP) in order to study the variation of N-SRD with varying reboiler thermal inlet temperature at 9.0 vol(%) CO₂.

It is found that a pilot-scale CO₂ capture process under the influence of simulated S-EGR reduces the N-SRD by 25.1 % under the test condition at 6.6 vol(%) CO₂. This test condition was observed to have lower N-SRD as with regards to the other tests with different CO₂ concentrations, below and above which the N-SRD begins to lose its value. It was also established that within the boundary of the process conditions used in these tests, the impact of the PHW temperature on the CO₂ capture efficiency increases with increasing the PHW temperature, but at the detriment of N-SRD, which begins to increase above 125 °C despite more CO₂ being captured.

KEYWORDS

Post-combustion CO₂ Capture (PCC)

Normalized - Specific Reboiler Duty (N-SRD)

Pressurized Hot Water Temperature (PHW)

Selective - Exhaust Gas Recirculation (S-EGR)

Monoethanolamine (MEA)

1. INTRODUCTION

In spite of global efforts to stabilize the global warming increase to 2 °C or even lower to an ambitious 1.5 °C by 2100 (UNCC; 2019a, UNCC; 2019b) fossil fuels are expected to continue to play a dominant role until the 2040's. Under the International Energy Agency (IEA) 2 °C Scenario in 2060, CCS is expected to contribute to at least 7 % of the decarbonisation endeavours of the fossil fuel electricity generation and heavy industry plants by 2060 (EIA; 2017). Natural gas is expected to have a much longer lifespan due to its lower carbon emission per kg of fuel burnt as compared to coal and oil, its substantial geological reserves, and lowering prices due to the projected increase in supply (EIA; 2019, EIA; 2017, The Economist; 2016,). As a result, Carbon Capture and Storage (CCS) if integrated with the Natural Gas Combined Cycle (NGCC) power plant, thus makes a rational means to continue utilizing natural gas for electricity production while attaining significant CO₂ capture efficiency. Furthermore, attaining potentially negative emission when biomass is used in part or completely as the source of fuel (Bui et al; 2018).

Having acknowledged that natural gas and CCS are going to play an important role in the global decarbonisation economy, PCC is likely to be the first technology to be retrofitted to existing power plants due to its ease of retrofit compared to pre-combustion and oxy-fuel combustion CO₂ capture (Siemens, 2019). However, PCC attracts a high solvent regeneration energy requirement due to the low CO₂ concentration in the exhaust gas of a gas turbine of about 3 - 4 vol(%) (Merkel et al; 2013). The ensuing low CO₂ partial pressure as a result necessitates that more steam (thermal energy), which will otherwise be utilized to generate electricity at the low pressure steam turbine in the Natural Gas Combined Cycle (NGCC), will be employed for solvent regeneration (DOE; 2019, Li et al; 2016). In excess of 70 % of the entire energy requirement for the CO₂ capture plant is consumed for regeneration of the amine solvent and this corresponds to approximately 3.2 – 4.2 MJ/kg of CO₂ captured for the flue gases of gas power plants using 30 wt(%) MEA at 90 % CO₂ capture efficiency (Rezazadeh et al; 2016, Mantripragada et al 2019, Raynal et al; 2011, Brigman et al; 2014, Leung et al; 2014). The S-EGR in series process conveys all of the CO₂ depleted flue-gas exiting the absorber to the selective membrane, where more of this CO₂ is transferred via the membrane to the compressor inlet, thus the resulting exhaust gas from the gas turbine will be richer in CO₂. This configuration has the advantage of lowering the required stripper CO₂ recovery rate and thereby further decreasing the solvent regeneration energy requirement. Whereas, the S-EGR in parallel

reduces the volumetric flow entering the SCCP and diverts part of it to the CO₂-selective membrane, consequently reducing the solvent flow regime and management of the CO₂ capture process. The CO₂ in the diverted fraction of the volumetric flow of the flue gas is captured by the CO₂-selective membrane and re-routed back to the compressor inlet. This configuration has the benefit of reduced CapEx and OpEx of the SCCP due to the reduced flue-gas flow rate.

A number of studies have been carried out on S-EGR to evaluate its practicability. Merkel et al, 2012 appraised the energy consumption of S-EGR and reported a 42 % decrease in the energy of CO₂ separation as the CO₂ concentration in the flue gas was increased from 4 to 19 %. Diego et al., 2017 reported based on process modelling results that 49.3 % net efficiency of NGCC with S-EGR in parallel at 53 % S-EGR ratio was achieved as compared to the 49.0 % and 49.8 % values obtained for the NGCC + CCS and EGR systems at 39 % EGR ratio, respectively. Laura et al., 2018 analysed the cost benefits of S-EGR in parallel and established that at 70 % recirculation, 97 % of selective CO₂ efficiency and 96 % of the PCC efficiency will translate to 46 % reduction of packing volume and 5 % reduction in SRD. However, it was Akram et al., 2016 who evaluated the performance of the CO₂ capture process on a pilot-scale based on varying CO₂ concentrations and reported a 7.1 % decrease in the SRD per unit percentage increase in CO₂ concentration. Having established by Abu Zahra et al., 2006, that higher amine concentration lead to reduction in the SRD, this study investigates the influence of S-EGR on the performance of a pilot-scale capture plant against 40 wt(%) MEA. S-EGR in series was mimicked in these tests by enriching the CO₂ concentration at the absorber gas inlet with the aim of increasing the CO₂ partial pressure, consequently increasing CO₂ mass transfer from the gaseous to the liquid phase. This technique is expected to facilitate the commercial deployment of a gas turbine-CCS on a global scale and to subsequently attain its full potential in the global decarbonisation disposition via reducing the solvent regeneration energy requirement and enhancing CO₂ capture, this will subsequently promote reduction in both the operational and overnight expenditure of the CO₂ capture facility. (Merkel et al; 2013, Herraiz et al; 2018, Diego et al; 2017, Huang et al; 2008, Nanomemc2; 2019, Energy-2050; 2019).

2 MATERIALS AND METHODS

2.1 MATERIALS

A 40 wt(%) aqueous solution of MEA was employed as the solvent for CO₂ absorption as against the benchmark 30 wt(%) MEA in order to improve CO₂ capture and reduce the N-SRD

of SCCP (Brigman et al; 2014, Abu-Zahra et al 2007). The synthesized flue gas, i.e. CO₂ and air were sourced from the onsite cryogenic CO₂ storage via the Gas Mixing Skid (GMS) and the ambient air respectively. The PHW employed for solvent regeneration was produced from an onsite electrical boiler. Furthermore, demineralized water was used for the flue-gas wash. (PACT; 2018, UKCCSRC 2013).

A pocket colorimeter II test kit was utilized for the determination of the Iron (Fe) content via the 1,10-phenanthroline method with a sensitivity to measure the Fe content as low as 0.01 mg/L Fe (HACH; 2019a, HACH; 2019, Pyenson et al; 1945, Marczenko et al; 2000).

2.2 METHODS

The UKCCSRC-PACT CO₂ Capture Plant was employed for this study and is schematically illustrated in Fig. 2.1. The PACT Carbon Capture is designed to capture in excess of 1 tonne of CO₂ per day using 30 wt(%) MEA at 90 % CO₂ capture efficiency using a coal-fired flue gas (PACT; 2018). The SCCP can be integrated to the on-site combustion rig to enable a PCC investigation from these facilities that includes a Turbec micro-Gas Turbine (mGT) (PACT; 2018). Moreover, the solvent-based CO₂ capture plant can be integrated to a GMS to enable the study of the carbon capture process of the synthesised flue gas and gain insight into how different flue gas composition affect the CO₂ capture and the SRD.

The PACT Amine CO₂ Capture Plant includes an absorber and stripper column reactors, which are 8.0 and 7.5 m in length respectively. The absorber is fitted with 6.5 m and 0.3 m of Sulzer Mellapak CC3 structured packing in length and diameter respectively. The absorber reactor is characterized by ten Resistance Temperature Detectors (RTD) along its length at intervals of approximately 0.68 m and at about 0.56, 1.25, 1.93, 2.62, 3.30, 3.99, 4.67, 5.36, 6.04 and 6.73 m from the bottom. The Stripper which is equipped with 6 m and 0.3 m IMTP 25 packing type has nine RTDs at roughly 0.9 m intervals at about 0.39, 1.34, 2.19, 3.04, 3.89, 4.84, 5.69, 7.09 and 7.59 m from the bottom (Akram et al; 2016, PACT; 2018, Tait et al; 2018). The results obtained at the temperature points are plotted in order to gain a better understanding of the thermal behaviour within the columns during the course of CO₂ absorption and desorption under the influence of varying CO₂ concentrations and PHW temperatures. Centrifugal pumps and Pneumatic Diaphragm valves were installed to condition the flow of the solvent within the SCCP and electromagnetic flow meters measure the flows as shown in Fig. 2.1. Furthermore, manual solvent sampling faucet are also installed at the rich solvent, lean solvent and water-

wash streams to enable sampling of the solvent streams for analysis. The solvent is analyzed using a Mettler Toledo Automatic Titrator.

