Efficient Utilization of Industrial Excess Heat for Post-combustion CO2 Capture: An Oil Refinery Sector Case Study

(article starts on next page)

GHGT-12

Efficient Utilization of Industrial Excess Heat for Post-combustion CO$_2$ Capture: An Oil Refinery Sector Case Study

Viktor Andersson[1]*, Henrik Jilvero[1], Per-Åke Franck[2], Fredrik Normann[1], Thore Berntsson[1]

[1]*Chalmers University of Technology, Department of Energy and Environment, 412 96, Göteborg, Sweden*
[2]*CIT Industriell Energi AB, Chalmers Teknikpark, 412 88, Göteborg, Sweden*

Abstract

A key issue in post-combustion carbon capture is the choice of absorbent. In this paper two different absorbents, monoethanolamine (MEA) and ammonia (NH$_3$), have been modeled in Aspen Plus at different temperatures for possible implementation at an oil refinery. The focus of investigation is the possibilities of heat integration between the oil refinery and the carbon capture process and how these possibilities could change in a future situation where energy efficiency measures have been implemented.

The results show that if only using excess heat from the refinery for heating of the carbon capture process, the MEA process can capture more CO$_2$ than the NH$_3$ process. It is shown that the configuration requiring least supplementary heat when applying carbon capture to all flue gases is MEA at 120 °C.

The temperature profile of the excess heat from the refinery suits the MEA and NH$_3$ processes differently. The NH$_3$ process would benefit from a flat section above 100 °C to better integrate the heat needed to reduce slip, while the MEA process only needs heat at stripper temperature.

Keywords: Oil refinery, Carbon capture, Heat integration

* Corresponding author. Tel.: +46-31-772-3019
  E-mail address: viktor.andersson@chalmers.se

1. Introduction

Since the industrial processes emitting CO_2 are inherently different, different carbon capture technologies are required. Case studies of various sectors (power, cement, oil refineries etc.) are therefore needed to determine a good match between the industrial process and the capture technology. This paper focuses on post-combustion capture through chemical absorption in the oil refining sector.

Monoethanolamine (MEA) has been used for many years to separate CO_2 in the gas processing industry, and has therefore become the benchmark to which other CO_2 capture technologies are compared. Two of the most important characteristics of a good absorbent of CO_2 are a low heat of reaction with CO_2 (low specific heat demand) and a low temperature for the regeneration process. The regeneration temperature is of special interest when excess heat is available, since the amount of excess heat may be considerably larger at lower temperatures. The residual heat from the CO_2 emitting process can be used to cover parts of or the whole heat requirement of the CO_2 capture process.

Ammonia (NH_3) emerged as a promising solvent in the beginning of the 21th centry [1], mainly due to its persistence to degradation, which is a major problem for alkanolamine-based solvents. The most notable disadvantage with NH_3 is its high vapour pressure, which result a loss of solvent, ammonia slip. The company Alstom is developing a proprietary NH_3-based post-combustion capture process called "the chilled ammonia process" (CAP). In a first configuration the absorber was chilled to very low temperatures (0-20°C), which would allow precipitation to occur in the solvent and a low slip of NH_3. However, recent developments have deemed this configuration too cumbersome and a design similar to that of the MEA-based process is suggested [2].

Early evaluations of NH_3 as a solvent have suffered from optimistic estimations of the heat requirements for NH_3 regeneration where as low requirements as 900-1000 kJ/kg CO_2 captured was reported [1,3]. These estimates have later been revised and more recent figures reports a heat requirement of approximately 2500 kJ/kg CO_2 captured [2]. MEA is a commercially used solvent and as such has a well mapped heat requirement for regeneration of around 3300 – 3800 kJ/kg CO_2 captured[4–6]. Excess heat utilization could prove to significantly reduce the primary energy demand of the CO_2 capture process [7,8].

