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SMART - Solvent Management At Reduced Throughput – a prototype demonstration plant

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Abstract

It is important to verify in advance by pilot testing that amine solvents used for a post-combustion capture application can be maintained in good condition indefinitely and that emissions can be kept at acceptable levels as the solvent ages. UK permitting guidance now recommends that this is done on a slipstream of the actual flue gas for extended periods (e.g. 8000 hrs or longer) with realistic process conditions, including full reclaiming and other solvent management techniques. This would ideally be done in a large pilot unit (e.g. 0.4 m diameter columns, ~ 10tCO2/day) to give realistic column packing performance, but equipment costs are of the order £10M or more. Also, with many potential UK sites with different, specific flue gas characteristics, and a number of potential solvents, the UK CCS sector would require multiple parallel tests to take place, and sharing of a small number of large facilities would not be feasible. And, of course, if different approaches are to be tested and each test takes many months that it is virtually essential to be able to run multiple tests in parallel.

To this direction, this paper describes fundamental laboratory research on the interaction between reclaiming and full PCC cycle operation. The equipment being developed at the University of Sheffield is representative, but small scale, of testing of amine post-combustion capture applications, and includes on-site use by UK industry. Features in the design include: integrated thermal reclaimer, externally heated reboiler with low skin temperatures to avoid thermal degradation hot spots, acid wash for emission control, absorber with low-liquid loading packing to give a 'short, fat' absorber (for easier transport and location on site) that can still achieve representative solvent loadings. The immediate objective is to build a laboratory version of the SMART equipment that will determine the design details for future units to be used on commercial sites. Building a prototype unit for use in laboratory setting very importantly allows modifications to be undertaken quickly in service – this is a novel system, and it will no doubt be necessary to adjust aspects of the equipment to achieve the desired performance. The prototype will also not be designed for unattended or fully automatic operation, which very significantly reduces costs and complexity (crucially, given the limited funds available) – but adding this capability in 'production' units is a routine matter once the fluid flow and heat and mass transfer aspects of the design and operation have been researched and defined. It will, however, be possible to run the prototype unit with minimal or no manual intervention for long periods, allowing extended trials as 'background activity' in the laboratory.

Reclaimer trials and full SMART operation is able to start with a charge of aged MEA from the TERC 1 tCO_2 day^{-1} unit to give accelerated results. Operation in the laboratory environment will also allow novel analytical techniques,

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such as online solvent concentration and loading measurements to be refined under realistic circulating-solvent conditions.

Anticipated CO_2 capture rates are 20-100 kg CO_2/day, depending on flue gas. Novel features include:

- Integrated reclaiming – based on previous UKCCSRC/industry-funded PCC-CARER research.
- Realistic desorber/reboiler pressures and hence temperatures –essential to match thermal degradation effects – the pressurised desorber also allows direct feed to a ‘short, fat’ absorber without pumping.
- Realistic solvent loadings – the absorber will use low-liquid-loading gauze packing to give a relatively ‘short, fat’ column, making the unit much easier to locate and move.
- Realistic solvent cycling rate – the number of solvent cycles per day and residence times at different stages in the cycle will also be reproduced, based on detailed commercial plant data available.
- Realistic flue gas conditioning and absorber exit emission countermeasures –water and acid washes will be included as separate columns, giving flexibility in reproducing different equipment options.

The reclaimer is initially being tested in a standalone mode before integration with the rest of the SMART rig.

Keywords: amine post-combustion capture; solvent management; reclaiming

1. Introduction

Removal of CO_2 emissions from post-combustion carbon capture (PCC) plants must be tackled to achieve global long term climate targets. It is important that amine solvents used in PCC applications can be maintained in good condition indefinitely and that, as the solvent ages, emissions are kept at acceptable levels.

UK permitting guidance recommends that pilot testing be done on a slipstream of flue gas for extended periods (8000 hours or longer) under realistic process conditions, including full reclaiming and other solvent management techniques [1]. However, recent experience by the authors suggests that the cost for sufficiently long testing periods makes testing difficult, together with the quantity of potential UK sites that have specific flue gas characteristics and the number of potential solvents, causing the UK CCS sector to require multiple parallel tests to take place simultaneously. And, of course, if different approaches are to be tested and each test takes many months then it is virtually essential to be able to run multiple tests in parallel. The equipment being developed at the University of Sheffield for representative, but small scale, testing of amine post-combustion capture applications will enable these required multiple parallel pilot tests to occur, provided adequate reclaiming/cleaning equipment can be installed – the authors are only aware of such a system, using thermal reclaiming, for MEA [2].

