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Comparative Energy and Environmental Performance of 40% and 30% Monoethanolamine at PACT Pilot Plant

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1. Abstract:

Post combustion CO_2 capture using amines is one of the most well understood processes. The most widely used and studied solvent for this purpose is 30% Monoethanolamine (MEA). The main issue with the process is the use of energy for stripping CO_2 out of the solvent. It is anticipated that higher concentrations of MEA can capture a higher amount of CO_2 and thus reduce energy consumption but may also result in a worsening of the environmental emissions due to potential increase in corrosion and solvent degradation.

In order to study the impact of 40% MEA (as opposed to 30% MEA) on the capture plant performance, a test campaign was carried out at the Pilot Scale Advanced Capture Technology (PACT) facilities of the UK Carbon Capture and Storage Research Centre (UKCCSRC) using 30% and 40% MEA. The absorber (9 m height x 0.3 m dia.) is packed with 28 sections (6.5 m) of Mellapak CC3 structured packing. The absorption column temperature profile is measured by 10 RTDs installed around 48 cm apart along the column length. The performance of the capture plant in terms of reboiler duty, capture efficiency, loading capacity and liquid to gas ratio is evaluated at different operating conditions.

It has been found that specific reboiler duty using 40% MEA drops by up to 14% as compared to that with 30% MEA under similar test conditions. It has also been observed that the process is very sensitive to reboiler temperature and slight changes in reboiler temperature can have a significant impact on the plant performance. Moreover, similar energy and capture performance can be achieved at different reboiler temperatures with right combination of temperature and pressure in the reboiler/stripper. Corrosion rate was found to be higher with 40% MEA than 30% MEA. Solvent degradation rate and solvent carry over has also indicated slightly higher levels for 40%. Water wash was shown to be effective in recovering most of the MEA from the flue gas.

Keywords: Monoethanolamine; Next generation solvents; Carbon capture; Solvent degradation

Nomenclature:

CCS	Carbon Capture and Storage
CCUS	Carbon Capture Utilisation and Storage
CHP	Combined Heat and Power
FGD	Flue Gas Desulphurisation
FTIR	Fourier-Transform Infra-Red spectroscopy
IMTP	INTALOX Metal Tower Packing
L/G	Liquid to Gas Ratio
MEA	Mono-Ethanol-Amine
NSRD	Normalised Specific Reboiler duty
PACT	Pilot-scale Advanced Capture Technology
PHW	Pressurised Hot Water
RTD	Resistance Temperature Detectors
SRD	Specific Reboiler Duty
TPD	Tonnes Per Day
UKCCSRC	UK Carbon Capture and Storage Research Centre

2. Introduction:

Climate change is becoming an increasingly bigger challenge as time is passing. Efforts are being made to reduce the use of fossils and move towards renewable sources of energy, improving existing technologies and inventing new technologies. Biomass, renewables, efficiency improvements, new technologies and CCUS all are required to save the planet. None of these alone can solve the problem due to technical, environmental, social and availability challenges. All has to play their part if global temperature rise has to be kept under 2 ⁰C.

Carbon capture, utilisation and storage (CCUS) is becoming increasingly important as one of the leading options to tackle climate change. In Oct 2017, UK Government announced new approach to CCUS as part of clean growth strategy. The main aim is to be ready to deploy CCUS at scale during 2030s if costs are significantly reduced. Under this plan, a CCUS task force was setup who delivered its report to the Government in July 2018. The report set out the view of the industry on how best to progress CCUS in the UK. The key message given by the report is that a roadmap is developed jointly by government and industry to show how CCUS can be developed and deployed across all sectors [CCUS Cost Challenge Task Force, 2018]. CCUS deployment pathway published at the end of 2018 has given an action plan aiming at first UK CCUS commercial plant to be operational from Mid-2020s [Clean Growth, 2018].

In order to meet the emissions reduction targets, decarbonising industrial sector is as important as power sector. Post combustion CO_2 capture by absorption using amines is one of the leading technologies to reduce CO_2 emissions. The technology is proven to work at commercial scale. The technology has been used in the industry for more than 60 years mainly applied to gas sweetening plants and petrochemicals but also applied to Urea manufacturing plants. The beauty of this technology is that it can be retrofitted to the existing infrastructure saving investments in costly process changes. The application of the technology to the power sector is relatively new. However, there are some commercial scale facilities operational around the globe. Sask Power (Boundary Dam 3) in Canada and Petra Nova in USA are two such facilities. There are a number of research facilities in industrial and academic setups working on different aspects of the technology (de Cazenove et al. 2016; Akram et al. 2016; Notz et al. 2012; Mejdell et al. 2011). Details of all the carbon capture plants operational, being built or planned can be found on the global CCS institute website.

Pilot-scale Advanced Capture Technology (PACT) situated at University of Sheffield is one such facility working on CO_2 capture from different sources including natural gas, coal, biomass, waste materials and synthetic flue gases. PACT has two of 330kW CHP gas turbines for firing gaseous fuels, 250kW pulverised fuel rig for firing biomass, coal and mixtures, 240kW chip boiler for firing waste wood of different grades.

Although post combustion CO_2 capture is well advanced and understood, it has some drawbacks. The major disadvantage is the cost of CO_2 capture, which reduces the power plant efficiency by 6-7% points. The other issue with this technology is degradation of the solvent due to heating "thermal degradation" and due to the presence of oxygen in the flue gas "oxidative degradation". The chemicals used in this technology are very aggressive and cause corrosion which is another weak point. Process modifications (Kang et al. 2016; Jassim et al. 2007; Le Moullec et al. 2014; Madan et al. 2013; Ahn et al. 2013; Amrollahi et al. 2011; Oh et al. 2018; Diego et al. 2017; Merkel et al. 2013; Herraiz, 2016) and new solvents (Aronu et al. 2010; Kumar et al. 2014; Hakka 2007; Yuan and Rochelle, 2018; Wang et al. 2015; Yang et al. 2016; Kim et al. 2013; Cheng et al.

2013) are being tested to minimise these issues. Monoethanolamine (MEA) 30wt% is the most widely studied solvent. It is understood that higher concentrations of MEA can have better energy performance. Abu-Zahra et al. (2007) have shown by Aspen plus simulations that reboiler duty drops as concentration of MEA in the solvent is increased. Therefore, as MEA has been used in the industry for many decades and its degradation pathways are better understood than most of the new solvents, concentrated MEA can potentially be a next generation solvent.

However, there are only limited studies on the use of 40% MEA as an alternative solvent (Abu-Zahra et al. 2007; Li et al. 2016; Delfort et al. 2011, Brigman et al. 2014; Morken et al. 2014). Abu Zahra et al. performed Aspen simulations to find out optimum conditions for lower reboiler duty by varying a number of parameters including variation of solvent concentration from 20% to 40%. They observed that reboiler duty drops with increase in solvent concentration and obtained minimum reboiler duty with 40% MEA at 0.3 mol/mol lean loading. Li et al. presented simulation results for different MEA concentrations varying from 25% to 40%. Morken et al. presented emissions data and concluded that 40% MEA gives higher levels of MEA emissions from absorber. Delfort et al. presented the impact of oxidation inhibitors on solvent degradation and found that specific inhibitors were able to reduce solvent degradation considerably. The only study authors found in open literature comparing the energy performance of 30% and 40% MEA in a real capture plant is by Brigman et al. (2014).

Above literature citation indicates that although lots of research work has been carried out on the CO_2 capture process using 30% MEA, practical data with higher concentrations of MEA is very limited. Therefore, in this test campaign 40% MEA has been tested under widely varying operational conditions and its energy and environmental performance, wherever appropriate, has been compared with 30% MEA. The impact of different reboiler temperature and stripper pressures has also been evaluated.