The solvent concentration was determined after the rich and lean solvent samples were extracted from the plant via dedicated sampling valves at the end of a stabilized state using the following relationship (Akram et al; 2016):

$$\text{Solvent (\%)} = \frac{\text{mL of NaOH} \times \text{NaOH molarity} \times 0.044 \times 100}{\text{Sample weight}} \quad \text{Eq. (1)}$$

whereas the CO₂ loading is determined as follows (Akram et al; 2016):

$$\text{CO}_2 \text{ loading} = \frac{\text{wt\% of CO}_2 \times 1.39}{\text{wt\% of MEA}} \quad \text{Eq. (2)}$$

Fig. 1 shows the process flow of the UKCCSRC PACT CO₂ capture plant.

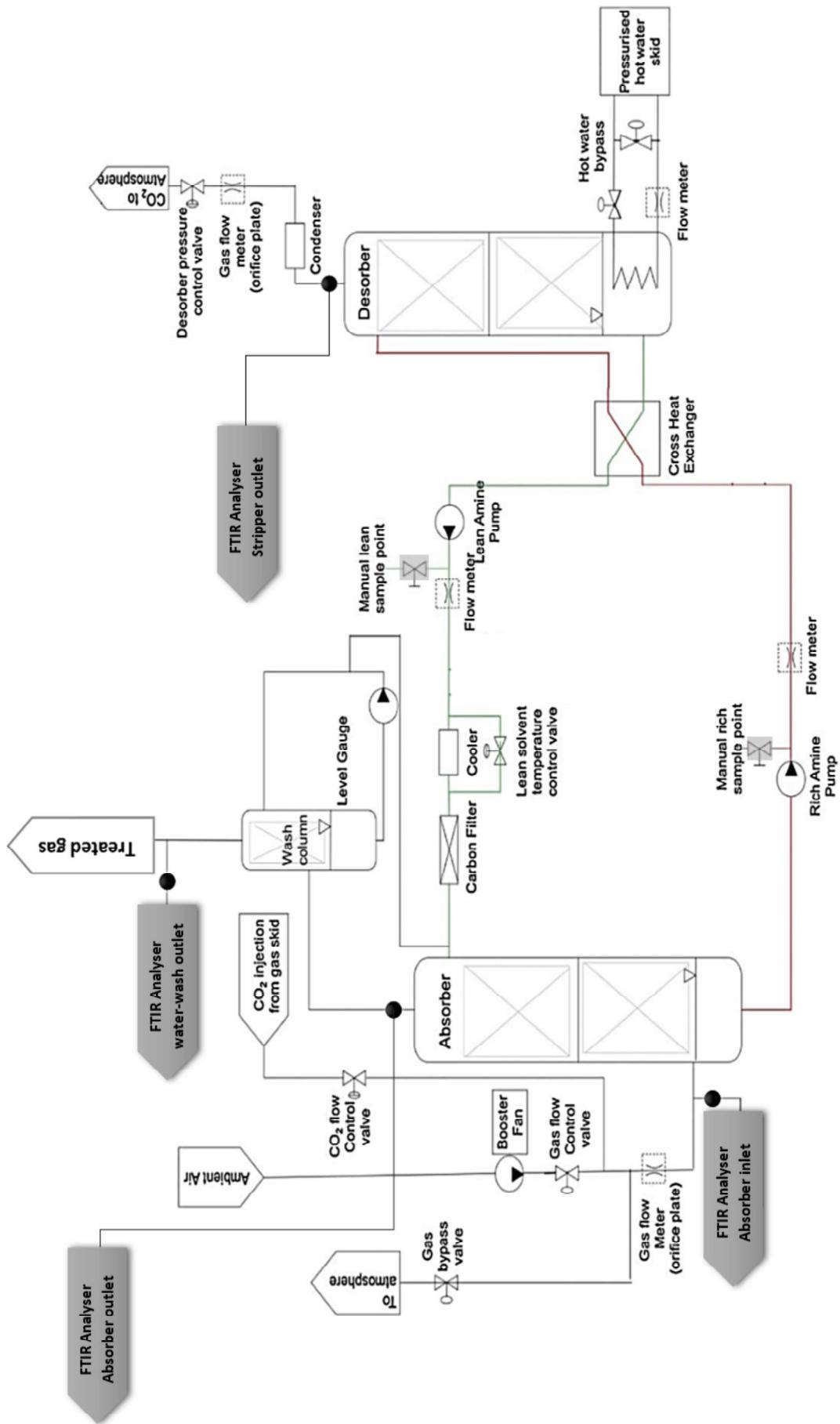


Fig. 1 A schematic of the PACT CO₂ capture plant, indicating the gas sampling line outlets in black spherical shapes (Akram et al., 2016, Tait et al.; 2016).

A booster fan is utilized to introduce the synthesized flue gas into the absorber gas inlet and the mass flow of the CO₂ from the skid is modified via a control valve until a desired concentration of CO₂ in the flue gas is achieved at the absorber gas inlet. The flow of the solvent which is conveyed in a counter-current direction to the gas, is activated by turning on the centrifugal pumps and set to a desired flow at the Human Machine Interface (HMI). The pneumatic diaphragm valves positioning is adjusted until the desired flow pressure is achieved. The PHW is supplied to the reboiler at a given flow rate, the temperature of the PHW at the inlet of the reboiler is set at the electric boiler unit and the flow into the reboiler is controlled via a three-way valve.

Four gaseous sampling points, that include the absorber inlet, absorber outlet/water-wash inlet, stripper outlet and water-wash outlet, were utilized to sample the gases by the gas probe and transported via insulated hot gas sampling lines to the heated sampling oven which is kept at a temperature of about 180 °C to avoid condensation. The sampled gas is thereafter pumped to the Gasmet DX 4000 (Fourier-Transform Infrared Spectroscopy) FTIR analyser at 40-50 L/min and analyses the sampled gas at about every 7 seconds in the course of the experiments. The FTIR identifies the composition of the gas sample via establishing a unique infrared spectral profile based on the individual gas compound absorption capacity to the infrared radiation (Gasmet; 2019, ThermoFisher; 2019). The desired sampling line is connected to the FTIR to obtain readings when the process conditions are observed to be stable. The composition and concentration of the gases on every line and process fluid temperature, pressure and flow rate are collected from the FTIR, Programmable Logic Controller (PLC) and Labview software logs. The result of the composition and flow rate regimes of the fluid across the CO₂ capture plant are utilized to evaluate for key dependable parameters. This include the CO₂ recovery rate, the N-SRD and emission studies; whereas the temperature result on the columns give an understanding of the behaviour of the thermal response of the CO₂ capture and solvent regeneration under different operating conditions.

The mass flow rate of the captured CO₂ (mCO₂) was calculated by ascertaining the difference between the flow rate at Standard Temperature and Pressure (STP) of the CO₂ at the absorber inlet and CO₂ in the treated gas (Akram et al, 2016) from the FTIR data. Whereas, the CO₂ capture efficiency (removal rate) was determined by the percentage ratio of the CO₂ captured to the CO₂ at the absorber inlet (Rezazadeh et al; 2016). The absorption capacity was

determined as the ratio of the flow rate of CO₂ captured (mCO₂) to the flow rate of the solvent amine (m_L).

and the degree of solvent regeneration (x_{reg}) was calculated as follows (Rezazadeh et al; 2016):

$$x_{reg} = \alpha_{CO_2}^{rich} - (\alpha_{CO_2}^{lean} / \alpha_{CO_2}^{rich}) \quad \text{Eq. (3)}$$

Where, $\alpha_{CO_2}^{rich}$ is the rich CO₂ loading and $\alpha_{CO_2}^{lean}$ is the lean CO₂ loading.

The reboiler (solvent regeneration unit) at the PACT is a kettle type heat exchanger (PACT; 2018) and utilizes the PHW produced in an electric boiler as the source of thermal energy to heat the rich solvent stream for the solvent regeneration/CO₂ stripping. A significant amount of energy is utilized at the CO₂ capture plant for the solvent regeneration energy requirement, i.e. the reboiler (thermal) duty and this is the product of the difference in the enthalpy of the PHW entering and leaving the reboiler and the mass flow rate of the PHW (Akram et al; 2016, Luo et al; 2015). The Specific Reboiler Duty (SRD) was determined as the thermal energy utilized to capture a unit weight of CO₂ whereas the Normalized Specific Reboiler Duty (N-SRD) was ascertained by the ratio of the SRD to the base-case SRD.

The colorimeter test kit mentioned in Section 2.1 is employed to determine the amount of iron present in the solvent produced by the corrosion of the equipment. A coloured complex is formed by the reaction of the Fe with a reagent introduced into the sample, i.e. 1,10-Phenanthroline. The amount of light absorbed by the sample is used to evaluate the concentration of the Fe against an earlier analysed sample without the reagent (Pyenson et al; 1945, Marczenko et al; 2000).

Two experiments were performed in the course of this study. Four tests were embarked on in experiment 1 and this was conducted under varying CO₂ concentrations. The CO₂ concentration in the flue gas of the Combined Cycle Gas Turbine (CCGT) is approximately 4 - 6 vol.% (Merkel et al; 2013) and as a result the CO₂ concentration values from 5.0 to 9.1 vol(%) were selected to simulate the S-EGR (see Table 1) based on the mGT's exhaust CO₂ concentration. The CO₂ was acquired from the onsite CO₂ storage tank and supplied to the SCCP via the GMS along with the ambient air. Using synthesized gas gave the operational flexibility that was required for the experiment. The flue gas was conditioned at approximately 40 °C at the inlet of the absorber. The solvent flow rates were increased from 476.4 kg/hr at the base case test to 975.0 kg/hr at 9.1 vol(%) of CO₂. Further, the flue gas flow was maintained at a flow rate of approximately 171 m³/hr (STP) while the CO₂ concentrations increases with

every test, technically increasing the partial pressure of CO₂. These tests were performed while the CO₂ capture efficiency was maintained at approximately 90 %. A summary of the experimental data is presented in table 1.