When excess heat is used to cover some of the heat requirement of the capture process it is important to evaluate most suitable operating temperature/pressure of the stripper. The heat requirement as a function of stripper conditions has been studied in Oyenekan & Rochelle [9] who investigated reboiler duty of three different stripper configurations when applying CCS to a coal power plant. The configurations were simple stripper, multipressure stripper and vacuum stripper (30 kPa, 60 – 80 °C). They concluded that for MEA the multipressure stripper resulted in the lowest reboiler duty, and that the highest reboiler duty was associated with vacuum stripping, approximately 22 – 62% higher penalty than for standard pressure stripping. Berstad et al. [10] varied the reboiler temperature between 107 and 142 °C when applying CCS to three different applications: a coal power plant, a natural gas combined cycle (NGCC) and a biomass power plant. The study concluded that the low temperature cases resulted in a lower efficiency penalty compared to the reference case for both the NGCC and the biomass plant due to the use of steam at lower pressure in the regeneration process. In addition to this Abu-Zahra et al. [4] performed a study varying the stripper temperature between 108 and 128 °C, determining that the lowest energy demand is present when regenerating MEA at 128 °C. The heat requirement as a function of stripper pressure has rarely been subject evaluation for the NH_3 process. In the work by Linnenberg et al. [11] it was concluded that the heat requirement for the stripper is slightly decreasing for a higher pressure. However, when integrated with a power plant net efficiency decrease would reach a minimum at a stripper pressure at 4 bar, due to the permission to use low quality steam from the steam cycle.

Since none of these studies took into account the benefits of using excess heat to decrease the primary heat demand of the process, the authors of this paper conducted a heat integration study where an MEA-based capture plant was heat integrated with an oil refinery while varying the temperature between 90 and 120 °C [12]. The paper concluded that when being restricted to a small amount of heat collecting units the drawbacks of an increased heat demand of the capture plant due to lower stripper temperature outweighs the benefit of being able to use more excess heat. Benefits were however present when being able to utilize the full excess heat potential of the refinery.

Currently, incentives are being set in place in an effort to increase energy efficiency in industry sector. As the implementation of CCS is halted by unclear future policies it is believed to be deployed on a medium term time scale whereas energy efficiency measures are likely to be implemented before CCS. The implementation of energy efficiency measures has at least two implications on excess heat driven CCS. The most tangible is the loss of extractable excess heat. However, this is partly counteracted by the fact that less CO_2 will be produced due to a

lower fuel demand. Case studies on current systems as well as future situations are therefore essential to map the possibilities of using excess heat for the CCS process in a future perspective.

This paper highlights the possibilities of decreasing the primary heat demand through heat integration when performing post-combustion CO_2 capture at an oil refinery with either MEA or NH_3 as absorption fluid. The authors want to illustrate how the availability of excess heat can change the operating parameters of the CCS process by investigating two different absorbents at varying temperature. In addition to the two absorbents, heat integration is performed at two different levels of excess heat availability to simulate possible future energy efficiency measures.

Aspen Plus is employed to evaluate the heat requirement of both capture processes at different operating conditions of the regeneration (e.g. pressure and temperature). The results will reveal at which stripper operating conditions the difference between the heat requirement of the capture processes and the available excess heat are at a minimum. This analysis will result in an identification of a suitable heat integration strategy for carbon capture at the present oil refinery both in a current and future perspective.

2. Methodology

The methodology in this paper is divided into two stages. In the first stage models were developed for both the MEA and NH_3 processes and in the second step these models were used in conjunction with operating data from an oil refinery to perform a heat integration study. The main process operating parameter studied in this paper is stripper temperature.

2.1. Modelling work

The models presented in this paper have both been published in previous works by the authors [2,12,13]. The models are modified to include the following key assumptions:

- Flue gas composition with 11.8 % CO_2 after dewatering (10.8 % before dewatering) [14]
- Minimum temperature difference (ΔT_{min}) in lean/rich heat exchanger of 10 K
- Basis of 400 000 t CO_2/y with an annual operating time of 8 200 h.

For MEA and NH_3 respectively, three stripper temperatures have been modelled. The MEA process has been modelled at stripper temperatures of 90, 105 and 120 °C whereas the NH_3 process has been modelled at temperatures of 105, 120, 135 and 155 °C. The NH_3 process has not been modelled for 90 °C due to the extensive NH_3 slip. The slip increases with decreasing temperature and has a noticeable impact already at 105 °C.

Both the MEA and NH_3 processes were simulated in Aspen Plus using the default property method packages for each absorbent. Table 1 presents the most important relevant process operating conditions for both processes. For a more detailed description of the MEA and NH_3 process models see Andersson et al. [12] and Jilvero et al. [2].

Table 1: Operating conditions for the MEA and the NH_3 processes.