3. Materials and Methods:

The tests campaign is carried out on 1 TPD CO_2 capture capacity pilot plant. Brief plant overview and experimental strategy including measurements and calculations is described in the following.

3.1 Plant overview:

The pilot plant is equipped with a Flue Gas Desulphurisation (FGD) column, an absorber, a desorber, a reboiler and ancillary equipment. Simplified process flow diagram of the plant is shown in Fig 1. Absorber and desorber specifications are given in Table 1. Both the absorber and the desorber are 300 mm in diameter. Absorber has 28 sections of Mellapak CC3 packing (total packed height of 6.5 m) while desorber has IMTP25 random packing. The absorber has 10 equally spaced RTDs along the height for temperature profiling. The absorber is also equipped with watch ports to observe foaming.

The plant uses Pressurised Hot Water (PHW) in the reboiler for desorbing CO_2. The hot water is produced by an electric boiler having four electric elements controlled by two separate relays. The temperature of the hot water is varied by user input and controlled by a thermostat.

In order to reduce costs during these tests, synthetic flue gas was used instead of real flue gas. Synthetic flue gas was formulated by mixing CO_2 from a cryogenic storage tank with air. Therefore, FGD has not been used during this test campaign. Flow of CO_2 was measured by a Coriolis type flow meter and controlled by a pneumatic control valve. Flow of mixed gas (air+CO_2) was measured at

the inlet to the absorber using a Pitot tube type flow meter. Solvent flow rate was measured by electromagnetic type flow meters and controlled by pneumatic control valves.

Table 1: Absorber and desorber specifications

Specification	Absorber	Desorber
Diameter (mm)	300	300
Height (m)	9	7.5
Packed height (m)	6.5	7
Packing type	Structured	Random
Packing specification	Mellapak CC3	IMTP25
No. of temperature measurements	10	9

3.2 Sampling and measurements:

In order to assess the performance of the plant, gaseous and liquid streams are analysed. Gas analysis are performed at five different locations of the plant (FGD inlet, absorber inlet, absorber outlet, water wash outlet and desorber outlet) using FTIR. Gas samples are extracted using isokinetic sampling probes. The gas is then passed through heated filters and routed to a heated cabinet via heated lines. The heated cabinet has solenoids for switching between different lines. The sequence of the lines and sampling time can be varied using FTIR software. The heated cabinet has two sets of solenoids for sampling and purging. When one line is being sampled, the next one is being purged to avoid cross contamination. Gas flow for both purging and sampling setups was measured by rotameter type flow meters installed in the heated cabinet. After leaving the heated cabinet, sampling line is routed to FTIR while purging line is routed to a safe location outside the lab. The gaseous stream exiting the FTIR after analysis is also discharged to a safe location outside the lab. In order to avoid condensation in the sampling system and that the analysis are as accurate as possible, the whole sampling system starting from the sampling point on the plant up till the FTIR is kept heated up to 180 ⁰C by thermostatic temperature controllers.
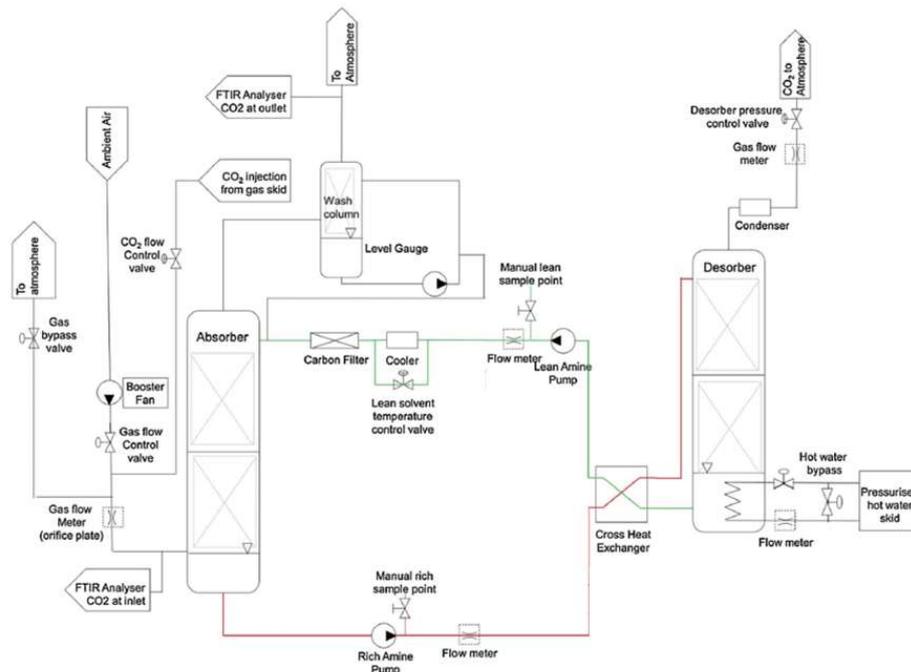

Figure 1: Simplified flow diagram of PACT CO_2 capture plant

For solvent analysis, samples from rich and lean solvent streams were collected from the plant at the end of each test, when the plant was operating at steady state for a period of time. The samples were analysed for CO_2 loadings, MEA concentration and iron content. CO_2 loadings and MEA concentration in rich and lean solvent streams for each test run was measured by endpoint titration method using Mettler Toledo auto-titrator. The samples were titrated against 0.2M HCl to determine MEA concentration and with 0.5M NaOH to determine CO_2 loading. Further details of titration procedure and calculations formulas are given in Akram et al. (2016).

Corrosion rate was measured by the amount of Iron (Fe) build up in the solvent over time. Iron content in the samples was measured using HACH Pocket II Colorimeter. Sample (10 mL) was taken in each of the two bottles. One of the bottle was used as baseline while Ferrover reagent was introduced into the second bottle. The working principle is the measurement of light getting through the sample. Light getting through each of the sample bottles is compared and the difference is correlated to the amount of Fe in the sample.

The apparatus can only measure up to 5 mg/L of Fe. In order to measure higher concentrations, samples were diluted by 50% with demineralised water. In this procedure, same sample is used as span as well as for measurement. Therefore, the impact of change in colour of the aged solvent on the measurements is neglected.

3.3 Calculations:

Capture efficacy is calculated using flue gas composition and conditions at the inlet and outlet of the absorber. Temperature, pressure and composition of flue gas are measured at the inlet and outlet while flow rate is measured only at the inlet of the absorber. There is no gas flow measurement at the outlet of the absorber due to the risk of flow meter being damaged due to potential of carrying over water droplets with the absorber exiting flue gas. Thus, outlet flow rate is calculated using mass balance across the absorber. Following equation (1) is used to calculate capture efficiency.

$$Capture\ Efficiency = (C_{in} - C_{out})*100/C_{in} \tag{1}$$

Where, C_{in} = Volume flow of CO_2 at NTP entering the absorber; C_{out} = Volume flow of CO_2 at NTP leaving the absorber

Specific Reboiler Duty (SRD) and the amount of CO_2 captured is calculated using energy used in the reboiler and the amount of captured CO_2 by using formulas previously given in Akram et al. (2016). Energy used in the reboiler is calculated using temperature of the Pressurised Hot Water (PHW) at the inlet and outlet of the reboiler and its flow rate by using Equation 2.

$$Q = M * C_p * (T_{in} - T_{out}) \tag{2}$$

Where, Q = energy consumption, kJ/h; M = mass flow rate of the pressurised hot water, kg/h; C_p = specific heat capacity of water, kJ/kg.K; T_{in} = inlet temperature of the pressurised hot water, °C; T_{out} = outlet temperature of the pressurised hot water, °C.

The amount of CO_2 captured was calculated using Equation 3.

$$M_{CO2} = (n_{CO2in} - n_{CO2out}) * MW_{CO2} \tag{3}$$

Where, M_{CO2}= mass of CO_2 captured, kg/h; n_{CO2in}= moles of CO_2 entering the absorber; n_{CO2out} = moles of CO_2 leaving the absorber; MW_{CO2} = molecular weight of CO_2.