In experiment 2, the CO₂ concentration and solvent flow rate were kept constant at approximately 9.1 vol(%) and 983 kg/hr, respectively, whereas the PHW was incrementally increased by 1°C from 124 to 127 °C. These tests were performed to assess the effect of varying the PHW temperature at the inlet of the reboiler on the performance of the CO₂ capture process. The solvent and flue gas were both conditioned at approximately 40 °C before being introduced into the absorber liquid and gas inlets, respectively. A summary of the experimental data is presented in table 2.

3 RESULTS AND DISCUSSIONS

The operating conditions and results of experiments 1 and 2 are given as follows in Table 1 and 2.

Table 1 Experiment 1: Summary of the tests data (varying CO₂ concentrations)

Parameter	Unit	Test 1	Test 2	Test 3	Test 4
CO ₂ concentration	vol%.	5.0	6.6	7.7	9.1
FG flowrate	m ³ /hr (STP)	170.1	171.1	169.3	171.4
FG inlet temperature	°C	41.7	39.4	39.7	39.3
FG outlet temperature	°C	48.1	49.1	42.4	41.0
Lean solvent flowrate	kg/hr	476.4	570.7	853.4	975.0
L/G ratio	kg/hr	2.1	2.5	3.8	4.2
PHW in temperature	°C	125.7	125.6	125.7	125.7
PHW out temperature	°C	122.7	122.6	121.6	121.4
Rich loading	mol CO ₂ /mol MEA	0.378	0.379	0.392	0.399
Lean loading	mol CO ₂ /mol MEA	0.262	0.267	0.304	0.319

CO ₂ captured	kg/hr	14.9	19.9	23.0	26.9
Solvent flow/CO ₂ captured	-	32.0	28.7	37.1	36.2
Capture efficiency	%	89	89	89	88
Absorption capacity	g(CO ₂)/kg(Solvent)	81.4	89.6	70.4	74.0
Normalized Specific Reboiler Duty (SRD)	-	1.00	0.75	0.92	0.86
decrease in energy consumption	%	-	25.1	8.2	16.3
Degree of solvent regeneration	%	30.69	29.55	22.45	20.05

Table 2 Experiment 2: Summary of the tests data (varying PHW temperature)

Parameter	Unit	Test 1	Test 2	Test 3	Test 4
CO ₂ concentration	-		9.1		
FG flowrate	m ³ /hr (STP)	168.9	170.2	171.4	169.5
FG inlet temperature	°C	40.9	40.1	39.3	41.0
FG outlet temperature	°C	41.8	41.8	41.0	41.5
Lean solvent flowrate	kg/hr	986.1	981.3	975.0	976.7
PHW in temperature	°C	123.8	124.8	125.7	126.8
PHW out temperature	°C	119.6	120.5	121.4	122.2
Rich loading	mol CO ₂ /mol MEA	0.414	0.419	0.399	0.373
Lean loading	mol CO ₂ /mol MEA	0.344	0.328	0.319	0.277
CO ₂ captured	kg/hr	23.4	25.4	26.9	27.7
Solvent flow/CO ₂ captured	-	42.1	38.6	36.2	35.3
Capture efficiency	%	78	83	88	91

Absorption capacity	g(CO ₂)/kg(Solvent)	64.5	71.2	73.9	74.6
Normalized Specific Reboiler Duty (SRD)	-	1.00	0.94	1.07	1.12
Change in energy consumption	%	-	6.2	-3.7	-12.3
Degree of solvent regeneration	%	16.91	21.72	20.05	25.74

3.1a Experiment 1: Influence of the CO₂ concentrations on the solvent loadings

The lean CO₂ loadings were observed to increase from 0.262, to 0.319 with increasing CO₂ concentration from 5.0 to 9.1 vol(%). This behaviour represents a percentage increase of 21.8 % between the aforementioned CO₂ recycle ratios. This trend was as a result of increasing the CO₂ concentration in the CO₂ rich solvent stream and solvent flow rate entering the stripper and reboiler as the reboiler thermal input was kept constant. As the lean CO₂ concentration increases, so does the rich CO₂ loading. The rich loading was further facilitated by increasing the CO₂ partial pressure in the flue gas with increasing the CO₂ concentration. A percentage increase of 5.56 % in the rich loading was observed from 0.378, to 0.399 between the aforementioned CO₂ concentrations. The rich loading would have been higher in the absence of increasing the solvent flow rate, which was increased in order to maintain the CO₂ capture efficiency of approximately 90 %. Notz et al. (2010), Akram et al. (2016) and Rezazadeh et al. (2016) all reported increasing CO₂ loadings of the solvent with increasing CO₂ concentrations. While Akram et al., 2016 reported a 14.2 % increase in the rich loading between 5.5 to 9.9 vol(%) CO₂ and which translates to 2.8 % increase for every unit percentage increment in CO₂ concentration with 30 wt% MEA. 5.5 % increase of rich loading from 5.0 to 9.1 vol% CO₂ was observed with 40 wt% MEA due to higher L/G ratio. Running the same process conditions at the same plant with higher MEA concentration is expected to produce higher rich loading for every unit percentage increment in CO₂ concentration, enhance CO₂ capture recovery and reduce the N-SRD. Brigman et al; (2014) has reported a reduction of SRD by 14 % when the amine concentration was increased from 30 to 40 wt(%). This is in line with a study by Abu Zahra et al., 2007 and Akram et al 2020., which established that higher MEA concentrations translates into a higher rich loading due to the increasing driving force for the CO₂ absorption. Higher CO₂ loading was however, reported to facilitate solvent degradation,

the products of such degradation were reported to facilitate corrosion in the CO₂ capture equipment, especially in the presence of oxygen (Gouedard et al., 2014). Degradation of the amine solvent is however beyond the scope of this study.

3.1b Experiment 2: Solvent loadings as a function of increasing PHW temperature

The lean and rich CO₂ loadings at a fixed CO₂ concentration of 9.1 vol(%) were observed to decrease with increasing PHW temperature. Both the lean and rich loadings decreased from 0.344 to 0.277 and 0.414 to 0.383, respectively. This represents a decrease of 19.48 % and 7.49 % for the lean and rich loadings, respectively. By increasing the PHW temperature, regeneration of the solvent intensifies, provided that the solvent flow rate and the CO₂ partial pressure remains constant, and consequently this produces a lower lean loading. The lean loading was observed to decrease by about 19.48 % as noted earlier with increasing PHW temperature, thus suggesting the influence of increasing the PHW temperature from 124 to 127 °C on the CO₂ desorption. As the lean CO₂ decreases, so does the rich loading with increasing the PHW temperature, thus signifying an increasing capacity to capture more CO₂ as the PHW temperature is increased. This behaviour is in agreement with the study by Sakwattanapong et al. (2005). For the reason that studies have reported thermal degradation to be facilitated above 130 °C, PHW inlet temperature was kept below the thermal degradation threshold at 127 °C (Rezazadeh et al; 2016, Global CCS Institute; 2019). This finding was on the backdrop of a study by Rochelle et al; 2009 who established that MEA degrades at a pace of about 2.5 – 6 % weekly when held at 135 °C.

3.2a Experiment 1: Influence of the CO₂ concentration on the Normalized Specific Reboiler Duty

The N-SRD is expected to decrease with increasing the CO₂ concentration in the flue gas stream prior to introduction at the absorber gas inlet. More CO₂ is captured with increasing the CO₂ concentration in the flue gas due to the higher driving force for the CO₂ absorption, thus increasing the value of the SRD (Akram et al; 2016, Li et al; 2011, Notz et al; 2012). However, the N-SRD in these tests was observed to behave in a sinusoidal manner as demonstrated in Fig. 2. With increasing the CO₂ concentration from 5.0 to 9.1 vol(%), the solvent flow was increased by 19.8 % to maintain a CO₂ capture efficiency at 90 % and as a result increases the amount of CO₂ captured. This reduces the N-SRD in spite of the increase in the reboiler heat duty utilization. With increasing CO₂ concentration in the flue gas (above a CO₂ concentration

of 5.0 vol(%) and the solvent flow rate, the solvent residence time in the reboiler drastically reduces, and more reboiler thermal energy is consumed for CO₂ desorption to accommodate the increased solvent flow rate. This increase in reboiler duty utilization causes a higher N-SRD despite the higher CO₂ captured. The marginal decrease in N-SRD from 7.7 to 9.1 vol(%) of CO₂ was due to the increased CO₂ capture at a slightly higher CO₂ partial pressure. Also, it was noted that a good indicator of the behaviour of the N-SRD is the ratio of the solvent flow rate to CO₂ captured as shown in Fig. 2. The higher is the ratio of the solvent flow rate to CO₂ captured, the higher is the N-SRD when both are represented as a function of the CO₂ partial pressure. This finding is on the backdrop of a study by Akram et al., 2020 who reported a decrease in the N-SRD as the MEA concentration was increased from 30 to 40 wt(%) by about 14 %. Increase in the CO₂ recycle ratio is expected to further deliver lower N-SRD due to increasing capacity of the amine solvent to absorb more CO₂.