	MEA process	NH_3 process
Absorbent concentration (CO_2-free) (wt%)	30	14.3
Lean CO_2 loading (mol CO_2/mol abs.)	0.32	0.25
Rich CO_2 loading (mol CO_2/mol abs.)	0.497	0.5
CO_2 Capture efficiency (%)	85	85

Figure 1 presents a general schematic of the MEA and NH3 processes. Both processes are based on the same absorber/stripper setup. However, the fact that NH_3 is considerably more volatile than MEA requires extra consideration in the design of the capture process. An extra absorber (A2) is used to reduce NH_3 discharge from the absorption process. The remaining ammonia in the flue gases are captured and retained by an NH_3 abatement cycle formed by a water-wash (A3) and an NH_3 stripper (S2). There is also a heat requirement for the NH_3 stripper which needs to be considered in the heat integration. The operating temperature of the NH_3 stripper is 100°C. Extra attention is also put to the stripper condenser where a water-wash condenser is used to eliminate any possible solid precipitation.
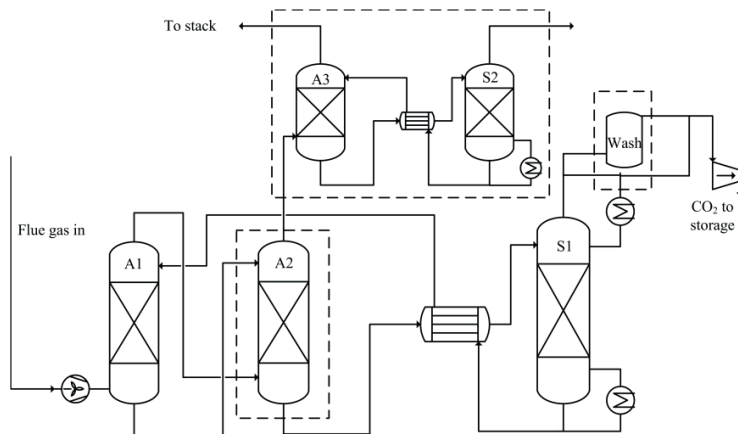
Figure 1: General setup of the MEA/NH$_3$ absorption processes. Equipment within dashed lines is only used in the NH$_3$ process. A1-3 are absorbers, S1-2 are strippers and the wash section is used in the NH$_3$ process instead of a condenser.

2.2. Reference oil refinery plant

The focus of this case study is a complex oil refinery, emitting approximately 1.8 Mt CO$_2$/y, situated on the west coast of Sweden. The main part (1.74 Mt CO$_2$/y) of the CO$_2$ is emitted from 4 chimneys situated in the middle of the refinery and these are the chimneys deemed available for CCS in this study. The CO$_2$ concentration in the flue gases ranges from 6.7% to 24% before dewatering [14].

The oil refinery has a primary heat demand of 409 MW but a detailed heat integration study revealed that theoretical minimum heat demand is only 199 MW [15]. Thus, the theoretical heat savings amount to 210 MW although this potential is neither economically nor practically feasible to implement. In this work it is assumed that 20 % (42 MW) of the proposed energy efficiency measures can be implemented in a medium-term time perspective. The energy savings are assumed to be evenly distributed in the temperature interval 170 – 130 °C and affect both the CO$_2$ emissions from the refinery (due to fuel savings) and the extractable excess heat.

The available excess heat in the two cases investigated is graphically presented in an so called "actual cooling load curve" (ACLC) [16], see Figure 2. The ACLC is constructed by adding all heat available in hot streams that are cooled by air or cooling water and plot the sum of heat against the temperature at which it is available. Flue gas currently not used to produce steam is also included in the analysis. The amounts of heat available are extracted from real plant data, thus no restrictions have been set as to how the heat is collected.

When calculating the CO$_2$ emission savings from decreased fuel use, a mixture of 50 % propane and 50 % butane is assumed as fuel. The boiler producing steam is assumed to have an efficiency of 0.9.

2.3. Heat integration

Heat from the process is assumed to be supplied to the reboiler via a heat collection system. The system consists of a water trunk pipeline leading condensate from the capture plant to the refinery process. The trunk pipeline branches out to a number of heat exchangers where the condensate is vaporized while refinery process streams are cooled. The steam is then lead back to the capture plant through a trunk pipeline. The heat collection system uses 10 K as a minimum temperature difference over the system, which in practice means that all streams in the refinery seen as 10 K below their actual temperature when extractable excess heat is mapped. The mapping of heat flows in the refinery processes is carried out first for the present situation and then for a future situation where energy efficiency measures have been taken. The energy efficiency measures of 42 MW steam changes the appearance of the ACLC. In Figure 2 the ACLCs before (current situation) and after (future situation) energy efficiency measures are shown.
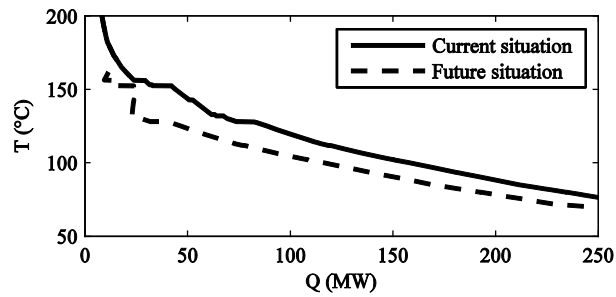
Figure 2: The ACLC before and after energy efficiency measures are carried out.