The energy consumption per unit mass of CO_2 captured (MJ/kgCO_2) also referred to as SRD is calculated by using Equation 4 by putting values calculated in equations 2 & 3.

$$SRD = Q/M_{CO2} \tag{4}$$

Normalised Specific Reboiler Duty (NSRD) is calculated by dividing SRD for each test with the SRD for the test having the lowest SRD.

$$NSRD = SRD/(SRD)_{min} \qquad (5)$$

Degree of regeneration is calculated by Equation 6.

Degree of regeneration = (Rich loading – Lean loading)*100/Rich loading   (6)

3.4 Test Matrix:

Operational conditions (solvent flow rate, gas flow rate, stripper pressure and reboiler temperature) were varied to test the performance of the solvent. One parameter was varied at a time keeping rest of the operational conditions fixed. The range of variation in operational conditions is given in Table 2.

Table 2: Test matrix

Parameter	Unit	Operational Range
Solvent flow rate	kg/h	600-1200
Stripper pressure	barg	0.2-0.5
Reboiler temperature set point	⁰C	120-128
Flue gas flow	m³/h	160-200
Lean solvent temperature	⁰C	40
Flue gas temperature	⁰C	40

Some of the tests were carried out under controlled conditions for both 30% and 40% MEA for comparison. Results of all the tests carried out during this test campaign are described in the following section and sub-sections.

4. Results and discussion:

First set of experiments was carried out for evaluating the performance of 40% MEA, followed by the comparative tests with 30% MEA and the impact of variation is PHW. The results of the tests are presented in here.

4.1 Tests with 40% MEA

A number of experiments were carried out under varying operational conditions. The results of the experiments and interdependency of operational and calculated parameters are presented in this section.

4.1.1 Capture Efficiency and Specific Reboiler duty

Figure 2 plots the impact of changes in capture efficiency on the reboiler duty for all the 23 experiments carried out with 40% MEA during this test campaign. It can be seen from the plot that, as the capture efficiency increases, the NSRD first drops and then increases, passing through a minimum. The data is scattered due to different operational conditions used but the trend is clear. Minimum NSRD is observed to be at around 75-77% capture efficiency.

4.1.2 Variation of L/G ratio:

Figure 3 shows variation in capture efficiency and NSRD with changes in Liquid to Gas (L/G) ratio for 40% MEA. It can be observed from the figure that both capture efficiency and NSRD increase with increase in L/G ratio under the experimental conditions. The figure shows three sets of data. For the first two sets of data flue gas flow rate was fixed at (170 m³/h and 190 m³/h) while solvent flow rate was changed to vary L/G ratio. For the third set of data, solvent flow rate was fixed at 1200 kg/h while gas flow rate was changed to vary L/G ratio. For all the tests, reboiler temperature was fixed at 125 ⁰C while stripper pressure was fixed at 0.5 barg.
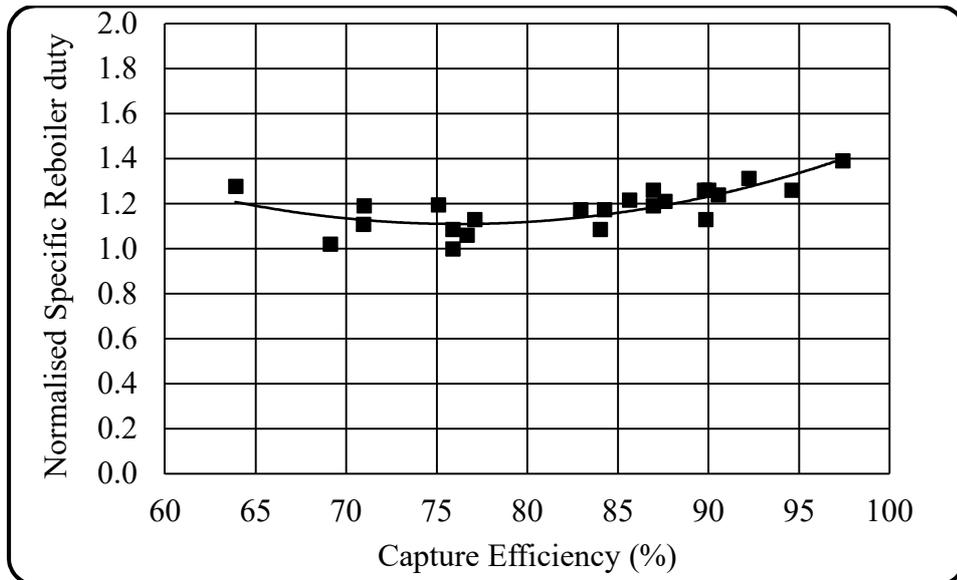
Figure 2: Capture efficiency vs. NSRD for 40% MEA

It can be observed from the figure that capture efficiency increases with increase in L/G ratio but at the cost of increase in NSRD. With increase in L/G ratio, more solvent is available to capture more CO_2 from the flue gas thus capture efficiency increases. On the other hand, at higher solvent flow rates, more energy is required to heat up the solvent which results in higher NSRD. It can be concluded from the data that both capture efficiency and NSRD increase with increase in L/G ratio regardless of the flow variation parameter i.e. solvent flow variation or gas flow variation. At lower L/G ratios, solvent has lesser capacity to absorb CO_2 resulting in increase in CO_2 loadings and thus drop in CO_2 capture efficiency. Although plot shows linear trend between L/G ratio and NSRD at the conditions tested, it is anticipated that at some lower L/G ratio NSRD will start to increase due to reduction in the amount of captured CO_2.
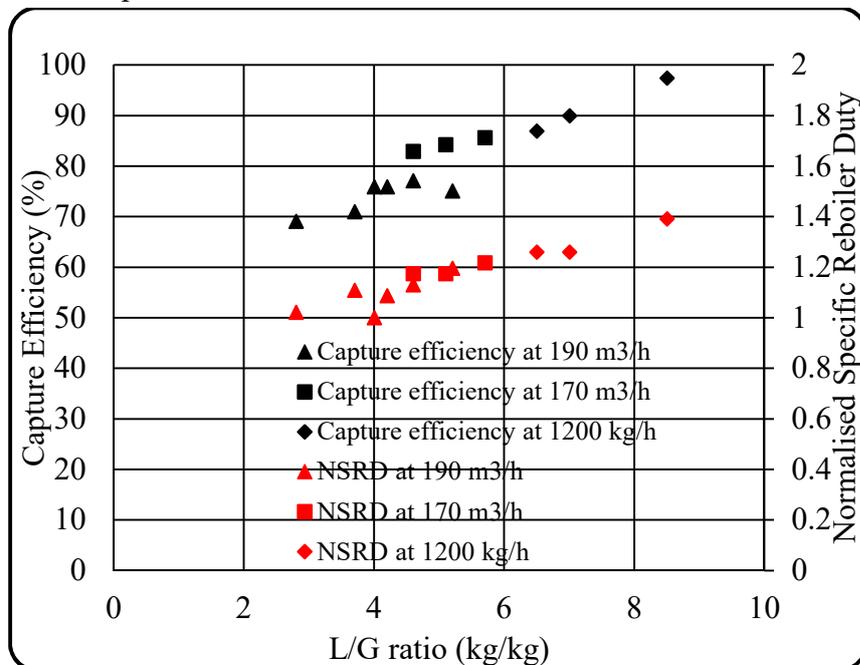

Figure 3: Capture efficiency and reboiler duty as a function of L/G ratio for 40% MEA