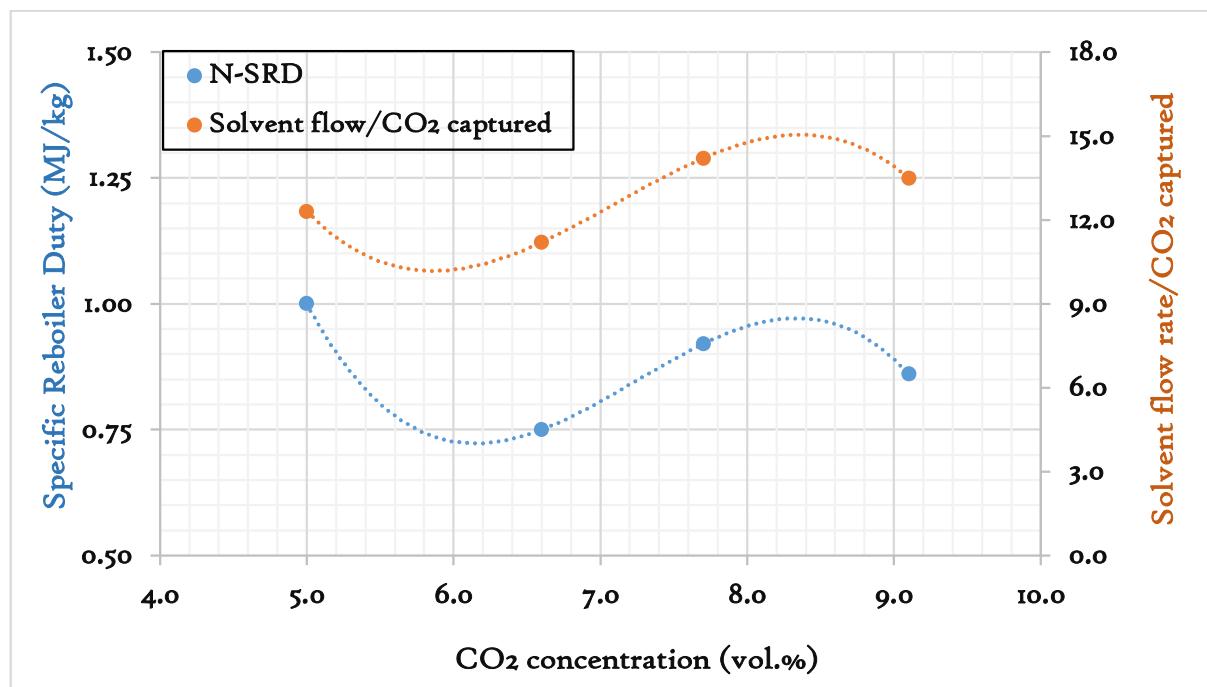


Fig. 2 N-SRD and ratio of solvent flow to CO₂ captured as a function of increasing the CO₂ concentration.

Figure 3 gives an insight into the behaviour of the N-SRD in experiment 1. From 5.0 to 6.6 vol(%) of CO₂, the absorption capacity increased from 81.4 to 89.6 g/kg and the energy consumption reduced by 25.1 %, thus prompting a decrease in the N-SRD from 1.0 to 0.75. The operational conditions at CO₂ concentration of 6.6 vol(%) is thus observed as the recirculation ratio with the lowest SRD in these tests, this value is also where the absorption capacity and reduction in energy consumption were highest. The absorption capacity and

reduction in the energy consumption both decreased thereafter a 6.6 vol(%) of CO₂ as shown in Fig. 3, and this is due to the increased reboiler heat duty utilization with increasing the solvent flow rate to maintain the CO₂ capture efficiency at approximately 90 %.

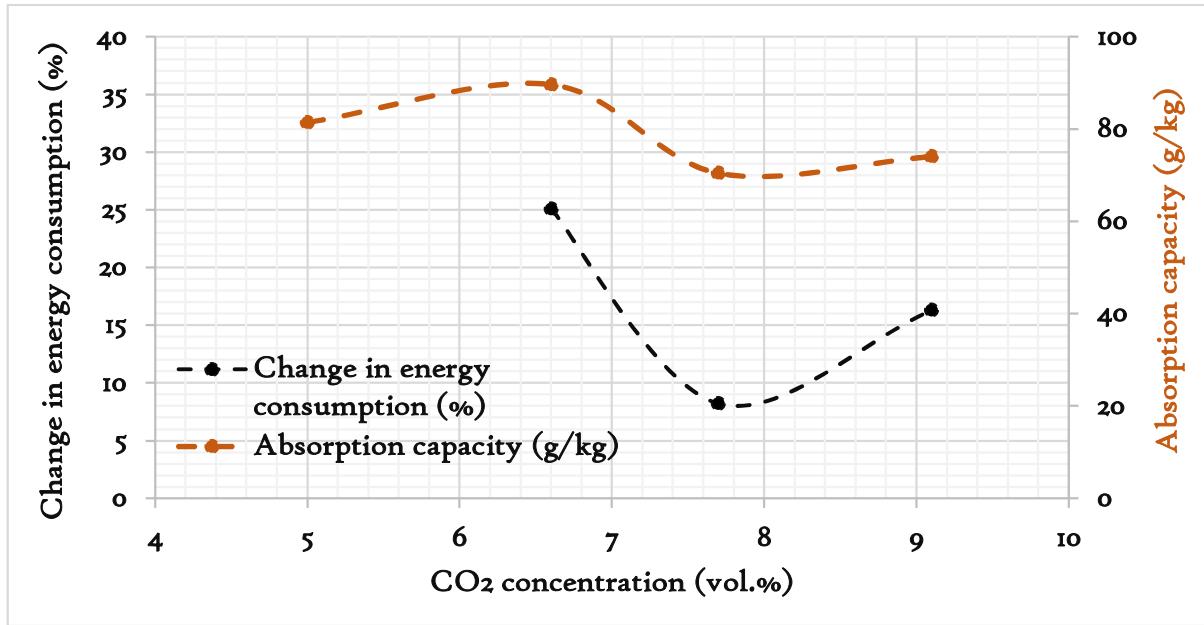


Fig. 3 Change in the energy consumption against the base-case and the absorption capacity as a function of the CO₂ concentration.

3.2b Experiment 2: Influence of the PHW temperature on the Normalized Specific Reboiler Duty

The N-SRD was observed to decrease from 124 °C of PHW temperature to 125 °C and increase thereafter from 125 °C to 127 °C as demonstrated in Fig. 4. The first phase in the decreasing N-SRD is suggestive of increasing the CO₂ capture due to the increasing capacity of the leaner solvent to capture more CO₂, this behaviour was also reported by a study by Akram et al, 2020. However, from 125 to 127 °C, the reboiler duty was high to translate into a decrease in the N-SRD despite capturing more CO₂. 125 °C was observed to be the PHW temperature with lower value of the N-SRD in this study. Also, the PHW temperature has a pronounced impact on the CO₂ capture efficiency which was increased from 78 to 91 % when PHW temperature was increased from 124 °C to 127 °C. This result is also in agreement with a study report by Akram et al 2020, who also observed an increment of capture efficiency from 72 to 88 %. A leaner solvent, due to the increasing PHW temperature, has a greater capacity to capture the CO₂ and consequently has a better CO₂ capture efficiency. The ratio of the solvent flow to CO₂ captured in these tests took a decreasing linear trend because the liquid-gas ratio and the flue gas CO₂

concentrations were kept constant. Fig. 4 shows the SRD and the ratio of the solvent flow to CO₂ captured as a function of the PHW temperature.

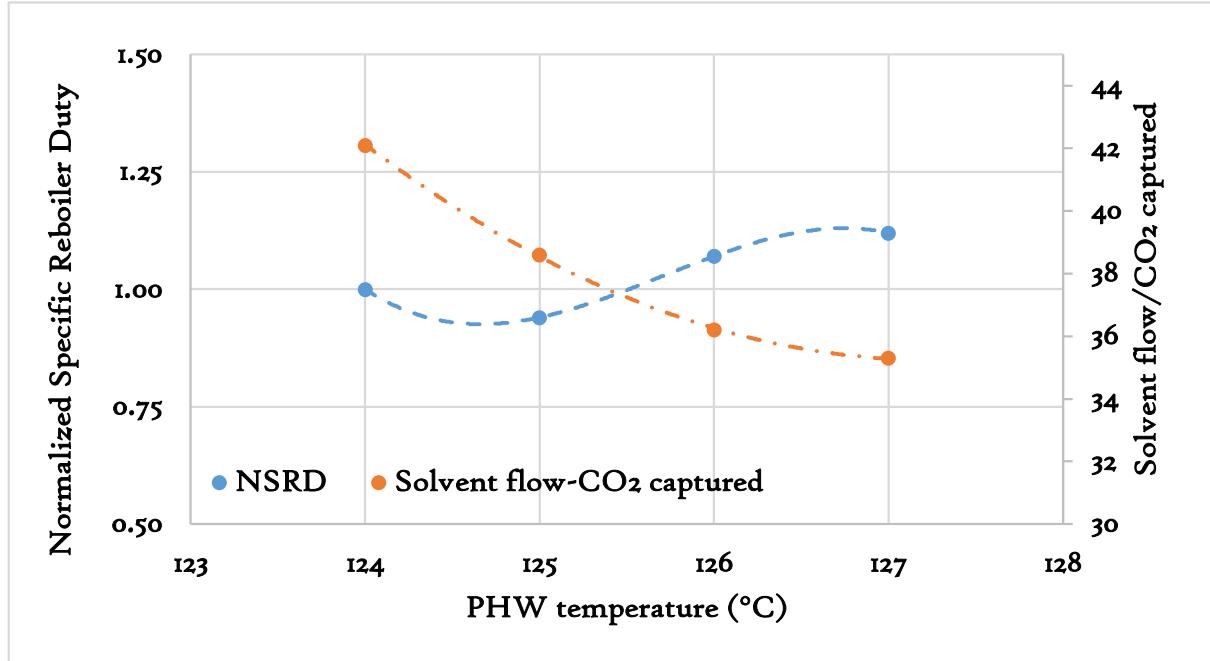


Fig. 4 N-SRD as a function of the PHW temperature in experiment 2.