Amine solvents used in the removal of CO_2 from flue gas streams are subject to thermal and oxidative degradation, causing problems such as corrosion, reduction in CO_2 capture efficiency, increased emissions to atmosphere and increases to operational costs [3]. The rate of degradation of solvents is often unreported for proprietary solvents, and where reported, pilot scale campaigns are too brief to quantify longer term degradation rates. However, maintaining solvent health for sustained operation is achievable by the continuous removal of degradation products from the circulating solvent at the same rate they are formed. In pilot scale campaigns, thermal reclaiming of the solvent tends to only happen between runs, or else when certain conditions are exceeded, such as heat stable salt (HSS) concentration [4,5]. However, as pilot scale campaigns are not sufficiently long to quantify long term degradation rates, long term pilot testing (>8000 hours) should be undertaken using *continuous* thermal, other reclaiming methods, or other comprehensive and effective solvent management methods to properly demonstrate the long-term effects of reclaiming on the solvents lifecycle.

Two common non-proprietary solvents used for capture studies are monoethanolamine (MEA) and CESAR-1, an aqueous blend of 2-amino-2-methyl-1-propanol (AMP) and piperazine. Both solvents have been thermally reclaimed in batch mode, although the limited published data on thermal reclaimability of CESAR-1 to date suggests incomplete removal of impurities [6,7]. However, reclaimability of MEA been readily documented, and tests from the Technology Centre Mongstad (TCM) indicated that thermal reclaiming of MEA demonstrates a high rejection of impurities [4]. Thermal reclaiming involves the separation of species based on their varying volatilities; hence, continuous thermal reclaiming of blended solvents may not be effective. For that reason, the Solvent Management at Reduced Throughput (SMART) project will focus on using MEA as a solvent, as it is able to be continuously thermally reclaimed.

The SMART rig is a lab-scale capture unit, built and operated to enable and achieve low lean loadings of CO_2 after the regeneration step (desorber) and high loadings of CO_2 in the absorption step (absorber), as suggested by a study conducted by Michailos and Gibbins [8] using an Aspen Plus CCSI MEA Steady State Model for high capture rates. The results from this study show that if the absorber has sufficient packing height and the desorber is capable of operating above 2 bar, then it is feasible for the plant to operate efficiently at higher capture levels, with only minimal adjustments to the plant operation. The desorber pressure required to reach the lean loadings necessary for effective high capture levels without excessive energy requirements is stated to be 2.4 bar.

However, further work [9] explores the extent of increased degradation of MEA when desorber conditions are modified to increase the pressure and temperature to the optimal conditions described [8] to achieve a lower lean loading of CO_2 in MEA, to capture 100% of additional CO_2 in flue gas capture. Carbamate polymerization, or thermal degradation, has the highest rate in the reboiler, as well as in the descending solvent in the desorber and the solvent pre-heating stream.

The rate of degradation is a function of CO_2 loading and temperature, and a recent process modelling study [9] suggested that under the conditions required for a lower lean loading of 0.1 $molCO_2\ molMEA^{-1}$ (i.e. higher pressure, and so temperature requirement) the rate of thermal degradation increases by 300% - 400%, compared to a lean loading of 0.2 $molCO_2\ molMEA^{-1}$ due to the effects of increased regeneration temperature outstripping the limiting effect of carbamate concentration in the lean solvent. It was found that the rate of total thermal and oxidative degradation increases by 24% - 138% as a result of increased reboiler temperatures and additional absorber residence times, depending on the desired capture fraction of up 100% additional CO_2 capture [9]. However, this study was limited as it considered only completely clean solvent, with no work on the catalytic effects of degradation products on the degradation rate of the solvent, which is likely to significantly increase over time. As such, long term testing of the desorber conditions described should be undertaken using continuous thermal reclaiming to manage the solvent to offset increases in the rate of thermal degradation resulting from an increased reboiler temperature to capture a higher fraction of CO_2 from flue gases. All the above will be tested in the SMART rig that is being constructed at the University of Sheffield, which aims to produce realistic amounts and rates of degradation products and degradation product formation and remove them using continuous thermal reclaiming at a rate of no more than 1 inventory per week, cycling the MEA solvent between two approximate loadings of 0.1 $molCO_2\ molMEA^{-1}$ and 0.45 $molCO_2\ molMEA^{-1}$. Regeneration conditions will be realistic, operating at pressures up to 2.4 bar(a) and temperatures of up to 135 °C, with, as far as possible, realistic temperatures in the absorber and realistic continuous solvent reclaiming. A simplified flow diagram is shown in Figure 1.

Degradation products will be analysed using gas chromatography and titration methods. Whilst it is not possible to analyse for every degradation product, the objective is to identify if any product is accumulating despite continuous thermal reclaiming.