4.2 Comparison of 40% and 30% MEA:

A number of experiments were carried out with 30% to compare the results with 40% MEA. The results of the tests are given in Figure 4. The figure compares NSRD as a function of capture efficiency for 30% and 40% MEA. It can be observed from the figure that NSRD is lower for 40% MEA as compared to that for 30% MEA. Moreover, the difference in the specific reboiler duty between the two concentrations of MEA is higher at higher capture efficiency. This could be due to higher capacity of solvent in the case of 40% MEA which can absorb more CO_2. It is interesting to note that the lowest NSRD for both the concentrations of MEA under the tests conditions is around 75%.
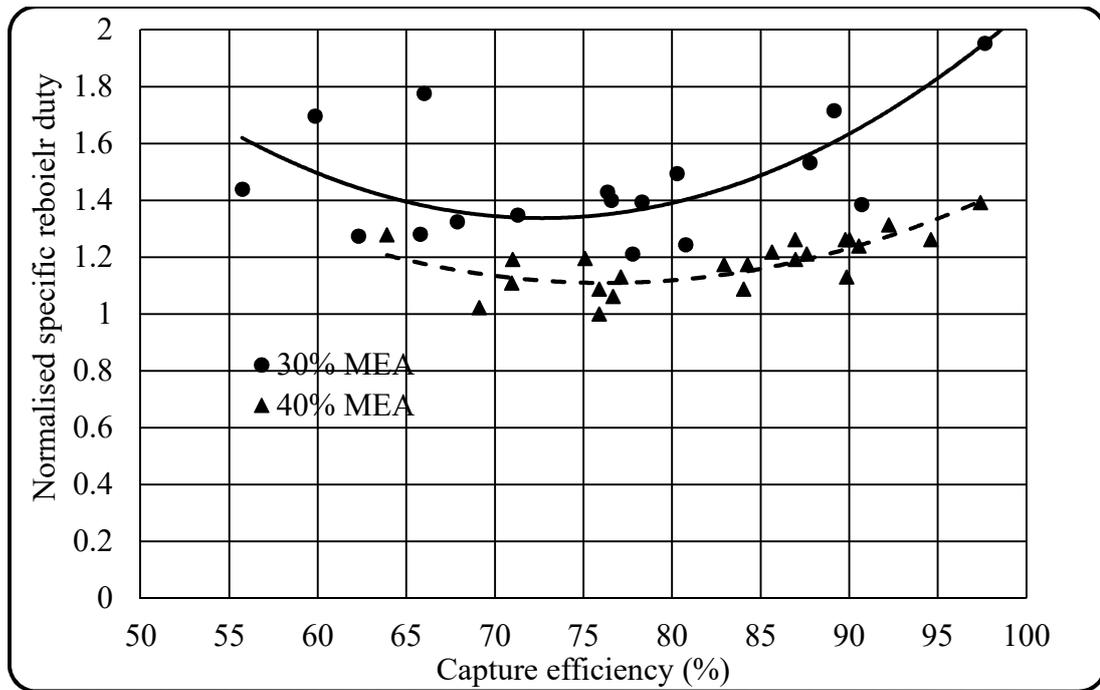

Figure 4: Comparison of capture efficiency vs NSRD for 30% and 40% MEA

Table 3: Operational conditions and results for comparative tests

Parameters	Test 1		Test 2		Test 3	
MEA (%)	40	30	40	30	40	30
Flue gas flow rate (m³/h)	197.9	197	161	158.5	162.4	158.5
CO_2 (%)	13.19	13.14	11.83	12.11	11.83	11.93
Solvent flow (kg/h)	897	890.7	709.7	677.9	705	684.6
Stripper pressure (barg)	0.2	0.2	0.2	0.2	0.4	0.4
Reboiler sump temperature (ºC)	115.4	114.4	115.39	115.62	117.32	117.84
Rich loading (mol/mol)	0.464		0.445	0.432	0.493	0.444
Lean loading (mol/mol)	0.236		0.224	0.209	0.407	0.272
Capture efficiency (%)	77.57	77.76	88.26	90.7	65.84	80.75
NSRD	1.06	1.21	1.21	1.38	1.28	1.24
Degree of regeneration (%)	49.1		49.7	51.6	17.4	38.7
CO_2 capture rate (kg/h)	33.7	33.5	28.3	29.3	21.6	26.1

In order to better understand the operational difference between 30% and 40% MEA, a set of experiments are carried out to compare 30% and 40% MEA under similar operational conditions. Operational conditions and the results for the three comparative tests are given in Table 3. Tests 1 and 2 were carried out at 0.2 barg stripper pressure while Test 3 was carried out at 0.4 barg stripper

pressure. Temperature of the PHW for all these tests was set at 120 ⁰C. For all the tests reported here, data is averaged over half an hour steady state test run.

It can be observed from the table that for Test 1, capture efficiency is the same (~78%) for both the concentrations of MEA while NSRD is 14% lower for 40% MEA. Similar behaviour is witnessed in Test 2 where capture efficiency for both of the MEA concentrations is very close (88-90%) while NSRD is 14% lower in the case of 40% MEA. Brigman et al. (2014) reported a 15% drop in reboiler duty when concentration of MEA was increased from 30% to 40%, with lean loading of 0.25 and 0.2-0.25 for 30% and 40% MEA, respectively, and capture efficiency of 85%. Therefore, it can be concluded that, under the conditions tested, 40% MEA has 14% less specific reboiler duty as compared to 30% MEA.

On the other hand, in the case of Test 3, NSRD for both the MEA concentrations is close but capture efficiency is 15% lower in the case of 40% MEA. The reason for this can be explained by comparing lean loading which is very high, above 0.4 mol/mol, in the case of 40% MEA resulting in solvent having very low capacity to absorb CO_2 in the absorber. This can be observed by looking at loadings and degree of regeneration values given in Table 3.

It can be observed from the table that lean loading increased from 0.224 to 0.407 mol/mol as stripper pressure was increased from 0.2 to 0.4 barg, all other parameters were kept the same. This indicates that there was not enough stripping at this point due to temperature limitations.

For tests 1 and 2, degree of regeneration is 49% while for test 3 it is dropped from 49% to just 17% with increase in stripper pressure from 0.2 to 0.4 barg. So, the poor performance of the stripper at these conditions is the reason for lower capture efficiency in the case of 40% MEA.

In order to further explain the above phenomenon, Figure 5 plots absorber temperature profile for the three comparative tests. It can be observed from the figure that the temperature profile for all the three test cases is similar but bulge temperature is different. The highest bulge temperature is recorded for Test 2 due to highest capture efficiency for this test case. The lowest bulge temperature is recorded in the case of Test 3. This is due to reduced absorption as a result of poor stripping as explained above.

The poor stripper performance resulted in higher lean loading which pushed the absorber conditions towards the equilibrium pinch. If lean loading is not sufficiently low, CO_2 equilibrium partial pressure in the lean stream entering the absorber is close to the partial pressure of CO_2 in the gas leaving at the top of the absorber. Under these conditions, mass transfer will drop in the upper section due to lower mass transfer driving force available. In order to avoid such a situation, lean loading should be dropped by increasing stripper temperature (Brigman et al. 2014). Therefore, in order to avoid the pinch performance of the absorber, PHW temperature was increased. Results of the experiments with elevated PHW temperature at 0.4 barg stripper pressure are explained in the next section.