The absorption capacity with increasing PHW temperature increased by 15.7 % from 64.5 to 74.6 g/kg. The percentage energy consumption decreased by 6.2 % from 124 – 125 °C and then increases to 12.3 % at 127 °C against the base case N-SRD. The test at 125 °C PHW temperature was observed to be lower in the N-SRD and this was because above this temperature the reboiler heat input was high to achieve lower value of N-SRD and below this temperature the CO₂ captured is low to translate to a low N-SRD. Fig. 5 shows the change in energy consumption and absorption capacity as a function of the PHW temperature.

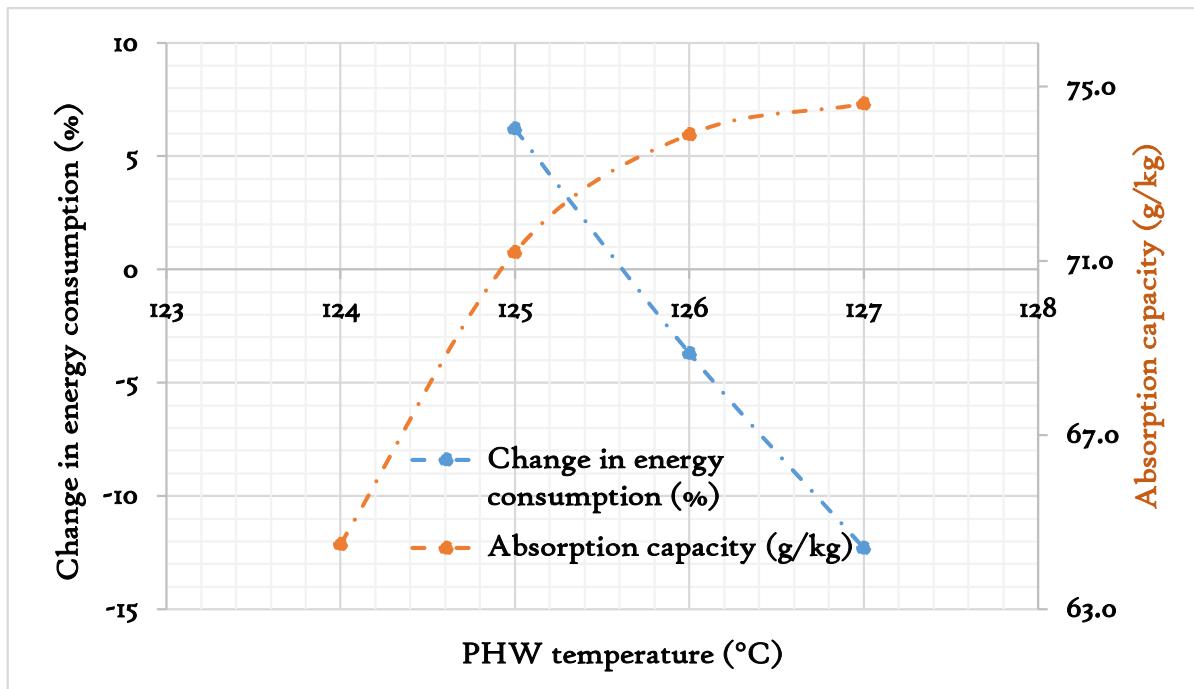


Fig. 5 Change in energy consumption against the base-case and absorption capacity as a function of the PHW temperature in experiment 2.

3.3a Experiment 1: Influence of the CO₂ concentration on the CO₂ capture efficiency

The CO₂ capture efficiency in experiment 1 was maintained at approximately 90 % as the CO₂ recycle ratio increases. The CO₂ captured was observed to increase from 14.9 to 26.9 kg/hr with increasing the CO₂ concentration due to the increasing rich loading. This trend is in line with earlier published studies (Akram et al; 2016, Diego et al; 2017, Li et al 2011) and represents a significant increase in CO₂ captured by 80.5 % from 5.0 to 9.1 vol(%) of CO₂. The increase underscores the influence of the driving forces for the CO₂ capture as the CO₂ partial pressure increases and under the influence of higher solvent concentration at 40 wt(%) MEA (Notz et al; 2012).

3.3b Experiment 2: Influence of the PHW temperature on the CO₂ capture efficiency

The CO₂ captured was also observed to increase from 23.4 to 27.7 kg/hr with increasing PHW temperature from 124 to 127 °C. As noted earlier, a leaner solvent stream is produced with an increasing PHW temperature at the reboiler hot-water inlet and as a consequence, a capacity to capture more CO₂, thereby increasing both the CO₂ capture efficiency and capture rate by 16.7 % and 4.3 kg/hr respectively. Prior study (Akram et al., 2020) has also noted on the importance of PHW temperature on the CO₂ capture efficiency. These results signify the impact of

increasing the PHW temperature on the amount of CO₂ recovery. However, the higher the PHW temperature, the higher the risk of thermal degradation of the solvent. Thermal degradation is more likely to transpire in regions of an elevated temperature in the CO₂ capture plant and this includes the reboiler (stripper sump) and lower part of the stripper (Rochelle et al; 2012, Davis et al; 2009).

3.4a Experiment 1: Influence of the CO₂ concentration on the solvent regeneration

The degree of solvent regeneration was observed to decrease from 30.69 to 20.05% with increasing CO₂ concentration from 5.0 to 9.1 vol(%) and this represents a 34.67 % decrease in the solvent regeneration, and this trend is in agreement with the published studies by Rezazadeh et al., 2011 and Akram et al., 2020. This trend was observed as the liquid-gas ratio increases from 2.1 to 4.2, representing a 100 % increase in the L/G. The decline in the degree of regeneration with increasing CO₂ concentration was due to the decreasing resident time of the solvent in the reboiler and stripper as the solvent flow rate was increased to maintain the capture efficiency at approximately 90 %. Fig. 6 shows the degree of solvent regeneration and L/G as a function of the CO₂ recycle ratio.

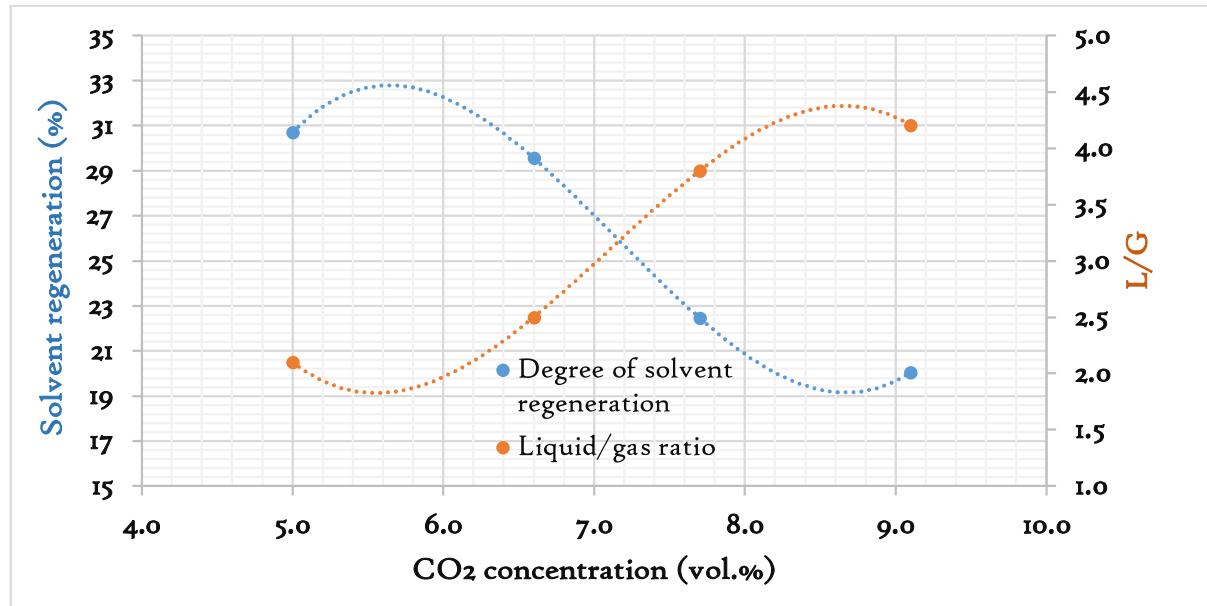


Fig. 6 Degree of the solvent regeneration and L/G as a function of the CO₂ concentration.

