Simulation and Economic Optimization of Amine-based CO_2 Capture using Excess Heat at a Cement Plant

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Abstract

In order to remove CO_2 from power or process industry, a well-known method is absorption in monoethanolamine (MEA) followed by desorption, and this technology has been in operation for decades. The major challenge is the high energy demand for CO_2 desorption. In many industrial cases, a limited amount of cheap waste heat is available and this makes partial CO_2 capture an interesting option. It is not obvious whether a high removal efficiency from a part of the exhaust or a low removal efficiency from the total exhaust is the optimum solution. In this work, simulations of traditional amine-based CO_2 capture processes are performed with full-flow and part-flow of flue gas. The cost of CO_2 capture is estimated using a detailed factor method and a Lang factor method. It is found that a full-flow alternative is the energy optimum while a part-flow alternative treating 80% of the exhaust gas is the cost optimum. This work shows that the calculated optimum is dependent both on the criteria used and on the selected method.

Keywords: CO_2 capture at cement plant, Aspen HYSYS simulation, Partial capture, Cost estimation

1 Introduction

Global warming due to increased greenhouse gas emissions, especially CO_2 emissions has become a major environmental issue. CO_2 emissions have been tripled from fossil fuel, cement industry and flaring since 1970 (IPCC, 2014). The cement industry accounts for around 5% of anthropogenic greenhouse gas emissions. CCS (Carbon capture and storage) is urgently needed along with other energy efficiency measures to reduce the industrial emissions to a level that will meet the 2°C goal (IEAGHG, 2013). United Nations has set this long term goal to limit the global average temperature to well below 2°C above pre-industrial levels since this would reduce the risks and impacts of climate change (IEA, 2015).

1.1 Aim

The aim of this work is to investigate the energy optimum and cost optimum conditions for CO_2 capture from a cement plant with the use of limited excess heat available from the process. Besides, a task is to compare two cost estimation methods, i.e., detailed factor method and Lang factor method.

The subsequent challenge is to perform a cost-benefit analysis of different cases to evaluate whether it is cost optimum to treat all the exhaust gas or only a part of it. Some previous studies (Park, 2016; Øi et al, 2017) have concluded both that a part-flow alternative is optimum and that a full-flow alternative is optimum. The objective of this work is to analyze whether the calculated optimum is dependent both on the criteria used and on the selected method.

1.2 Literature

There have been numerous studies that perform techno-economic analysis of different CO_2 capture concepts, not only for power industry but also for process industries (Rao et al, 2002; Kuramochi et al, 2012) but detailed studies that investigate waste or excess heat potential from process industries to power post combustion CO_2 capture plants are rare.

(Dong et al, 2012) performed a study of the possibility to utilize waste heat from a cement plant to capture CO_2 effluent from the plant. Up to 78 % capture could be achieved using only waste heat by integrating heat recovery with CO_2 capture.
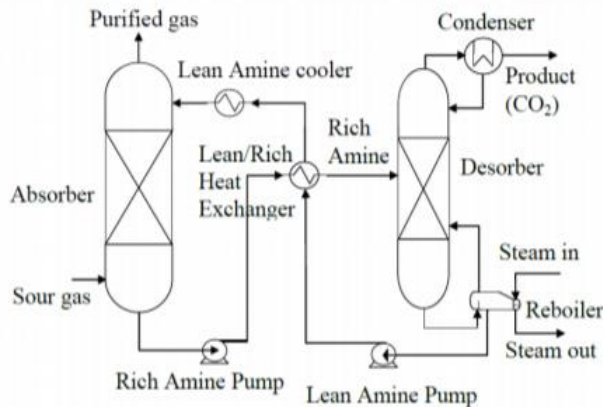
A techno-economic analysis of an oil refinery with amine-based based carbon capture plant has been performed (Andersson et al., 2016). In this work, excess heat from the refinery was shown to decrease specific cost of carbon capture.

The (NORDICCS, 2017) project has evaluated the potential of using waste heat from cement industry to cover the reboiler duty of the stripper for an amine-based CO_2 capture plant and concluded that utilisation of waste heat is necessary in order to lower the cost of CO_2 capture. The CO_2stCap project (Skagestad et al, 2017) is in progress in Norway and Sweden to evaluate different possibilities for partial CO_2 capture from industrial sources.