The temperature at the bottom of the absorber in test 3 is lower than 40 ⁰C due to poor absorption rate resulting in lower temperatures throughout the absorption column. For the other two tests, due to higher absorption rate, temperatures along the column are higher so as the temperature at the bottom of the column.
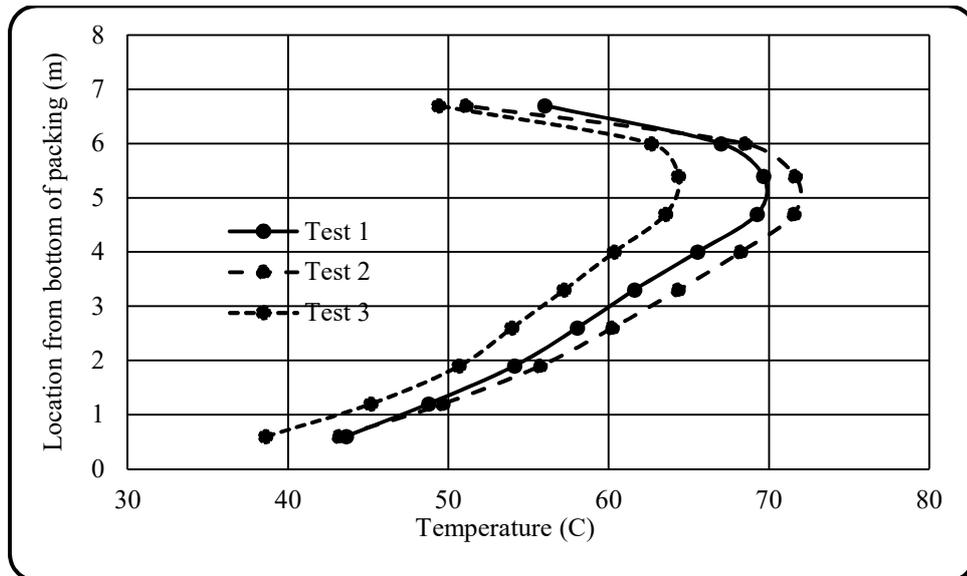
Figure 5: Absorber temperature profile for 40% MEA for the comparative tests

4.3 Variation of PHW Temperature:

In the previous section, results of 30% and 40% MEA are compared for three experiments under similar conditions. The results of Test 3 indicate that the process was equilibrium pinched due to high lean loading. In order to decrease lean loading and push the conditions away from pinch, PHW set point was increased gradually from 120 ºC to 125 ºC.

Figure 6 plots capture efficiency and NSRD for these tests. When the PHW temperature is increased, capture efficiency increased almost linearly under these conditions. With the increase in PHW temperature, stripper performs better and therefore, lean loading decreases. NSRD first dropped with increase in PHW temperature and then increased with further increase in PHW temperature. At the start, with increase in PHW temperature from 120 ºC to 122 ºC, NSRD dropped due to enhancement in stripping process. At these conditions, solvent absorbed more CO_2 in the absorber as a result of lower lean loading. With further increase in PHW temperature to 124 ºC, NSRD did not change although capture efficiency increased from 72% to 88%. At this point increase in energy consumption in the reboiler is balanced with the increased amount of absorbed CO_2. With further increase in PHW temperature to 125 ºC, NSRD started increasing due to relatively higher increase in the amount of energy used in the reboiler as compared to the increase in the amount of captured CO_2 under these conditions.

Figure 7, plot of loadings and degree of regeneration, shows that both lean and rich loadings drop with increase in PHW temperature. Lower the lean loading, higher the solvent capacity for more CO_2 absorption. The plot shows that lean loading drops sharply while rich loading drops less dramatically. In other words, the difference between lean and rich loadings, represented as degree of regeneration in Figure 7, increases as PHW temperature is increased which results in more CO_2 to be captured. The figure shows that as the PHW is increased from 120 ºC to 125 ºC, degree of regeneration increases from 17% to around 65%. It is interesting to note that values of degree of regeneration, capture efficiency and NSRD achieved for stripper pressure of 0.2 barg at PHW temperature of 120 ºC (test 2 in Table 3) are similar to those achieved at stripper pressure of 0.4 barg and PHW temperature of 124 ºC (Figures 6&7) indicating that capture process for 40% MEA under these conditions is similar.

In order to better understand the phenomenon, a RTD (Resistance Temperature Detector) was inserted into the reboiler to measure the actual solvent temperature in the reboiler sump. The reboiler sump temperature is plotted against PHW temperature in Figure 8. PHW is the water providing heat to the reboiler. PHW temperature is the boiler set point temperature while reboiler temperature is the actual solvent temperature as measured by inserting RTD right into the middle of the reboiler. As can be observed from the figure that, at the start, the reboiler sump temperature increase with increase in

PHW temperature. However, when PHW temperature increased from 124 ⁰C to 125 ⁰C, reboiler sump temperature did not increase indicating that most of the excessive energy introduced by increase in PHW temperature may have been used in the production of steam under these conditions which has resulted in increase in NSRD. This indicates that the optimum PHW temperature for 40% MEA under these operational conditions in this plant is around 124 ⁰C.
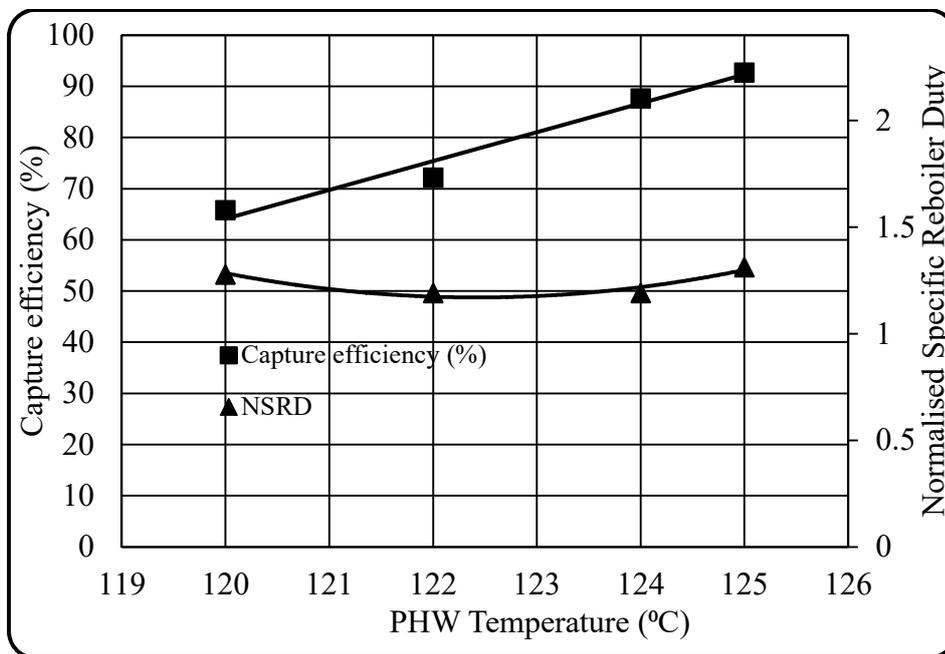

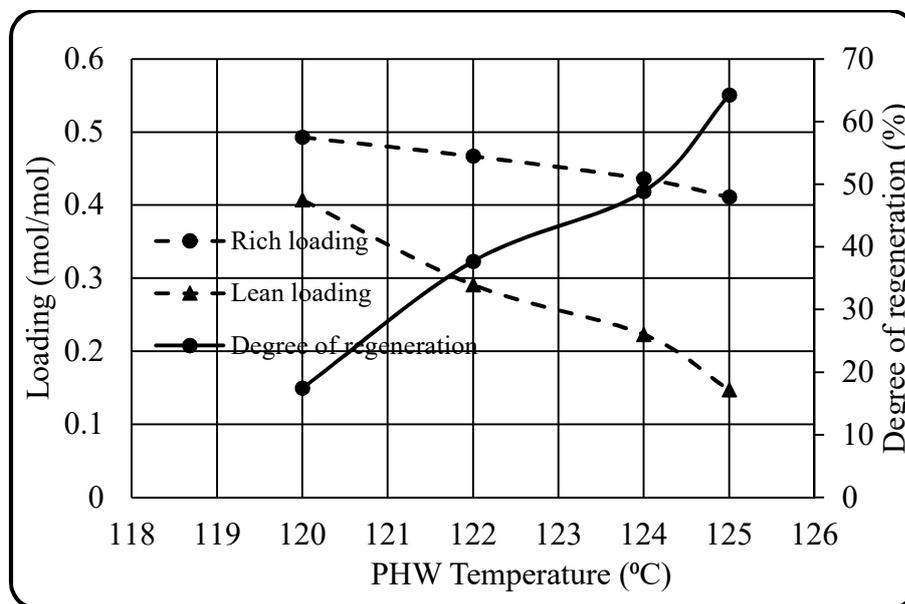Figure 6: Impact of regeneration temperature on capture efficiency and NSRD