In the mapping of heat in within the capture process, the minimum allowed temperature difference in a heat exchanger consists of the sum of one contribution from each side in the unit. For condensing/evaporating streams this contribution is 2.5 K, for liquid streams 5 K and for gaseous streams 7.5 K. The mapping of heat flows in the capture process is done by Pinch technology. Pinch technology is a systematic analysis of heat demands in an industrial process, and is based on first and second law of thermodynamics. The method consists of systematic collection and treatment of heat flows and can be used as a targeting method towards a theoretical minimum heat demand. The process´ minimum heat demand is calculated for a given minimum temperature difference in the heat exchanger network ΔT_{min} and the methodology also gives indications on suitable heat exchange within the process. For a comprehensive methodology description, see Smith [17].

The largest heat demands in the two processes are the stripper reboiler (for both processes) and for the NH_3 process also the NH_3 abatement cycle stripper, which removes the slip from the absorber.

The outcome from the heat integration study is how much excess heat that can be collected from the refinery to the MEA and NH_3 plants at the different temperature levels, how much CO_2 that can be captured using only excess heat and also the need for supplementary heating if choosing to capture 85 % of the CO_2 from the 4 main chimneys.

3. Results

The modelling of the MEA and the NH_3 processes yields the heat demands of the CO_2 capture processes. Implementation of energy efficiency measures would result in approximately 100 kt less CO_2/y which corresponds to 6 % of the annual CO_2 emissions. The amounts of CO_2 to be captured are therefore 1.48 Mt/y if CCS is deployed before energy efficiency measures and 1.40 Mt/y if CCS is deployed after energy efficiency measures, resulting in a 5 % decrease in heat demand. The heat demands as well as available amounts of excess heat are shown in Figure 3.
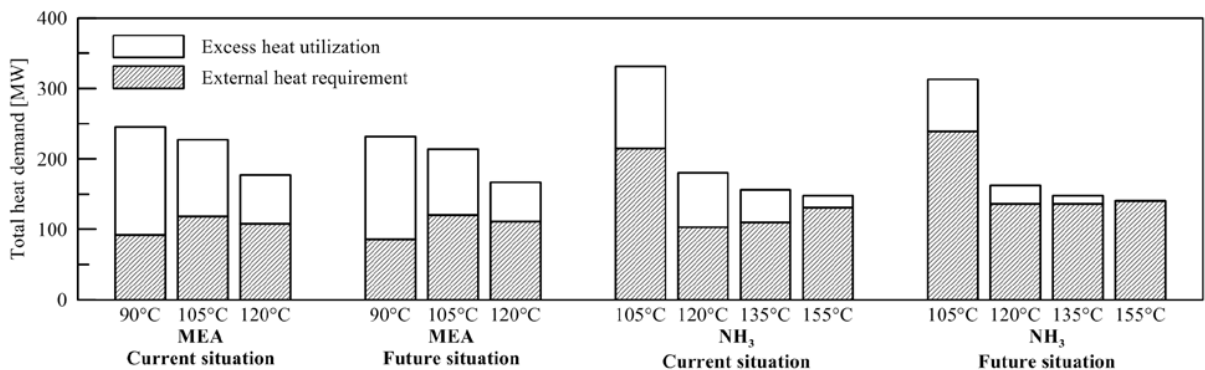

Figure 3: Heat demands from the CCS capture plant for before (Current situation) and after (Future situation) energy efficiency measures are being implemented.

It is evident that NH_3 is sensitive to operating at temperatures below 120 °C. As expected, and also as has been shown in previous work, the heat demand increases for all absorbents when the stripping temperature is decreased. However, the magnitude of the change for NH_3 when shifting down to 105 °C is larger than expected.