3.4b Experiment 2: Influence of the PHW temperature on the solvent regeneration

The degree of solvent regeneration was observed to take a different trend when the CO₂ recycle ratio was fixed at 6.12 and the PHW temperature increased from 124 °C to 127 °C and this is because the liquid/gas ratio was maintained at approximately 4.3. As anticipated, an increase in the solvent regeneration was observed with increasing PHW temperature as presented in Fig. 7.

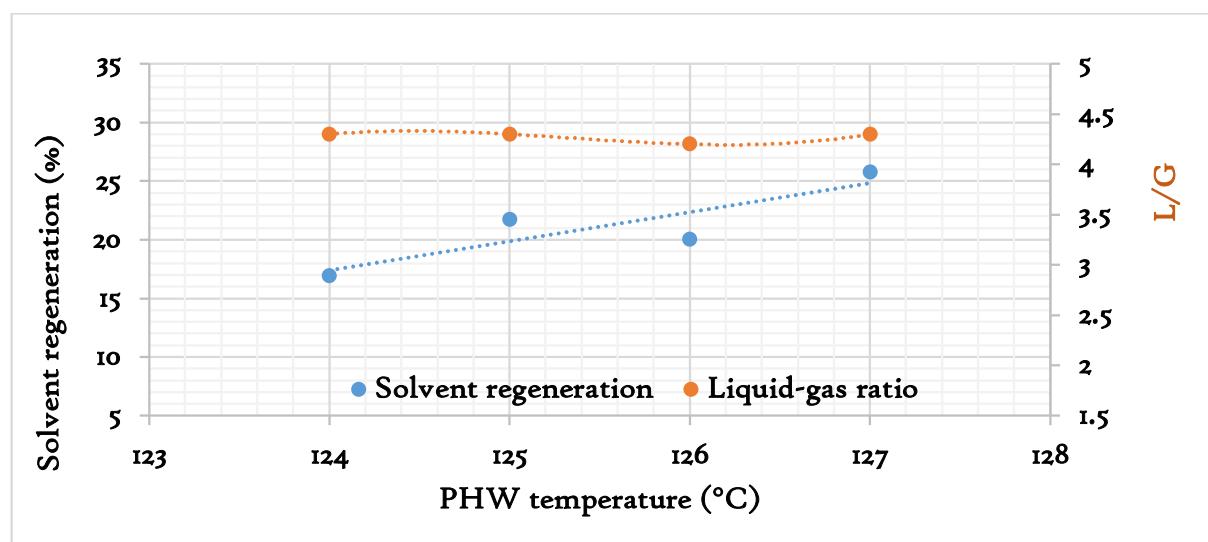


Fig. 7 Degree of the solvent regeneration and L/G as a function of the PHW temperature in experiment 2.

3.5a Experiment 1: The absorber temperature profile with varying CO₂ concentration

The absorber temperature profile gives an insight into the temperature behaviour of the fluid across the length of the absorber reactor. About 6.6 % increase in temperature was observed in the absorber temperature profile at the peak region when the CO₂ concentration was increased from 5.0 to 9.1 vol(%) as shown in Fig 8. This suggests increasing exothermic CO₂ absorption reaction as the CO₂ partial pressure increases. Also, a noticeable temperature bulge around the middle section of the absorber is also an indicator that the gas and liquid in the bulge region is close to equilibrium with regards to the CO₂ concentration (Kvamsdal et al; 2008, Sanpasertparnich et al; 2011, The Contactor; 2015, Digital Refining; 2017, Rezazadeh et al; 2017). However, the temperature profile at the thermocouples located at the top of the absorber column appears to decrease with increasing CO₂ concentration. This could be attributed to cooler and higher solvent flows into the absorber, which diminishes the rate at which the temperature builds up due to the CO₂-solvent reaction. This phenomenon is also behind why

the flue gas outlet temperature decreases with increasing CO₂ concentration. Identifying region of highest temperature bulge could be valuable with regards to enhancing CO₂ absorption via withdrawing part of the solvent from the bulge region and transporting it to the top of the absorber column or routing a fraction of lean amine solvent to the bulge region (Moulléc et al; 2014). The temperature profile at CO₂ concentration at 7.7 vol(%) appears to be higher than 9.1 vol(%) of CO₂ due to the likelihood that higher flow rate at 9.1vol(%) of CO₂ may have impacted a cooling effect on exothermic CO₂ absorption. Higher CO₂ recycle ratio is to be avoided to prevent risking reaching the flash point of the amine solvent, which is expected to reduce the solvent capacity to capture CO₂. Higher recycle ratio also promote the tendency to reach equilibrium pinch, which should also be avoided in order to enable the driving force of CO₂ absorption by the amine solvent (Akram et al., 2020).

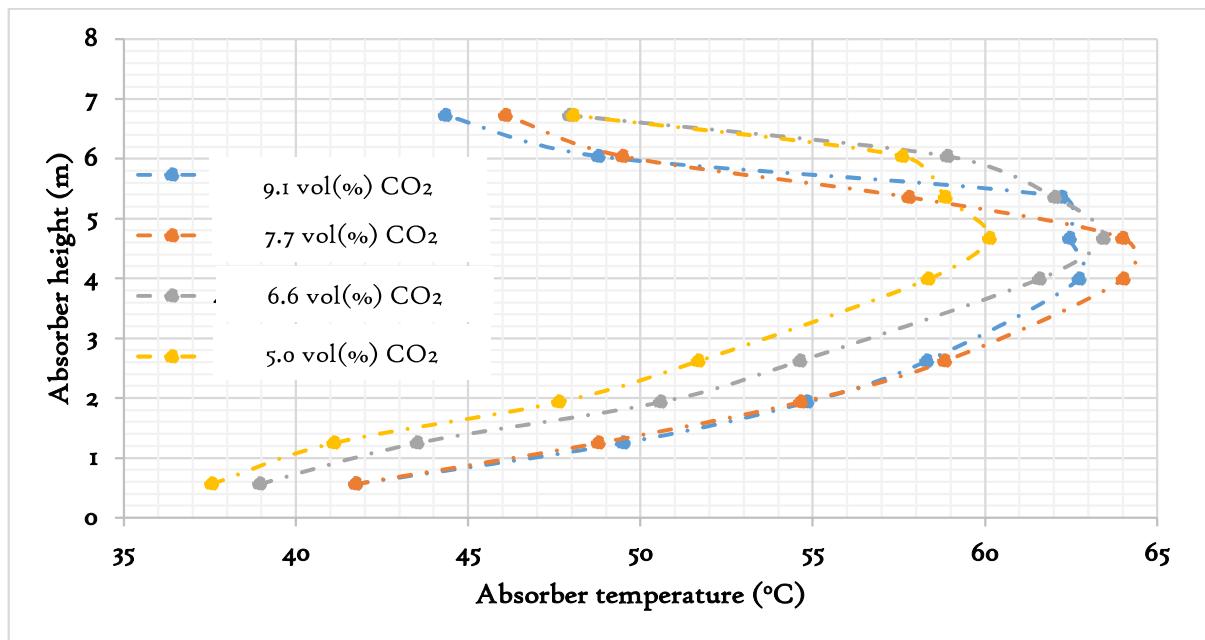


Fig. 8 Absorber temperature behaviour with increasing CO₂ concentrations.

3.5b Experiment 2: The absorber temperature profile with varying PHW temperatures

The higher is the PHW temperature, the lower is the lean loading and the higher is the solvent absorption capacity and consequently the higher is the exothermic CO₂ capture reaction and this results in a higher temperature prevalence in the absorber reactor. Thus, the PHW temperature at 127 °C was observed to have the highest absorber thermal profile and 124 °C the lowest absorber thermal profile as shown in Fig. 9.

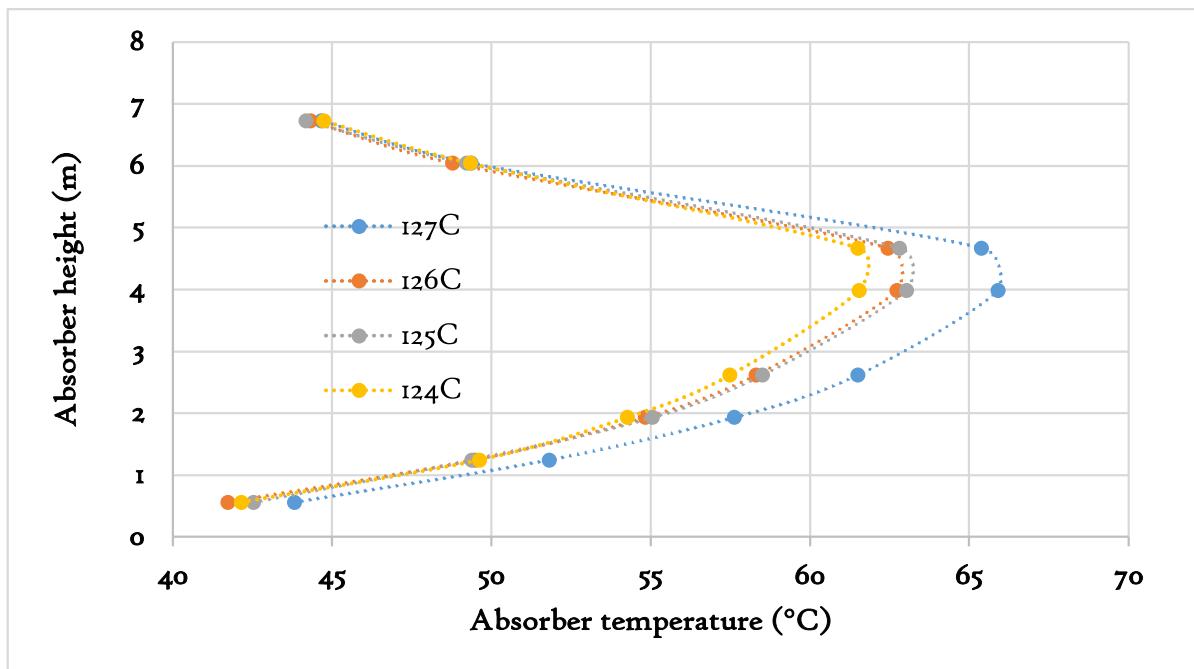


Fig. 9 Absorber temperature behaviour with increasing PHW temperature.