Figure 7: Comparison of rich and lean loadings and degree of regeneration for PHW temperature variation
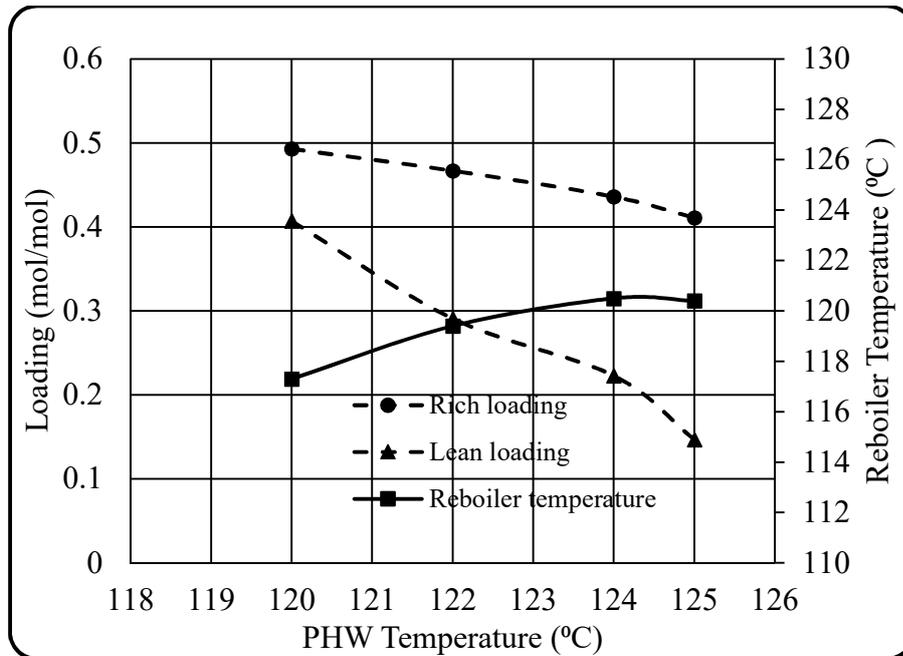
Figure 8: Comparison of rich and lean loadings and reboiler sump temperature for PHW
temperature variation

In order to reduce the production of steam, pressure was increased to 0.5 barg at reboiler temperature of 125 ºC. The results are plotted in Figure 9. With further increase in reboiler temperature, NSRD dropped slightly and stayed stable. Capture efficiency also dropped slightly due to reduced stripping at increased pressure at the same reboiler temperature, however, capture efficiency kept on increasing with increase in reboiler temperature due to increased amount of stripping at higher temperature at constant stripper pressure of 0.5 barg. It is clear from the figure that capture efficiency increases linearly with increase in reboiler temperature at both pressures under the test conditions.

It is also interesting to observe in the plot that similar capture efficiency can be obtained at 0.5 barg as at 0.4 barg but at slightly higher temperature with 0.5 barg.   This highlights that capture efficiency can be maintained at elevated stripper pressure by the right combination of temperature and pressure. It is anticipated that compression costs can be reduced if performance of the capture plant can be maintained at elevated stripper pressures.
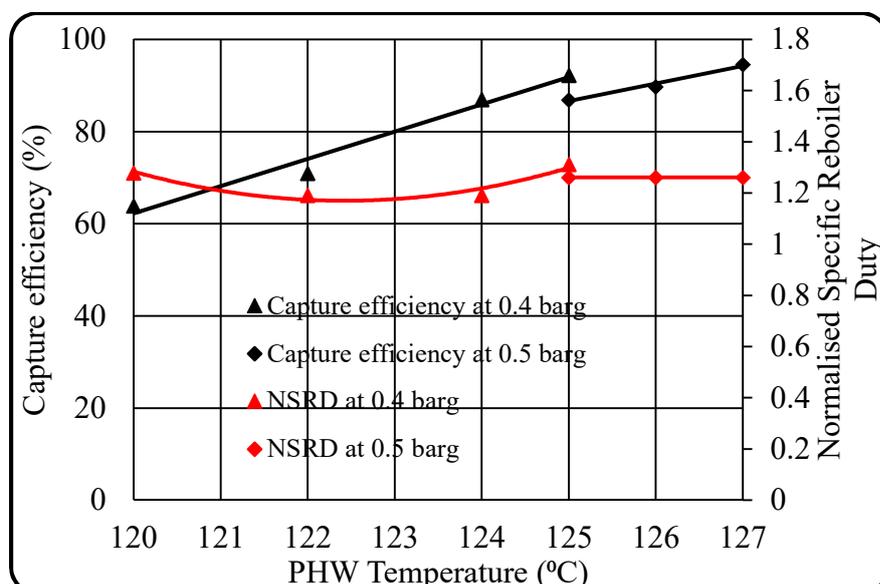

Figure 9: Capture efficiency and NSRD as a function of reboiler temperature

4.4 Emissions:

Gas composition was analysed at absorber inlet, absorber outlet and water wash outlet periodically. Figures 10 and 11 plot concentration of MEA and Ammonia recorded on a typical test day by FTIR for 40% and 30% MEA concentrations, respectively. In order to distinguish between different measurements points only part of the data is plotted. The measurement intervals at absorber inlet, absorber outlet and water wash outlet are marked in the graphs. In both cases, peak values for MEA in the plot are the measurements at the outlet of the absorber while the lower values are for water wash outlet. As can be seen from the plot in Figure 10, concentration of MEA at the outlet of the absorber for 40% MEA was quite high, mostly between 100 and 150 ppm while at the outlet of the water wash was below 10 ppm indicating that simple water wash was quite effective in removing most of the MEA from the flue gas. Concentration of ammonia was always measured to be around 50 ppm and was similar at the inlet and outlet of the water wash. Fractionally higher values noted at the water wash outlet are due to changes in gas composition due to water condensation.
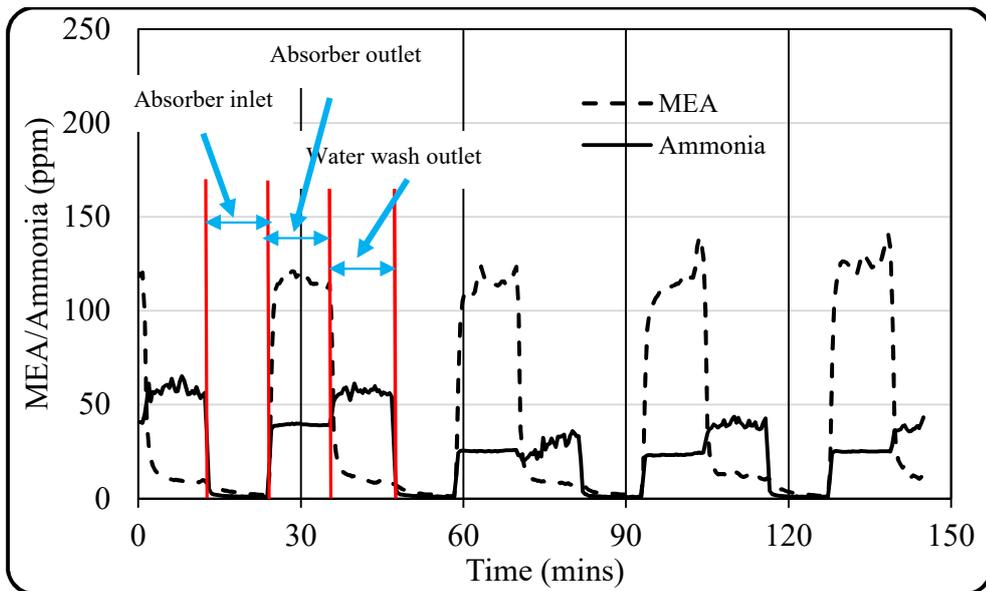