Figure 4 shows the equilibrium conditions at the top of the stripper as a function of temperature with NH_3 (a) and MEA (b) as absorbent of CO_2. As a representative case for both the MEA and NH_3 processes a CO_2 loading of 0.5 and a vapour fraction of 0.05 is shown. The flow of solvent and water varies considerably relative the flow of CO_2 to compression (index of 1). For the MEA system, the dominating component is water; MEA is not present in the gas phase. The amount of water evaporated with every kg of CO_2 is increasing with decreasing temperature, which means that the heat of regeneration is increased. For the NH_3 system CO_2 is the dominating component into the compressor from 90 °C and upwards. However, at lower temperature almost as much NH_3 as CO_2 exits the stripper. The ammonia needs to be separated from the carbon dioxide before compression. This can be achieved by washing the gas with water, which would cause some of the carbon dioxide to be absorbed and increase the heat required per kg CO_2 captured. This means that both processes are penalized by a lower stripper pressure, but in different ways. The MEA process suffers from a high evaporation of water while the NH_3 process suffers from loss of absorbent.

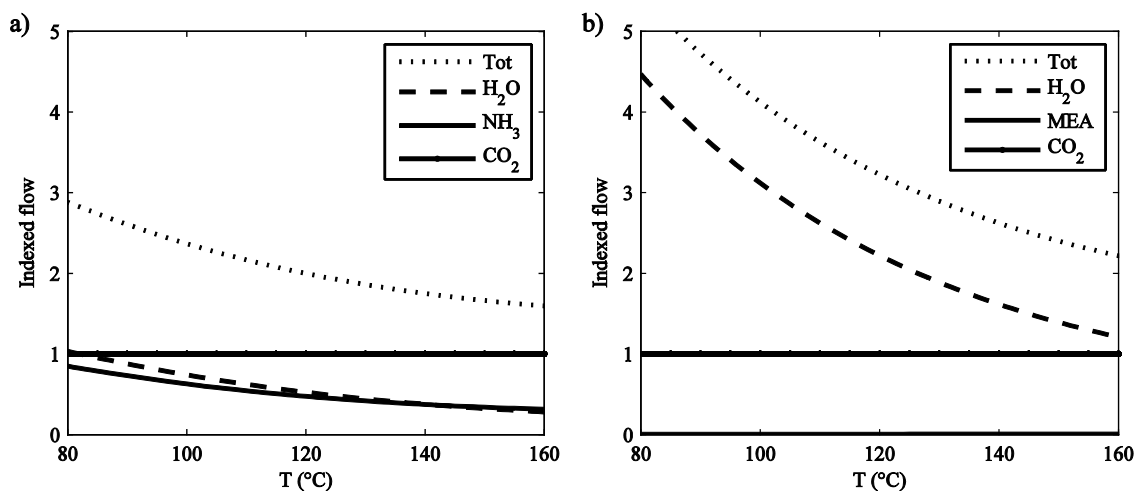Figure 4: Equilibrium of gas phase of an absorbent/CO_2/H_2O mixture at varying temperatures and a vapour fraction of 0.05. The absorbents are a) NH_3 and b) MEA.

A Background/Foreground analysis shows the heat flows of the processes that are being integrated with each other in the same figure. The overlap between the two curves is the amount of heat that can be recovered from the background process (in this case the oil refinery) to cover a demand in the foreground process (the CCS plant). Between the refinery process and the CCS process there needs to be a heat collecting system, as described in Andersson et al. [12], and in order to account for the ΔT_{min} the curves have been shifted 10 K. Figure 5 shows the background/foreground analysis of the NH_3 process at 105 °C and 135 °C respectively. When comparing the graphs representing the NH_3 processes in Figure 5 they both have the largest deficit of heat at the required stripper reboiler temperature. The slip of NH_3 from the absorber increases as the stripper temperature decreases. This is due to that the slip of NH_3 is higher as the temperature of the stripper decreases and thus more NH_3 needs to be added to the lean stream as make-up. Thus, the partial pressure of NH_3 is increased.