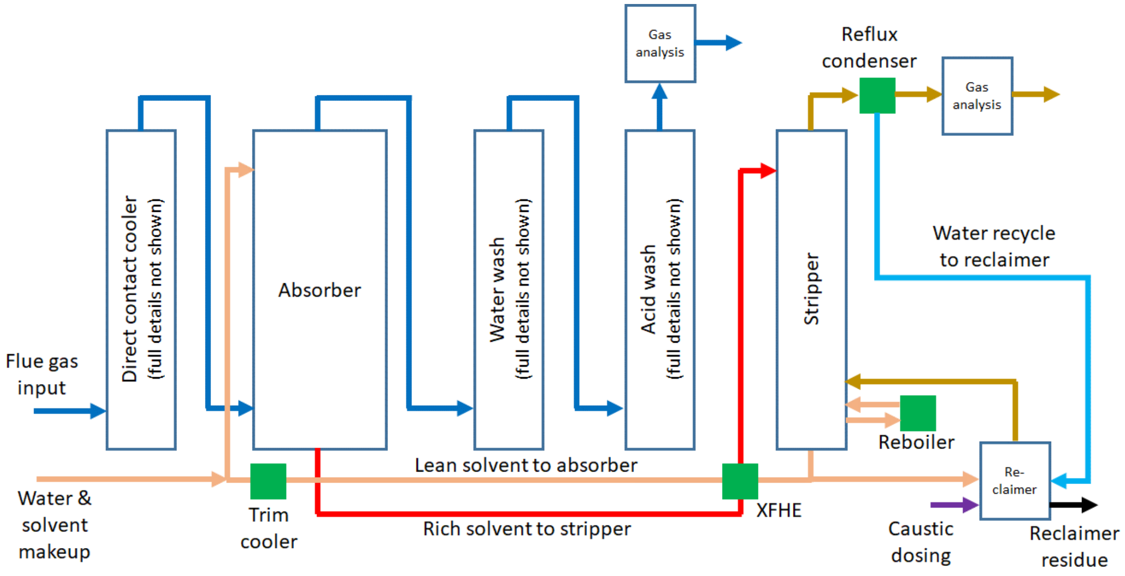

Figure 1: - Simplified SMART test prototype flow diagram (pumps, blowers, cooling water flows, temperature control details omitted)

2. Apparatus Design and Methods

2.1. Construction Materials

Degradation of MEA solvent has been shown in increase when in contact with transition metals such as iron [10] , copper [11], chromium [12], manganese [12] and cobalt [12]. Zinc has also been linked with thermal decomposition of MEA at low temperatures [13]. Contact of the solvent with these metals catalyses oxidative degradation of the solvent, resulting in the production of ammonia, organic acids, and eventually HSSs [10]. As a result, the PCC rig has been constructed using stainless steel (mostly 316) and PTFE on high temperature areas, and other plastics such as MDPE in lower temperature areas, with particular care to avoid contact of any wetted areas with other metals.

2.2. Desorber

The desorber has a packing height of 1m of random packed stainless steel 3/8” Pall ring packing in a column diameter of 4”. The reboiler has two immersion heaters with a combined variable power input of up to 6 kW, giving an estimated maximum CO_2 removal rate of 6 kg h^{-1}. The reboiler has a volume of around 7 litres, and a residence time of 2 - 5 minutes. Maintaining a low desorber sump residence time is critical as the rate of thermal degradation increases linearly with solvent residence time [9], which becomes more significant at higher temperatures.

The desorber is where the solvent regeneration stage occurs, designed to provide a lean loading of 0.12 $molCO_2/molMEA$ in line with the conditions described [8] aiming to achieve a temperature and pressure of ~130 °C and 2.4 bara, respectively. Increasing the pressure above atmospheric operation will ensure the target lean loading without subjecting the solvent to temperatures above 135 °C, where thermal degradations rates are expected to significantly increase [14].

2.3. Absorber

A key part of the current design is ‘short, fat’ absorbers using gauze packing to limit the height of the unit in the laboratory. These are also easier to install and operate in any location. High capture rates in the absorber are desirable but not essential. It is also possible that the absorber inlet CO_2 may need to be elevated to unrealistically high levels achieve the rich loading with the current absorber. The absorber has a diameter of 300mm and uses state of the art Sulzer BX packing, shown in Figure 2, which was recommended for use for the design solvent flux of 1.14 – 4.68 $m^3\ m^{-2}\ h^{-1}$, to achieve a high rich loading for a low packing height.

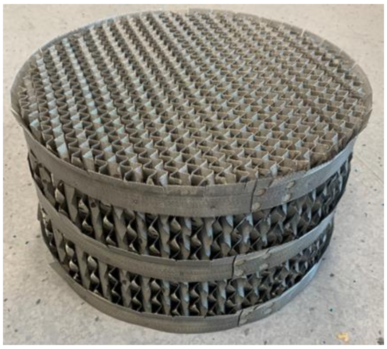

Figure 2: - BX Packing used in absorption column

2.4. Liquid-release system

A standard SMART unit has been developed to release liquid while maintaining pressure. This is achieved through a high- and low-level float switch, which triggers a solenoid valve to release liquid at the high level and closes at the low level. The volume released between the high and low trigger is approximately 350 ml, enabling liquid discharge whilst maintaining pressure upstream of the float switches. These units can be designed to operate at up to 180°C.