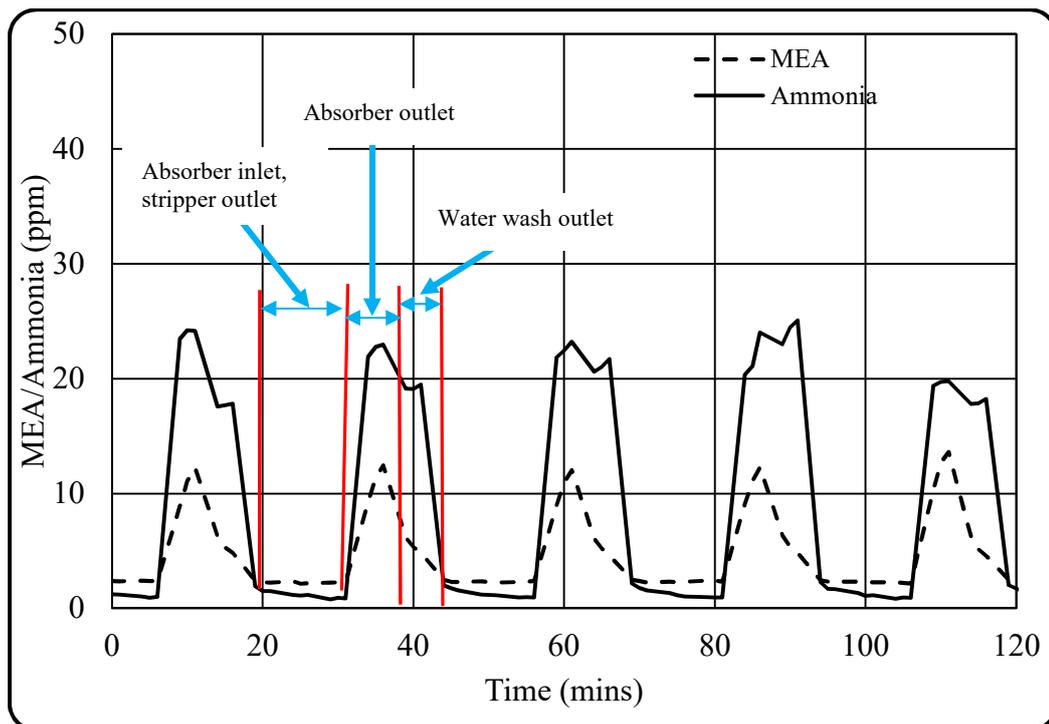Figure 10: Emissions of MEA and Ammonia for 40% MEA (sample data)

Figure 11: Emissions of MEA and Ammonia for 30% MEA (sample data)

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In the case of 30% MEA, concentration of MEA in the flue gas was always below 15 ppm with slightly lower values measured at the outlet of the water wash as compared to the inlet. This suggests that higher concentrations of MEA can result in higher carryover of MEA from absorber but simple water wash is good enough to tackle the issue. Concentration of ammonia in the case of 30% MEA was always around 25 ppm or below indicating that higher concentration of MEA can degrade at faster rate. In order to make full understanding of the phenomenon, further studies are being carried out at the PACT plant which will be published in the public domain very soon.

Morken et al. (2014) reported emissions of MEA at the outlet of water wash of around 20 ppm and 100 ppm for 30% and 40% MEA, respectively. Similar data is reported by Rochelle et al. (2011) who reported that 30% MEA at a CO_2 loading of 0.45 has a MEA volatility of 30 ppm.

This indicates that higher concentrations of MEA have higher tendency for carry over with the flue gas. Water wash section has to be designed to be very efficient in removing MEA from exiting flue gases if higher concentrations of MEA are to be used.

4.5 Corrosion:

Baseline solvent used in the industry is 30% MEA and all other new solvents and concentrations are compared to the performance of the baseline standard. Higher concentrations of MEA can result in higher degradation rates (Morken et al. 2014) and possibly higher material corrosion rates (Brigman et al 2014).

In order to compare the rate of degradation of 30% and 40% MEA, samples were collected every two days and were analysed for Iron (Fe) by HACH Pocket Colorimeter II. The results are shown in Figure 12. After 8 days of operation with 40% MEA, Fe content was measured to be 5.5 mg/L while that with 30% MEA was 3.2 mg/L. This indicates that 40% MEA has higher corrosion rate as compared to 30% MEA. However, Brigman et al (2014) reported that they have not measured any significant increase in iron concentrations when using 40% MEA in a stainless steel absorber but their absorber was lined with polypropylene. Further pilot scale studies are being carried out at PACT pilot plant to quantify the impact of MEA strength on the plant corrosion and resulting degradation. However, the results presented in Figure 12 show a clear trend that higher concentrations of MEA can result in higher degree of solvent degradation.
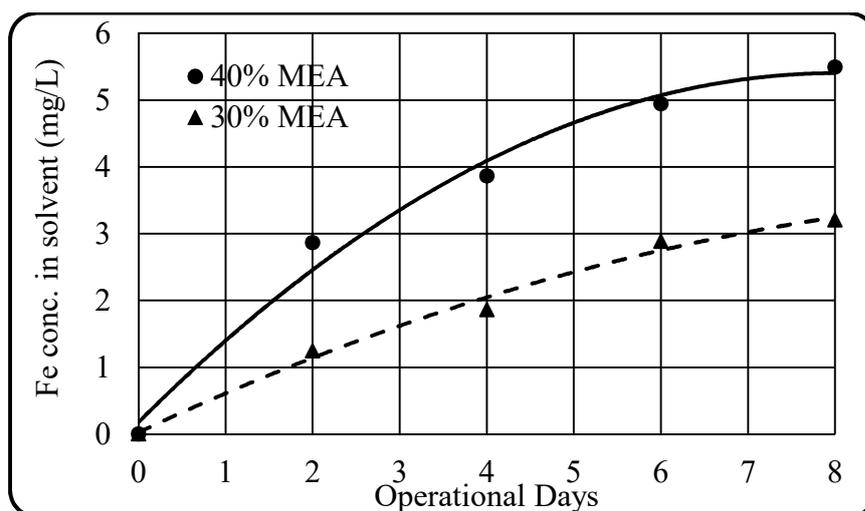

Figure 12: Corrosion measurement with 40% MEA

5. Conclusions:

Experiments are carried out with 30% and 40% MEA concentrations to compare performance of the two solvent concentration at PACT CO_2 capture pilot plant. The results have shown that higher concentrations of MEA give better specific reboiler duty but result in increase in rate of corrosion

and degradation. It has been observed that 40% MEA offers around 14% lower specific reboiler duty than 30% MEA under identical operational conditions.

The process is very temperature sensitive and slight change in reboiler temperature can have a big impact on the performance. It is observed during these tests that it may be possible to achieve similar capture efficiency and specific reboiler duty for different reboiler temperatures with right combination of stripper pressure and reboiler temperature. In order to reduce burden on the downstream compression process it is desirable to operate the CO_2 capture strippers at elevated pressures. The results here suggest that it is possible to achieve similar performance at elevated pressures. However, higher operating temperature at higher pressures will have impact on solvent degradation and corrosion.