Background/foreground analyses have also been made for the different setups of the MEA system and the resulting figures for 90 °C and 120 °C are shown in Figure 6. The heat profile of the MEA process does not change to the same extent as for NH_3 but is altered due to higher heat demand.
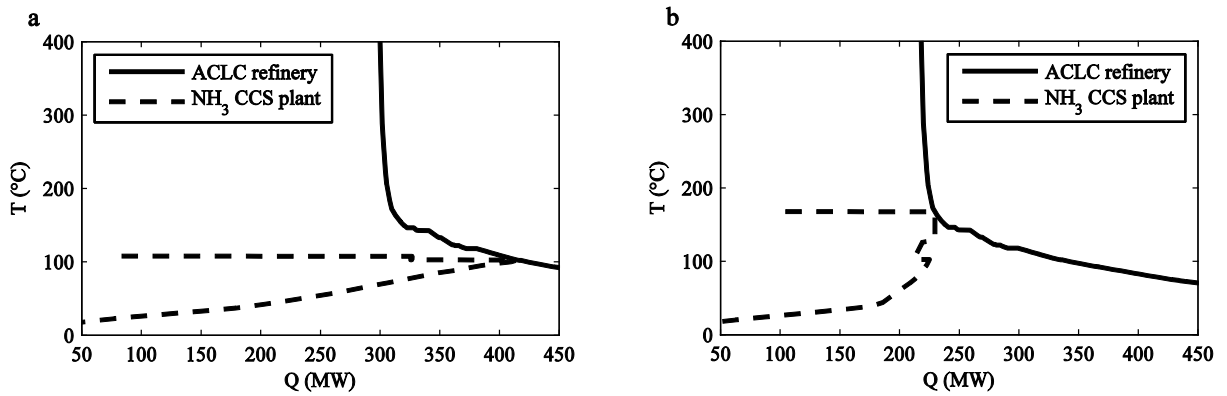
Figure 5: Background/foreground analyses showing the energy demand and the heat integration possibilities for the NH$_3$ process at a stripper temperature of a) 105 °C and b) 135 °C.
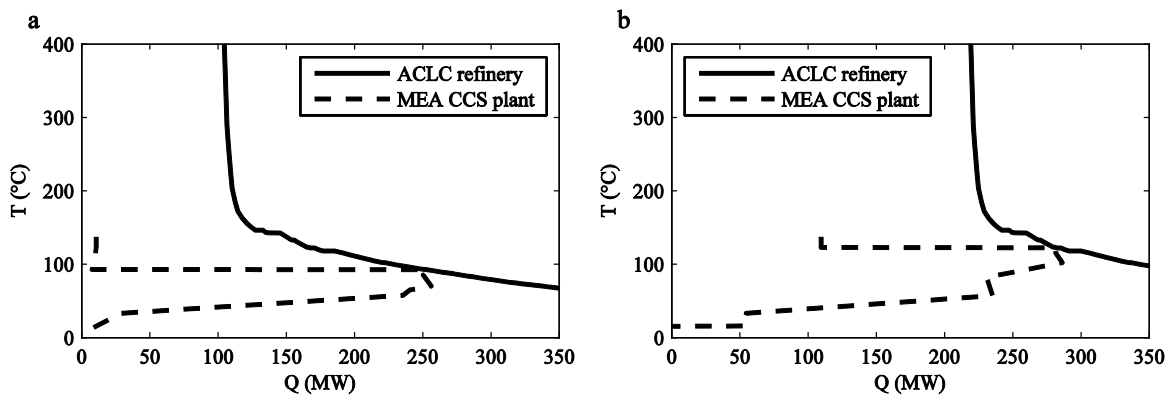

Figure 6: Background/foreground analyses showing the energy demand and the heat integration possibilities for the MEA process at a stripper temperature of a) 90°C and b) 120 °C.

For both setups of the MEA process the pinch point (where the two curves collide) is located at stripper temperature, which means that all available heat from the refinery process streams may be used in the stripper. This can be compared to the NH$_3$ process at 105 °C shown in Figure 5a where the pinch point is located at abatement cycle temperature, which means that the abatement cycle requires more heat than what is available at its temperature level and it will require some of the heat that could have been used in the stripper, increasing the need for primary heat to the CCS plant.

In Figure 5 and 6 the amount of heat that can be recovered from the refinery processes to the CCS process is illustrated by the overlap between the refinery ACLC and the CCS process curve. The remaining heat demand has to be covered by an external source e.g. a boiler or a heat pump. After performing a background/foreground analysis for the different setups in this study the remaining heat demands are shown in Figure 3. The remaining heat demand can be satisfied with a variety of boilers or a heat pump, but the refinery can also decide to separate only as much CO$_2$ as can be covered by the available excess heat. Table 2 shows how much CO$_2$ that can be separated using only excess heat, and also how much CO$_2$ that would remain from the 4 main chimneys after CCS is implemented. The results presented give the somewhat paradoxical conclusion that energy efficiency combined with CCS in this case actually would result in larger CO$_2$ emissions than only performing CCS. Only if operating with MEA at a stripper temperature of 90 °C do the CO$_2$ emissions decrease more for the combined measures. For the 90 °C MEA case the heat demand is larger than for the other processes, and the 42 MW of fuel that could be saved give rise to more CO$_2$ emissions than what could be captured using 42 MW of fuel for carbon capture. The results in Table 2 are sensitive to assumptions regarding fuel mix and temperature span of energy efficiency measures.