3.6a Experiment 1: The stripper temperature profile with varying CO₂ concentration

Solvent regeneration is an endothermic process and high thermal energy is required to dissociate the solvent and CO₂ bonding (Akram et al; 2016, Zhang et al; 2014, Gao et al, 2017, Li et al; 2013, Soltani et al; 2017). The stripper temperature was observed to decrease by about 3 – 9 % at the temperature bulge with increasing CO₂ concentration and this is believed to be due to the increasing endothermic reaction of the solvent-CO₂ dissociation as shown in Fig. 10. Also, the temperature is observed to be higher at the lower part of the stripper column and this is due to the influence of the hot captured CO₂ and water vapour leaving the reboiler and entering the stripper column. The stripper temperature profile was not observed to have a smooth curve and this may be attributed to changes in fluid properties under the physical and thermochemical environment which include preferential pathways across the Sulzer Mellapak CC3: Structured packing, changes in viscosity, surface tension, vapour pressure, capillarity and changes in concentration of solvent degradation products of the fluid across the length of the stripper column. The apparent decrease in temperature at the top and bottom of the stripper is thought to be credited to the cooling effect of the rich solvent flow rate and intensified endothermic CO₂ dissociation reaction at the bottom of the stripper respectively. Fig. 10 shows

the relationship between the stripper temperature and the CO₂ concentration along the length of the stripper.

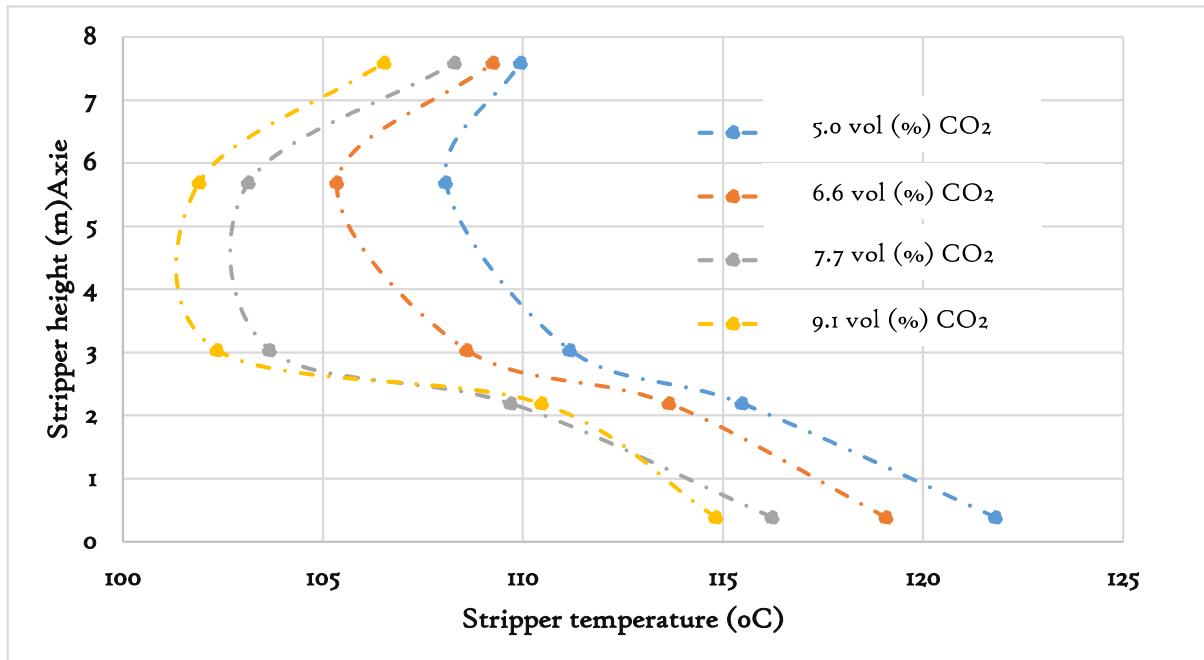


Fig. 10 Stripper temperature behaviour with increasing CO₂ concentration.

3.6b Experiment 2: The stripper temperature profile

The stripper temperature was observed to increase by about 10.42 % at the highest temperature bulge with increasing PHW temperature from 124 to 127 °C. As the PHW temperature increases, the captured CO₂ and water vapour travelling through the stripper becomes hotter and this manifests in a higher temperature profile in the stripper. A noticeable increase in temperature was observed within the bottom 2 m of the stripper column, indicating an elevated endothermic CO₂ desorption process at that region, this is where the temperature is highest in the stripper. The cooling effect at the top of the stripper is attributed to the rich solvent stream flow going into the absorber at relatively lower temperature than the stripper temperature. The stripper temperature profile with increasing PHW temperature is given in Fig. 11.

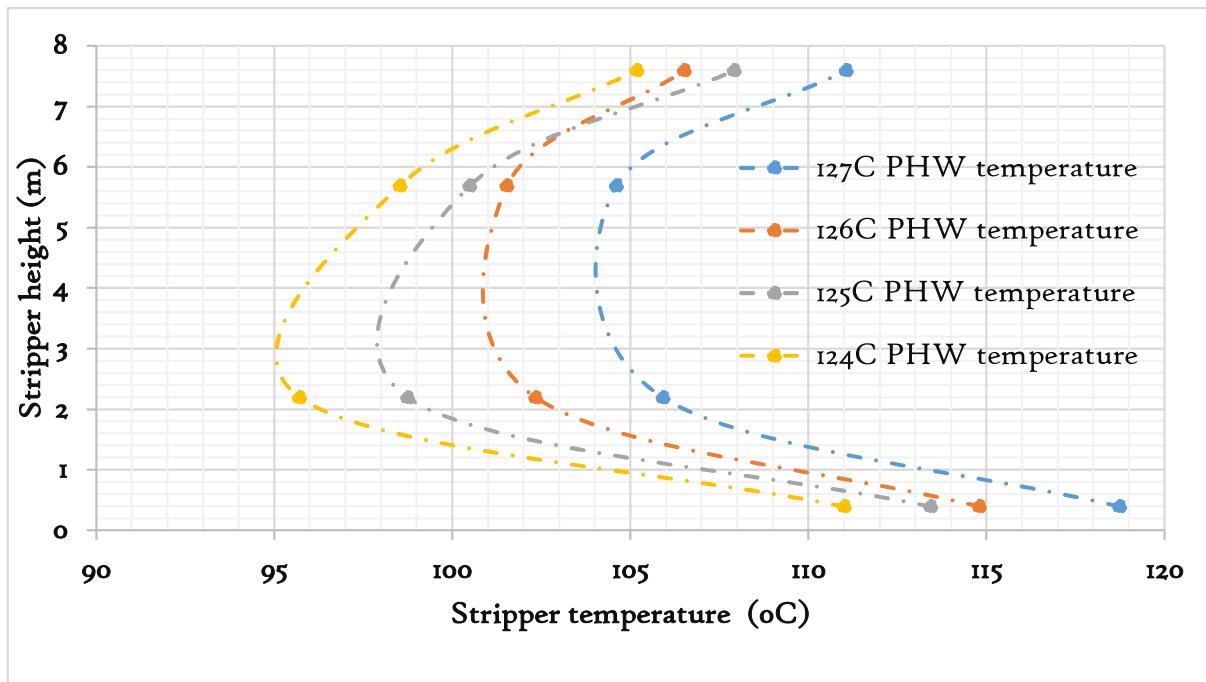


Fig.11 Stripper temperature behaviour with increasing the PHW temperature.