Figure 3: - Liquid release system internals (left), and external (right)

2.5. Removal of more volatile degradation products

Thermal reclaiming relies on the difference in volatility of MEA and degradation products for effective separation from products such as non-volatile HSSs. As more-volatile degradation products than MEA, such as acetaldehyde and acetone, will be carried over from the thermal reclaiming of MEA, these volatiles must be removed from elsewhere in the system to prevent accumulation of these compounds and their further degradative reactions. As such, an offtake in the water condensate stream after the desorber is included in the design, which can be operated in a feed-and-bleed, removing volatile-rich water, and making up the water balance through the water wash.

2.6. Analysis methods

Various analytical techniques are used to monitor amine concentration and CO_2 loading in commercial settings, including titration, gas chromatography, spectroscopy, and Total Inorganic Carbon (TIC) analysis. However, these techniques present trade-offs between accuracy, cost, robustness and speed. To address this challenge, the SMART rig incorporates advanced real-time solvent monitoring techniques, supported by data-driven analysis. These techniques, being developed at the University of Sheffield, are fast, cost-effective, and adaptable, enabling precise monitoring, better process control, optimization, and more responsive operational adjustments.

COMCAT (Control, Optimization, and Measurement in CO_2 Absorber Transients) continuously infers online amine concentration and CO_2 loading based on solvent density and viscosity measurements from industrial-grade instruments. The process leverages on offline calibration data derived from well-established correlations to ensure accuracy. COMCAT is designed to monitor rich solvents exiting the absorber.

HAPTICS (High Accuracy Pressure-Temperature Inferential Control in Stripping) infers lean CO_2 loading and solvent concentration in the desorber reboiler/sump by measuring solvent density, reboiler pressure, and reboiler temperature. It capitalizes on the equilibrium state between vapor and solvent to estimate real-time solvent conditions in lean solvent streams.

Analysis of solvent performance will also be determined using gas chromatography with a flame ionization detector, enabling qualitative and quantitative analysis of degradation products when compared to standard samples. The gas chromatograph uses a CP-Sil 8 CB column of length 30 m, internal diameter 0.32 mm and film thickness 1.0µm.

2.7. Reclaimer Design

The reclaimer design can be operated as integrated with the rest of the SMART system, or standalone for testing reclaiming conditions separately.

The reclaimer is heated using a 1.1 kW stirring hotplate, and a 1.75 kW heating tape, and operated under the same pressure as the desorber, enabling energy recovery by returning reclaimed solvent back into the desorber, above the reboiler and below the packing, with the aim to reclaim 1% of the solvent inventory every hour. The percentage of the inventory to be reclaimed is determined through power input, as the reboiler maintains a constant volume of around 2 litres and is fed from an overflow from the reboiler liquid outlet, whilst excess solvent overflows to the crossflow heat exchanger. A low volumetric flow rate is extracted from the bottom of the reclaimer and cooled, allowing it to be sent to a secondary, semi-continuous reclaiming stage. This first, continuous step, is designed to prevent the accumulation of degradation products within the main reclaiming vessel

It is important to note that the collected solvent that is extracted from the continuous reclaimer still contains a significant amount of useful MEA solvent. Therefore, the collected solvent undergoes a second reclaiming stage offline, which aims to maximise solvent recovery. This second-stage reclaiming also handles solvent with higher concentrations of degradation products, which can be more challenging to process. Conducting this reclaiming stage in a batch mode provides additional control over the process, allowing for adjustments such as applying a low vacuum when necessary to ensure all MEA is recovered while minimising further degradation. The solvent recovered from the second-stage reclaiming can then be reused in the SMART capture unit.

The first stage of the reclaimer has a water feed into the reclaimer inlet with the option of dosing caustic straight into the bottom of the reclaimer, if this is required. However, testing will determine if this is necessary in the first stage, as this might be reserved for only the second stage, to displace MEA in HSS complexes.

The reclaimer is designed to run at the same pressure as the desorber to maximise energy recovery back to the desorber. This increased pressure causes an increase in the boiling point of MEA, resulting in a higher temperature in the reclaimer. Whilst this higher temperature may cause increased degradation in the reclaimer itself, it is expected that the energy recovery offsets the increased rate of degradation in the reclaimer as less-volatile degradation products will be continuously removed. Recent work has also suggested that this increased rate of thermal degradation through operating the desorber at lower lean loading is partially or fully offset by the reduced carbamate concentration and in turn the reduced reaction rates [15,16]. Figure 4 shows the setup of the standalone reclaimer, which will be able to test thermal reclaiming operation under a range of conditions.