Table 2: The table shows how much CO_2 that can be captured using only excess heat, and how much that will be emitted to the atmosphere or to be captured using primary heat.

Temperature (°C)	90		105		120		135		155	
	CO_2 capt.	CO_2 rem.	CO_2 capt.	CO_2 rem.	CO_2 capt.	CO_2 rem.	CO_2 capt.	CO_2 rem.	CO_2 capt.	CO_2 rem.
MEA Current Situation ($ktCO_2/y$)	920	560	710	770	580	900	-	-	-	-
MEA Future Situation ($ktCO_2/y$)	880	520	610	790	470	930	-	-	-	-
NH3 Current Situation ($ktCO_2/y$)	-	-	520	960	630	850	440	1040	170	1310
NH3 Future Situation ($ktCO_2/y$)	-	-	330	1070	230	1170	110	1290	0	1400

4. Discussion

Industrial based CCS is a key technology to reduce CO_2 emissions to a sustainable level and post-combustion CCS is the most flexible technology since it is an end-of-pipe solution that has no implications on the rest of the industrial process. Post-combustion CCS is, however, expensive, with the main cost being attributed to regeneration heat for the absorbent. A scenario where an industrial plant captures as much CO_2 as possible with internal heat and emits the remaining part of the CO_2 may therefore be of interest. This approach may, as pointed out in connection to Table 2, result in the counterintuitive effect that the implementation of energy efficiency measures increases the emission of CO_2. However, energy efficiency measures will always lead to lower generation of CO_2. It is therefore important to consider the capture unit when optimizing the plant. It is, however, also important to consider the time perspective as energy efficiency measures are being implemented at present and in the near future, while CCS will probably not be deployed at a large scale for 10-20 years. One could also argue that efficiency measures should target low temperature streams, since heat at higher temperature is more usable. In future works investigations regarding temperature span of the energy efficiency will be done to see how large impact this has on the results.

The availability of excess heat can change what characteristics that are desirable in the CCS process. Looking at a stripper temperature of 120 °C the MEA process has a somewhat lower heat demand than NH_3 before integration, but after integration the remaining heat demands show opposite relation. This is due to the fact that, in comparison to the MEA process, more heat in the NH_3 process is attributed to the NH_3 abatement cycle which operates at a lower temperature than the stripper. When the pinch point is located at the stripper temperature the heat demand for washing can be completely covered and does not increase the heat demand of the combined processes. For MEA the heat demand occurs almost entirely in the stripper and the possibilities for heat integration thus occurs at this temperature level. If, however, the temperature level of 105 °C is investigated the benefits of having two different heat demands are not as significant, as the amount of excess heat that can be used in the NH_3 and MEA processes are very similar. As previously stated the NH_3 abatement cycle has a constant temperature independent of stripper temperature, and Figure 5 shows that when having a stripper temperature as low as 105 °C the temperature difference between the NH_3 abatement cycle and the stripper is small, cancelling the possibilities to improve heat integration by doing it at different temperatures. It should be investigated further if the NH_3 process has a point between 105 °C and 120 °C where the benefit of having the reboiler heat demand and NH_3 abatement cycle heat demand can be more utilized. The process at 120 °C could sustain a higher slip since there is unused heat at temperatures corresponding will to the NH_3 abatement cycle.

Modeling the MEA process at a stripper temperature of 90 °C is associated with some uncertainty. Very little experimental work has been done at such low temperatures and the sources available for validation are therefore limited. Some work has however been carried out, for instance Dugas [18] but more in depth studies would be necessary in order to successfully operate a full size plant at this temperature.

5. Conclusions

For the system presented in this case study an MEA process operated at a stripper temperature of 90 °C was shown to have the smallest need for supplementary heating when capturing 85 % of the CO_2 in the flue gases.

Implementing energy efficiency measures in an oil refinery can have implications on design of the CCS process. The preferred operating parameters of the CCS process changed after energy efficiency was improved.

Dividing the heat demand into two different units with different temperature levels (the NH_3 process) did not give the desired gains in heat integration for the cases investigated.

The MEA and NH_3 capture processes are affected by the excess heat temperature profile in different ways. If the temperature profile is steep and the sections between stripper temperature and 100 °C none of the NH_3 abatement cycle can be integrated and MEA could then be beneficial. If the temperature profile is flat below stripper temperature and 100 °C the NH_3 process could benefit from being able to integrate the slip of NH_3.