Although 40% MEA has better energy and capture performance, degradation and corrosion rates are higher with 40% MEA as compared to those with 30% MEA. In order to control degradation and corrosion rate it may be required to introduce corrosion inhibitors into the system. Further research is required to compare the impact of corrosion inhibitors on controlling the corrosion rate in different MEA concentrations.

Solvent emissions at the absorber outlet are higher for 40% MEA than 30% MEA indicating that concentrated MEA has higher tendency for carry over. During these trials it was observed that simple water wash was able to lower the MEA emissions to below 10 ppm. However, in order to control the emissions to the atmosphere, if higher concentrations of MEA are to be used, it may be required to use two stage water wash, acid wash in the first stage and simple water wash in the 2nd stage.

6. Acknowledgement:

7. References:

1. Abu-Zahra M.R.M., Niederer J.P.M., Feron P.H.M., Versteeg G.F., CO_2 capture from power plants: Part 2. A parametric study of the economical performance based on mono-ethanolamine, International Journal of Greenhouse Gas Control, Volume 1, Issue 1, April 2007, Pages 135-142.
2. Abu-Zahra M.R.M., Schneiders L.H.J., Niederer J.P.M., Feron P.H.M., Versteeg G.F., CO_2 capture from power plants: Part I. A parametric study of the technical performance based on monoethanolamine, International Journal of Greenhouse Gas Control, Volume 1, Issue 1, April 2007, Pages 37-46.
3. Ahn, H., Luberti, M., Liu, Z., Brandani, S., 2013. "Process configuration studies of the amine capture process for coal-fired power plants." Int. J. Greenh. Gas Control 16: 29–40.
4. Akram M., Ali U., Best T., Blakey S., Finney, K.N., Pourkashanian M., Performance evaluation of PACT Pilot-plant for CO_2 capture from gas turbines with Exhaust Gas Recycle, International Journal of Greenhouse Gas Control, Volume 47, April 2016, Pages 137-150.
5. Amrollahi, Z., Ertesvag, I.S., Bolland, A., 2011. "Optimized process configurations of post-combustion CO_2 capture for natural-gas-fired power plant-exergy analysis." Int. J. Greenh. Gas Control 5: 1393–1405.
6. Aronu, U.E., Svendsen H.F., and Hoff, K.A., Investigation of amine amino acid salts for carbon dioxide absorption. International Journal of Greenhouse Gas Control, 2010. 4: 771-775.
7. Brigman N., Shah M.I., Falk-Pedersen O., Results of amine plant operations from 30 wt% and 40 wt% aqueous MEA testing at the CO_2 Technology Centre Mongstad, Energy Procedia 63 (2014) 6012 – 6022.

8. CCUS Cost Challenge Taskforce, "Delivering Clean Growth: CCUS Cost Challenge Taskforce Report", July 2018.

9. Cheng H., Lai C., Tan C. 2013. "Thermal regeneration of alkanolamine solutions in a rotating packed bed." Int J Greenhouse Gas Control 16: 206–16.

10. Clean Growth: The UK Carbon Capture Usage and Storage deployment pathway-An action Plan, 2018.

11. de Cazenove T., Bouma RHB., Goetheer ELV., van Os PJ. and Hamborg ES. 2016. "Aerosol measurement technique: Demonstration at CO2 Technology Centre Mongstad." Energy Proc 86:160–170.

12. Delfort B., Carrette P.L, Bonnard L., MEA 40% with improved oxidative stability for CO_2 capture in post-combustion, Energy Procedia, Volume 4, 2011, Pages 9-14.

13. Diego ME., Akram M., Bellas JM., Finney KN. and Pourkashanian M. 2017. "Making gas-CCS a commercial reality: The challenges of scaling up." Greenhouse Gases: Science and Technology 7 (5): 778-807.

14. Hakka, L., Cansolv Technologies Inc. 2007, private communication.

15. Herraiz L., 2016. "Selective Exhaust Gas Recirculation in Combined Cycle Gas Turbine Power Plants with Post-combustion Carbon Capture." PhD Thesis. University of Edinburgh, Scotland, UK.

16. Jassim, M.S., Rochelle G., Eimer D. and Ramshaw C., Carbon dioxide absorption and desorption in aqueous monoethanolamine solutions in a rotating packed bed. Industrial & Engineering Chemistry Research, 2007. 46: 2823-2833.

17. Kang J.L., Wong D.S.H., Jang S.S., and Tan C.S., A comparison between packed beds and rotating packed beds for CO_2 capture using monoethanolamine and dilute aqueous ammonia solutions. International Journal of Greenhouse Gas Control, 2016. 46: 228-239.

18. Kim, Y.E., Lim, J.A., Jeong, S.K., Yoon, Y.I., Bae, S.T., Nam, S.C., 2013. "Comparison of carbon dioxide absorption in aqueous MEA, DEA, TEA and AMP solutions." Bulletin of Korean Chemical Society 34: 783–787.

19. Kumar, S., Cho J.H. and Moon I., Ionic liquid-amine blends and CO_2 BOLs: "Prospective solvents for natural gas sweetening and CO2 capture technology – A review". International Journal of Greenhouse Gas Control, 2014. 20: 87-116.

20. Le Moullec Y., Neveux T., Al Azki A., Chikukwa A. and Hoff KA., 2014. "Process modifications for solvent-based post-combustion CO_2 capture." Int J Greenhouse Gas Control 31: 96–112.

21. Li K., Cousins A., Yu H., Feron P., Tade M., Luo W. & Chen J., 2016., "Systematic study of aqueous monoethanolamine-based CO_2 capture process: model development and process improvement", Energy Science and Engineering, 4(1): 23–39.

22. Madan, T., Sachde, D., Lin, Y.-J., Frailie, P., Rochelle, G.T., 2013. "Improved process configurations for amine scrubbing." In: 7th Trondheim Conference on CO_2 Capture, Transport and Storage, Trondheim (NO), 4–6 June.

23. Mejdell T. Vassbotn T. Juliussen O., et al. 2011. "Novel full height pilot plant for solvent development and model validation." Energy Procedia 4:1753–1760.

24. Merkel TC. Wei X., He Z., White LS. Wijmans JG. Baker RW. 2013. "Selective exhaust gas recycle with membranes for CO_2 capture from natural gas combined cycle power plants." Ind Eng Chem Res 52:1150–1159 (2013).

25. Morken AK. et al., Emission results of amine plant operations from MEA testing at the CO_2 Technology Centre Mongstad, Energy Procedia 63 (2014) 6023 – 6038.

26. Notz R, Mangalapally HP. and Hasse H. 2012. "Post combustion CO_2 capture by reactive absorption: Pilot plant description and results of systematic studies with MEA." Int J Greenhouse Gas Control 6: 84–112.

27. Oh SY. Yun S. Kim JK. 2018. "Process integration and design for maximizing energy efficiency of a coal fired power plant integrated with amine-based CO_2 capture process." Applied Energy 216:311–322.

28. Rochelle G., Chen E., Freeman S., Wagener DV., Xu Q., Voice A., Aqueous piperazine as a the new standard for CO_2 capture technology, Chemical Engineering Journal, 171 (2011) 725 – 733.

29. Wang M., Joel AS., Ramshaw C., Eimer D., Musa NM. 2015. "Process intensification for post-combustion CO2 capture with chemical absorption: A critical review." Applied Energy 158: 275–291.

30. Yang Q., Puxty G., James S., Bown M., Feron P., Conway W.2016. "Toward Intelligent CO_2 Capture Solvent Design through Experimental Solvent Development and Amine Synthesis." Energy Fuels 30: 7503−7510.

31. Yuan Y. and Rochelle G.T., CO_2 absorption rate in semi-aqueous monoethanolamine. Chemical Engineering Science, 2018. 182: 56-66.