Figure 4: - Prototype Standalone Thermal Reclaimer

The P&ID of the standalone reclaimer is shown in Figure 5.

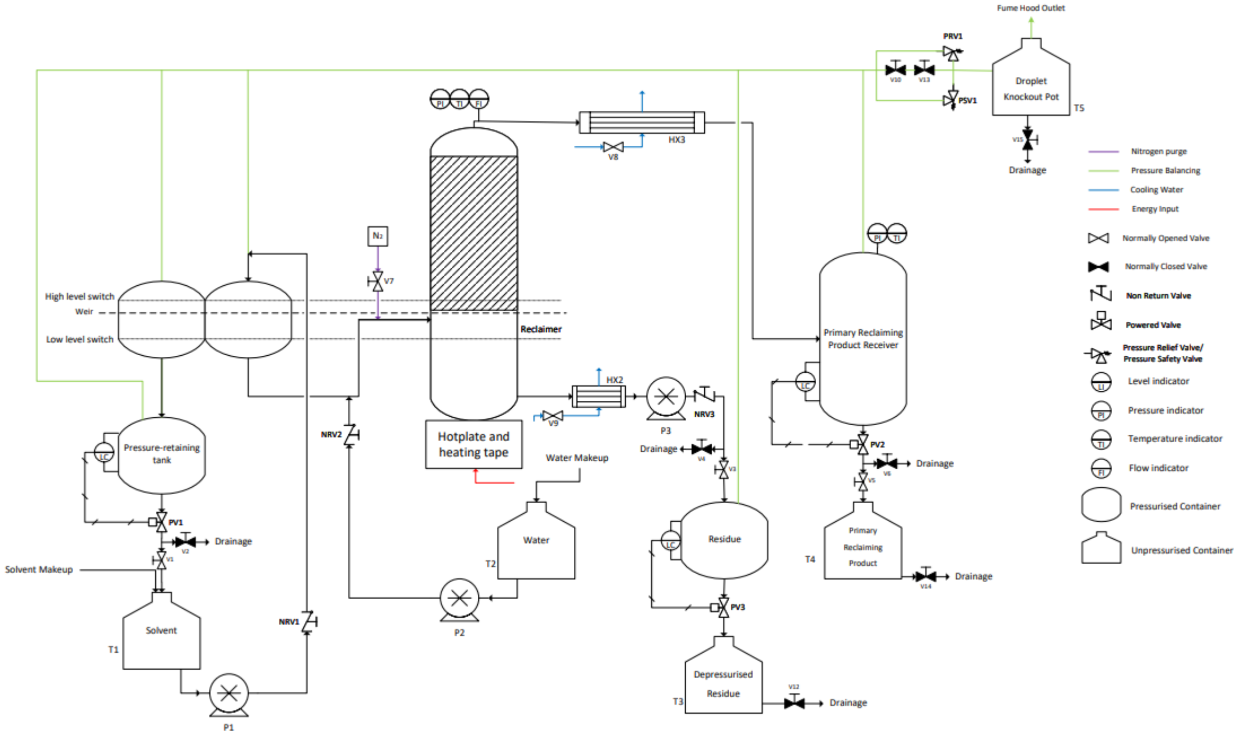

Figure 5: - P&ID of standalone reclaimer

3. Reclaimer Process Modelling

Limited research on continuous thermal reclaiming of MEA under different reclaimer conditions has been undertaken to date [17]. As such, a model of the thermal reclaiming process conditions and how they affect the efficacy of reclaiming was undertaken using MathCAD, validated using data from ASPEN Plus and experimental data, and made into an accessible Excel file. The intention is to provide an open access Excel model which is able to predict trends, allowing thermal reclaimer operatives to see the effects of changing conditions in the reclaimer, verified both experimentally, and using data from ASPEN Plus. The authors plan to make this a readily available downloadable file in the future. The similarity of the excel file to the MathCAD data is covered elsewhere [18] The derivation of the model is shown in Appendix 1.

The model requires the vapour pressure for any produced degradation products to ensure accuracy. However, the only common degradation product with accessible vapour pressure data to the authors' knowledge is HEEDA, from Aspen Plus. As the composition of degraded MEA is variable, the degradation product HEEDA is a good degradation product to use to indicate the less volatile degradation products, as HEEDA is a molecule that is always found within MEA degradation product systems, and is also a precursor to HEIA, one of the most stable degradation products. If vapour pressure data becomes available for other degradation products, the model could be expanded to include these.

3.1. Model results, discussion and analysis

At a constant pressure of 2 bar, the effect of the reclaiming conditions for a constant reclaimer feed of 100 mol h^{-1} on the temperature operation are as is shown in Figure 6.