3.7a MEA emissions with increasing CO₂ concentration

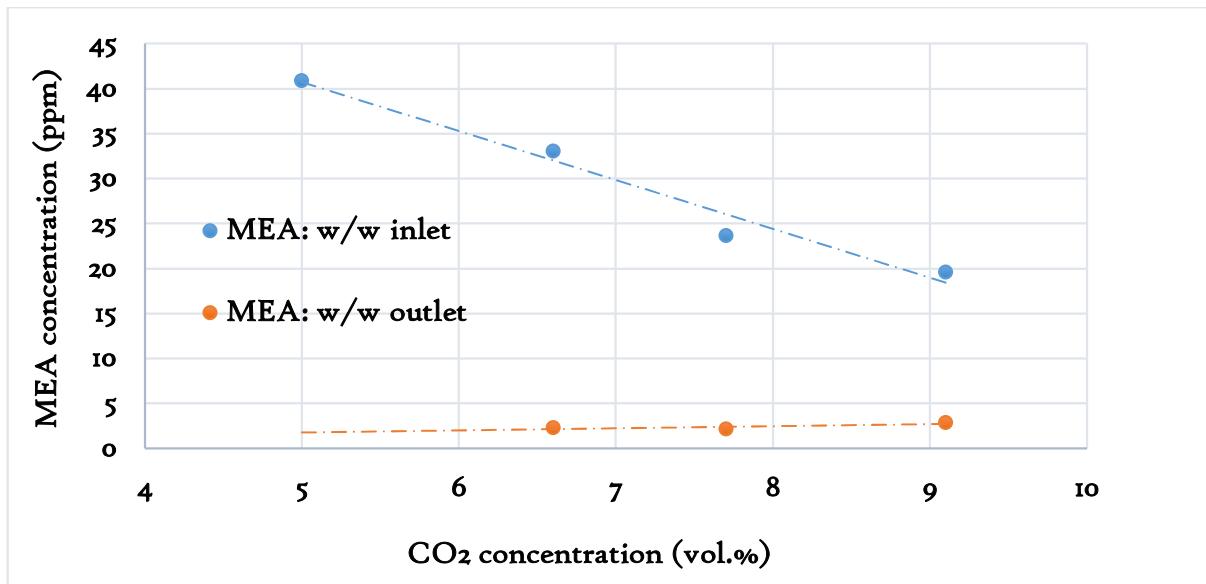


Fig. 12 The MEA emission with varying CO₂ concentrations in Experiment 1.

3.7b MEA emissions with increasing PHW temperature

The MEA at the absorber outlet is observed to increase from 12.2 to 19.6 ppm as the PHW temperature is increased from 124 to 127 °C. In Section 3.2, it was observed that the leaner is

the solvent; the higher is the absorption capacity and consequently the higher is the exothermic reaction in the absorber. This results in a higher amount of MEA being carried over by the evaporating water into the water-wash column. The MEA at the water-wash column inlet and outlet was recorded to be between 12.2 to 19.57 and 2.8 to 4.7 ppm, respectively, and this signifies about 76% of the water-wash efficiency as shown in Fig. 14.

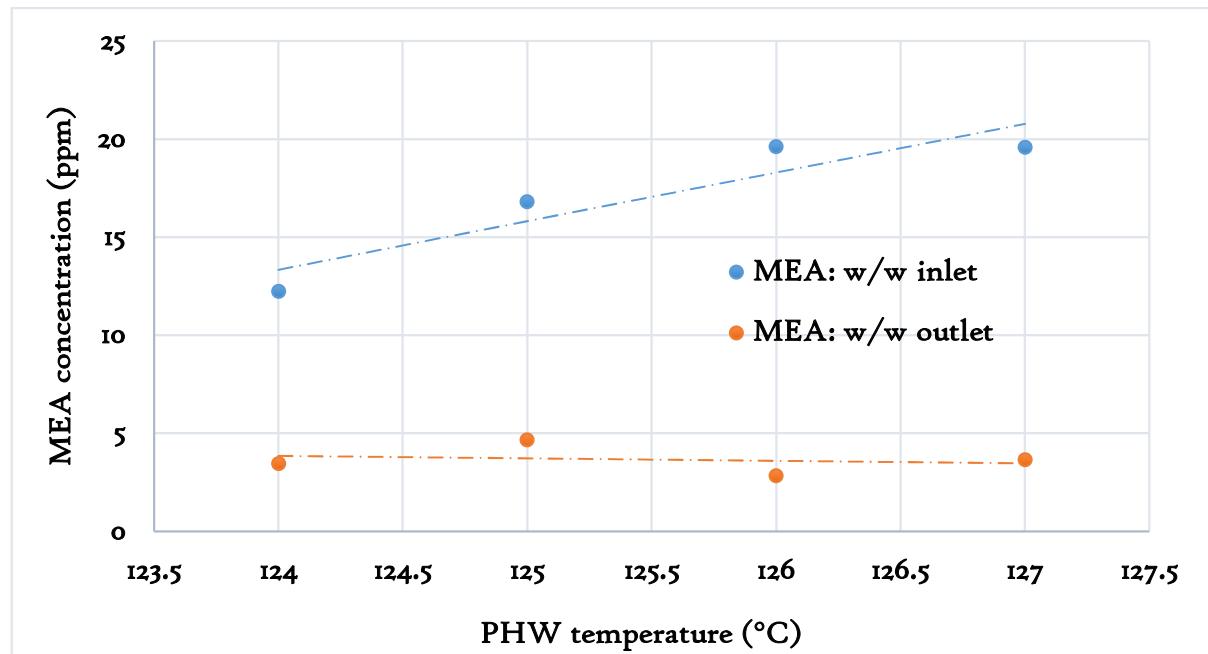


Fig. 14 The MEA emission with varying PHW temperature in Experiment 2.

3.8 Corrosion

Corrosion is a key challenge in an aqueous solution of amine-based carbon capture installations, and this is envisaged to increase the OpEx as well as the CapEx of the SCCP. Recent research has suggested that this phenomenon is caused as a result of anodic (iron dissolution) and cathodic (reduction of oxidisers present in the solution) electrochemical reactions on metal surfaces (Carbon Capture Journal; 2012). Furthermore, higher levels of CO₂ loadings, elevated levels of amine concentration, higher levels of operating temperature, higher levels of O₂ concentration, presence of Heat Stable Salts (HSS) and mobile solid particles are all observed to facilitate corrosion (J. Kittel, S. Gonzalez; 2014, Carbon Capture Journal; 2012, Gouedard et al; 2012). The result from this study appear to support the hypothesis that increase in Fe concentration is a key indicator of corrosion of the carbon capture internal equipment and a gauge to how the amine solvent has degraded as Fe concentration was observed to increase by 70.7% (from 9.16 to 15.64 mg/L) within the course of the two experiments. This noticeable

amplification of Fe concentration in these experiments underscores the impact of the corrosion in the carbon capture using an aqueous solution of amines. However, employing lower MEA concentration at 30 wt(%) MEA has been demonstrated on a pilot-scale study to corrode the plant's internal equipment slower than higher MEA concentration at 40 wt(%) MEA (Akram et al., 2020).

The treatment of flue gas in order to get rid of exhaust impurities, reduction of oxygen levels via Dissolved Oxygen Removal Apparatus (DORA) (Monteiro et al; 2018), designing corrosion resistant equipment and using corrosion inhibitors are expected to reduce the severity of corrosion processes.

CONCLUSION AND FUTURE WORK

The experimental campaigns were undertaken at the UKCCSRC-PACT facility in Sheffield, UK to evaluate the performance of the CO₂ capture process using 40 wt(%) MEA at

- i. Varying the CO₂ concentrations (simulating S-EGR) with a cap of 90% CO₂ capture efficiency and
- ii. Varying the PHW temperatures at the reboiler inlet.

In conclusion

- Increasing the partial pressure of the CO₂ in the flue gas of the gas turbines via increasing the S-EGR ratios under the influence of 40 wt(%) MEA to enhance the CO₂ absorption and reduce the N-SRD is advantageous only up to a certain CO₂ concentration i.e. 6.6 (vol%) based on the process conditions and plant specifications used in this study. Above which the N-SRD begins to increase. A good indicator of the N-SRD trend was observed to be the ratio of solvent flow/CO₂ captured and the absorption capacity, which is lowest and highest respectively when the N-SRD is at its lowest value. Achieving a lower value of the N-SRD via S-EGR and at an elevated amine concentration however comes with the risks of potentially advancing solvent degradation and system corrosion, which consequently translates to increasing OpEx and CapEx. Higher CO₂ recycle ratios induces a thermal prominence in the absorber column and as a result, risk the possibility of reaching the solvent's flash point

consequently reducing the performance of CO₂ capture. Further study, which is beyond the scope of this study is required for sustainable de-oxygenation techniques of the flue-gas to counteract oxidative solvent degradation.

The PHW temperature has a profound effect on the N-SRD. This study has determined based on the operating conditions used that the N-SRD was lowest at a certain PHW temperature i.e. 125 °C based on the plant's test process conditions utilized, below and above this identified temperature the N-SRD increases. The CO₂ capture absorption capacity increased in an almost linear manner with increasing the PHW temperature from 124 – 127 °C, but despite this behaviour, the N-SRD was observed to increase above 125 °C. For the reason that studies have reported that reboiler temperature above 130 °C accelerates the thermal degradation of the solvent, the PHW temperature above the thermal degradation threshold should only be utilized for flash CO₂ desorption in order to either promptly increase the CO₂ capture rate or lower the lean CO₂ loading.

- Higher solvent flow limits the emission of MEA from the absorber gas outlet as it induces a cooling effect at the top of the absorber reactor, thereby negating traces of MEA and associated volatile compounds to be carried over into the atmosphere, where it can degrade to form carcinogenic compounds in the form of nitrosamines and nitramines. However, higher liquid/gas leads to higher recurrent expenditure but lower cost of solvent treatment and replacement due to lower solvent stress.
- The Fe concentration was observed to increase to about 70.7 % in the course of the pilot-plant tests, underscoring equipment corrosion as a major challenge in CO₂ capture plants.

DECLARATION OF COMPETING INTEREST

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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