Implementing energy efficiency measures increased CO_2 emissions from the refinery when at the same time implementing CO_2 capture. This is valid when not using supplementary heating.

The stripper in the NH_3 process is difficult to operate below 120 °C. This is mainly due to that the increased slip of NH_3 from the stripper lead to a cumbersome and energy-intensive process.

6. Acknowledgements

References

[1]   EPRI. Chilled-ammonia Post Combustion CO_2 Capture System - Laboratory and Economic Evaluation Results 2006.
[2]   Jilvero H, Normann F, Andersson K, Johnsson F. Heat requirement for regeneration of aqueous ammonia in post-combustion carbon dioxide capture. International Journal of Greenhouse Gas Control 2012;11:181–7. doi:10.1016/j.ijggc.2012.08.005.
[3]   McLarnon CR, Duncan JL. Testing of Ammonia Based CO_2 Capture with Multi-Pollutant Control Technology. Energy Procedia, 2009;1;1027-34, doi:10.1016/j.egypro.2009.01.136.
[4]   Abu-Zahra MRM, Schneiders LHJ, Niederer JPM, Feron PHM, Versteeg GF. CO_2 capture from power plants. Part I. A parametric study of the technical performance based on monoethanolamine. International Journal of Greenhouse Gas Control 2007;1:37–46.
[5]   Knudsen JN, Jensen JN, Vilhelmsen P-J, Biede O. Experience with CO_2 capture from coal flue gas in pilot-scale: Testing of different amine solvents. Energy Procedia, 2009;1;783–90.
[6]   Mimura T, Simayoshi H, Suda T, Iijima M, Mituoka S. Development of energy saving technology for flue gas carbon dioxide recovery in power plant by chemical absorption method and steam system. Energy Conversion and Management 1997;38:S57–62.
[7]   Hektor E, Berntsson T. Future CO_2 removal from pulp mills - Process integration consequences. Energy Conversion and Management 2007;48:3025–33.
[8]   Johansson D, Sjöblom J, Berntsson T. Heat supply alternatives for CO_2 capture in the process industry. International Journal of Greenhouse Gas Control 2012;8:217–32.
[9]   Oyenekan BA, Rochelle GT. Energy Performance of Stripper Configurations for CO_2 Capture by Aqueous Amines. Industrial and Engnineering Chemistry Research 2006;45:2457–64. doi:10.1021/ie050548k.
[10]   Berstad D, Arasto A, Jordal K, Haugen G. Parametric study and benchmarking of NGCC, coal and biomass power cycles integrated with MEA-based post-combustion CO_2 capture. Energy Procedia 2011;4:1737–44. doi:10.1016/j.egypro.2011.02.048.
[11]   Linnenberg S, Darde V, Oexmann J, Kather A, van Well WJM, Thomsen K. Evaluating the impact of an ammonia-based post-combustion CO_2 capture process on a steam power plant with different cooling water temperatures. International Journal of Greenhouse Gas Control 2012;10:1–14. doi:10.1016/j.ijggc.2012.05.003.
[12]   Andersson V, Franck P-Å, Berntsson T. Industrial excess heat driven post-combustion CCS: The effect of stripper temperature level. International Journal of Greenhouse Gas Control 2014;21:1–10. doi:10.1016/j.ijggc.2013.11.016.
[13]   H Jilvero, Eldrup, N-H, Normann, F, Andersson, K, Johnsson, F, Skagestad, R. Techno-economic evaluation of an ammonia-based post-combustion process integrated with a state-of-the-art coal-fired power plant. Submitted to International Journal of Greenhouse Gas Control 2014.
[14]   Grönkvist S. Specifika förutsättningar för koldioxidavskiljning i Sverige (Specific conditions for carbon dixode capture in Sweden). Ångpanneföreningens forskningsstiftelse; 2010.
[15]   Andersson E, Franck P-Å, Åsblad A, Berntsson T. Pinch analysis at Preem LYR. Chalmers University of Technology; 2013.
[16]   Nordman R, Berntsson T. New pinch technology based HEN analysis methodologies for cost-effective retrofitting. Canadian Journal of Chemical Engineering 2001;79:655–62.
[17]   Smith RM. Chemical Process: Design and Integration. John Wiley & Sons; 2005.
[18]   Dugas. Pilot Plant Study of Carbon Dioxide Capture by Aqueous Monoethanolamine. University of Texas at Austin, 2006.