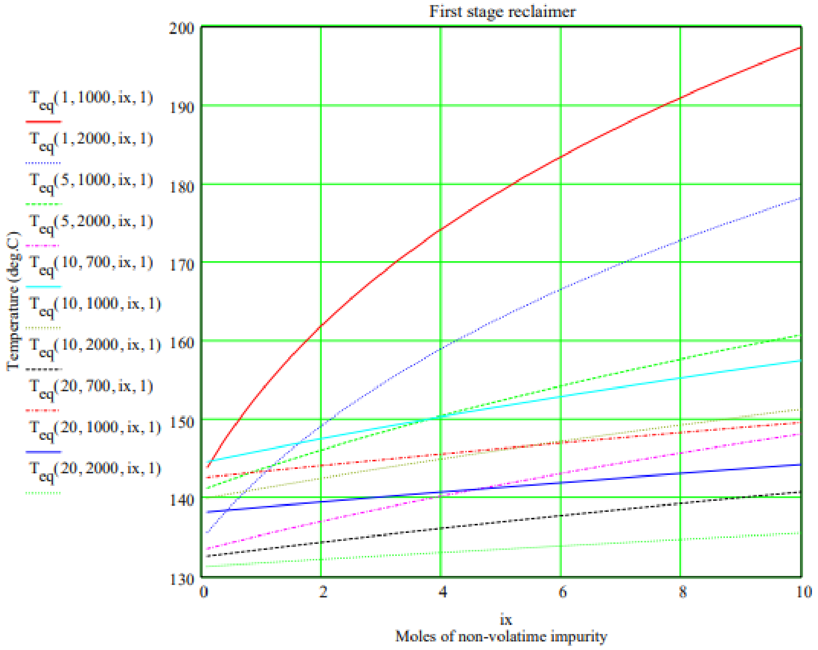

Figure 6: - Reclaimer temperature as a function of moles of non-volatile impurity in the blowdown.

Where for T_{eq}(A, B, C, D), A is the MEA in the blowdown [mol h^{-1}], B is the H_2O in the reclaimer feed [mol h^{-1}] C is the non-volatile impurity in the reclaimer feed [mol h^{-1}] and D is the moles of HEEDA in the reclaimer feed [mol h^{-1}] at 2 bar(a).

Figure 6 shows that as the moles of MEA in the blowdown increases, the temperature in the reclaimer significantly decreases, and the less significant an increase in the moles of non-volatile impurity has on the temperature increase in the reclaimer. For a constant molar flowrate of MEA and non-volatile impurity in the blowdown, increasing the molar flowrate of water in the reclaimer feed decreases the reboiler temperature, whilst increasing the moles of non-volatile impurity significantly increases the temperature in the reboiler. Knowledge of the effects of changing these conditions in the reclaimer is key to balancing factors such as energy use and MEA recovery on one hand with minimising additional degradation in the thermal reclaimer on the other.

Figure 7 shows the effects of operating the model at different reclaimer pressures for a constant water feed rate.

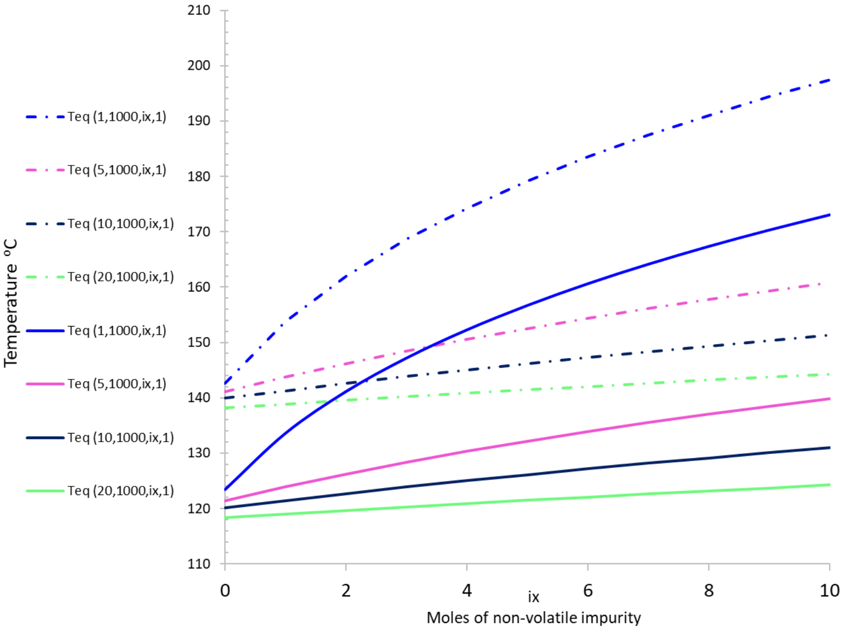

Figure 7: - Reclaimer temperature as a function of moles of non-volatile impurity in the blowdown at different pressures. The dotted lines show reclaimer operation at 2 bara, the solid lines show the operation at 1 bara.

Figure 7 shows that in the case of the SMART reclaimer, operating at the same pressure as the desorber (~2.4 bara) means that elevated temperatures in the reclaimer are unavoidable, compared with operating at atmospheric pressure. As such, a second, semi-continuous reclaimer operating under different conditions will be required to maximise MEA recovery through caustic dosing and high rates of MEA evaporation, decreasing the temperatures required by operating at reduced pressures (probably atmospheric, but if necessary sub-atmospheric pressure operation can also be considered) . The development of this second stage reclaimer is comprehensively covered in another paper [18].

The limitations of this model include that it produces more accurate results with clean solvent composed of only MEA and water, however, actual composition in the solvent includes a complex mixture of degradation products, which are not simple to model.

4. Conclusion

The SMART project addresses the requirement of multiple, parallel long-term tests (>8000 hours) of thermally reclaimable solvents on different flue gasses through the production of a standard rig with integrated continuous thermal reclaiming. The overall aim with the SMART unit is to produce realistic degradation products and to verify that they can be removed with an integrated reclaimer that does not impose severe energy demands and can be run up to 1 inventory turnover/week, or even higher if needed, for upsets. Very precise solvent replacement rates are going to be plant-specific given the complex mechanisms suggested in literature and the dependence on the actual materials of construction, residence times and operating methods – even if exactly the same flue gas is available to use.

Whilst it may not be possible to reproduce degradation rates, it may be possible to reproduce reclaiming. To successfully run the reclaimer, the input energy must be minimized to make it economically feasible to run for a long term at a rate of up to one inventory per week. The temperature in the reclaimer must also be minimised whilst still maintaining efficacy, so as to not needlessly increase the thermal degradation of the solvent in the reclaimer. Long term testing will enable analysis of the solvent to pick up if any degradation products are accumulating.

Most of the detailed design and procurement work with SMART is now completed and construction is at an advanced stage. The stand-alone reclaimer is being built first, as this is the key innovation and useful in its own right.

5. Future Work

The next stages in developing and deploying the SMART rig will be HAZOP, followed by absorber and stripper operation verification and integration. Future work will include experimental and computational work running the standalone reclaimer in single and two-stage configurations, modelling for a variable first stage blowdown composition. Experimental work will be done running the SMART rig long term on a slipstream of flue gas.

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Appendix 1 – Derivation of an approximate reclaiming model

This starts with the Raoult's Law approximation for vapour molar ratio:

$$\frac{\boldsymbol{MEA_{BD} \times P_{MEA}(T)}}{\boldsymbol{H2O_{BD} \times P_{H2O}(T)}} = \frac{\boldsymbol{MEA_{VAP}}}{\boldsymbol{H2O_{VAP}}} \tag{A1.1}$$

Where MEA_{BD} and MEA_{VAP} is the amount in moles of Monoethanolamine in the blowdown and in the vapour phase respectively, $H2O_{BD}$ and $H2O_{VAP}$ is the amount in moles of water in the blowdown and in the vapour phase respectively and $P_{MEA}(T)$ and $P_{H2O}(T)$ are the vapour pressures of MEA and H2O at temperature T.

Assuming ideal gas mixtures:

$$\boldsymbol{y_i = \frac{P_i}{P_T}} \tag{A1.2}$$

Where y_i is the vapour mole fraction of component i, P_i =the partial pressure of component i and P_T is the total pressure of the system.

Ratio of mole fractions in the vapour phase of two components MEA and H_2O is proportional to the ratio of their partial pressures:

$$\therefore\ \boldsymbol{\frac{y_{MEA}}{y_{H2O}} = \frac{P_{MEA}}{P_{H2O}}} \tag{A1.3}$$

According to Raoult's Law:

$$\boldsymbol{P_i = x_i \times P_i(T)} \tag{A1.4}$$

Where x_i is the liquid mole fraction of component i, and $P_i(T)$ is the vapour pressure of component i at temperature T.

$$\therefore\ \boldsymbol{\frac{y_{MEA}}{y_{H2O}} = \frac{MEA_{BD} \times P_{MEA}(T)}{H2O_{BD} \times P_{H2O}(T)}} \tag{A1.5}$$

Now writing y_i as y_{VAP} yields:

$$\boldsymbol{\frac{MEA_{VAP}}{H2O_{VAP}} = \frac{MEA_{BD} \times P_{MEA}(T)}{H2O_{BD} \times P_{H2O}(T)}} \tag{A1.6}$$

Derivation of H_2O and HEEDA in the blowdown, $H2O_{BD}$ and $HEEDA_{BD}$:

$$\boldsymbol{H2O_{BD} = \frac{H2O_{in}}{[MEA_{BD} \times P_{MEA}(T) + (MEA_{in} - MEA_{BD}) \times P_{H2O}(T)]} \times \left(MEA_{BD} \times P_{MEA}(T)\right)} \tag{A1.7}$$

$$\boldsymbol{HEEDA_{BD} = \frac{HEEDA_{in}}{[MEA_{BD} \times P_{MEA}(T) + (MEA_{in} - MEA_{BD}) \times P_{HEEDA}(T)]} \times \left(MEA_{BD} \times P_{MEA}(T)\right)} \tag{A1.8}$$

Where MEA_{in}, $H2O_{in}$ and $HEEDA_{in}$ is the amount of MEA, H_2O and HEEDA in moles entering the system.

From mass balances:

$$\boldsymbol{H2O_{BD} = H2O_{in} - H2O_{VAP}} \tag{A1.9}$$

$$\boldsymbol{MEA_{BD} = MEA_{in} - MEA_{VAP}} \tag{A1.10}$$

$$\boldsymbol{HEEDA_{BD} = HEEDA_{in} - HEEDA_{VAP}} \tag{A1.11}$$

From the Raoult's Law approximation for vapour molar ratio:

$$\boldsymbol{\frac{MEA_{VAP}}{H2O_{VAP}} = \frac{MEA_{BD} \times P_{MEA}(T)}{H2O_{BD} \times P_{H2O}(T)}} \tag{A1.12}$$

$$\boldsymbol{\frac{MEA_{VAP} \times H2O_{BD} \times P_{H2O}(T)}{H2O_{VAP}} = MEA_{BD} \times P_{MEA}(T)} \tag{A1.13}$$

$$\boldsymbol{\frac{1}{H2O_{VAP}} = \frac{MEA_{BD} \times P_{MEA}(T)}{MEA_{VAP} \times H2O_{BD} \times P_{H2O}(T)}} \tag{A1.14}$$

$$\boldsymbol{H2O_{VAP} = \frac{MEA_{VAP} \times H2O_{BD} \times P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}} \tag{A1.15}$$

Sub in $\boldsymbol{H2O_{VAP} = H2O_{in} - H2O_{BD}}$

$$\boldsymbol{H2O_{in} - H2O_{BD} = \frac{MEA_{VAP} \times H2O_{BD} \times P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}} \tag{A1.16}$$

$$\boldsymbol{H2O_{BD} = H2O_{in} - \frac{MEA_{VAP} \times H2O_{BD} \times P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}} \tag{A1.17}$$

$$\boldsymbol{H2O_{BD} = H2O_{in} - MEA_{VAP} \times H2O_{BD} \times \frac{P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}} \tag{A1.18}$$

$$\boldsymbol{H2O_{BD} + MEA_{VAP} \times H2O_{BD} \times \frac{P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)} = H2O_{in}} \tag{A1.19}$$

Sub out for $H2O_{BD}$:

$$\boldsymbol{H2O_{BD}\left[1 + MEA_{VAP} \times \frac{P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}\right] = H2O_{in}} \tag{A1.20}$$

$$\boldsymbol{H2O_{BD} = \frac{H2O_{in}}{\left[1 + MEA_{VAP} \times \frac{P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}\right]}} \tag{A1.21}$$

Sub in $\boldsymbol{MEA_{VAP} = MEA_{in} - MEA_{BD}}$

$$\boldsymbol{H2O_{BD} = \frac{H2O_{in}}{\left[1 + (MEA_{in} - MEA_{BD}) \times \frac{P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}\right]}} \tag{A1.22}$$

Multiplying numerator and denominator by $[\mathbf{MEA_{BD}} \times \mathbf{P_{MEA}}(\mathbf{T})]$

$$\boldsymbol{H2O_{BD} = \frac{H2O_{in} \times [MEA_{BD} \times P_{MEA}(T)]}{[MEA_{BD} \times P_{MEA}(T)]\left[1 + (MEA_{in} - MEA_{BD}) \times \frac{P_{H2O}(T)}{MEA_{BD} \times P_{MEA}(T)}\right]}} \tag{A1.23}$$

$$\boldsymbol{H2O_{BD} = \frac{H2O_{in} \times [MEA_{BD} \times P_{MEA}(T)]}{[MEA_{BD} \times P_{MEA}(T)] + (MEA_{in} - MEA_{BD}) \times P_{H2O}(T)}} \tag{A1.24}$$

$$\boldsymbol{\therefore H2O_{BD} = \frac{H2O_{in}}{[MEA_{BD} \times P_{MEA}(T)] + (MEA_{in} - MEA_{BD}) \times P_{H2O}(T)} \times [MEA_{BD} \times P_{MEA}(T)]} \tag{A1.25}$$

Similarly, for HEEDA:

$$\boldsymbol{\therefore HEEDA_{BD} = \frac{HEEDA_{in}}{[MEA_{BD} \times P_{MEA}(T)] + (MEA_{in} - MEA_{BD}) \times P_{HEEDA}(T)} \times [MEA_{BD} \times P_{MEA}(T)]} \tag{A1.26}$$