At the University College of Southeast Norway there have been performed simulations of possible CO_2 capture from Norcem cement plant in Brevik (Svolsbru, 2013). (Park, 2016) simulated partial CO_2 capture and concluded that in case of partial CO_2 capture of approximately 40 % of the CO_2 in the flue gas from a cement plant, treating all the flue gas would probably be more cost optimum compared to treat only a part of the flue gas. (Øi et al, 2017) have performed partial CO_2 capture on a traditional amine-based process and a vapour recompression process and concluded that the process with a low absorption column treating the total exhaust gives the lowest cost per ton CO_2 captured.
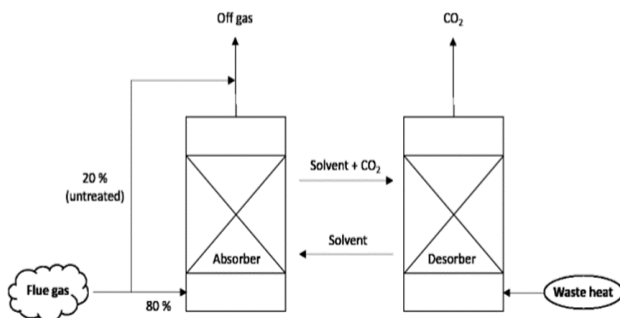
1.3 Process description

Figure 1 shows a standard process for CO_2 absorption into an amine-based solvent. It comprises an absorption column, a stripping column including a reboiler and condenser, circulating pumps and heat exchangers.

Figure 1. Process flow diagram of a standard amine-based CO_2 capture process (Aromada and Øi, 2015)

A sketch of a general post-combustion partial CO_2 capture process is presented in Figure 2. The whole or a part of a flue gas is sent to an absorber where CO_2 is absorbed in a solvent. The solvent is regenerated by releasing the CO_2 in a desorber and the regenerated solvent is sent back to the absorber.

Figure 2. A schematic of partial CO_2 capture (Park, 2016)

2 Methodology

Four case studies are analysed for partial CO_2 capture using only excess heat as mentioned in Table 1.

Table 1. Case studies description

Case study	Description	Flow type
C100	All the flue gas from String 1 goes to the CO_2 capture plant	Full-flow
C80	80% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow
C60	60% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow
C40	40% of flue gas from String 1 goes to the CO_2 capture plant	Part-flow

The cost and energy optimum alternative from the above four case studies was selected for two more case studies, one with lower reboiler temperature (115 °C)

and the other with a plate & frame heat exchanger to be used as lean/rich heat exchanger. The case studies in this work are performed in two parts:

1. Simulation of amine-based CO_2 capture plant
2. Dimensioning and cost estimation of CO_2 capture plant

2.1 Specifications and simulation of standard CO_2 capture process

All case studies were simulated for a standard process as in Figure 1 using Aspen HYSYS version 8.6 by selecting the Kent-Eisenberg vapour/liquid equilibrium model. Aspen HYSYS is a commercial general purpose process simulation program from AspenTech. It contains several equilibrium models, process unit operation models and flow-sheeting calculation alternatives.

The specifications for the full flow case simulation (case C100) are presented in Table 2. The flue gas (string 1) data are from a cement plant, and the excess heat is assumed to be constant 24.5 MW (NORDICCS, 2017). The absorption and desorption columns are simulated with equilibrium stages including a stage efficiency.

Table 2. Aspen Model parameters and specifications for the full flow alternative (Case study: C100)

Simulation parameter	Value
Flue gas (string 1) temperature from process	80 °C
Inlet flue gas temperature to absorber	40 °C
Inlet gas pressure to absorber	1.1 bar
Inlet flue gas molar flow rate	5788 kmol/h
CO_2 in inlet flue gas	22.1 mol-%
Lean MEA temperature	40 °C
Lean MEA pressure	1.01 bar
Lean MEA mass flow rate	527500 kg/h
MEA content in Lean MEA	29.0 mass-%
CO_2 in Lean MEA	5.5 mass-%
Number of stages in absorber	15
Murphree efficiency in absorber stages	0.11 – 0.21
Temperature in amine before desorber	101.2 °C
Number of stages in desorber	10
Murphree efficiency in desorber stages	0.5
Reflux ratio in desorber	0.3
Desorber pressure	2.0 bar
Reboiler temperature	120 °C
Reboiler Power (only excess heat)	24.5 MW
Pressure increase across Lean amine pump	3 bar
Pump efficiency	0.75
ΔT_{min} in Lean/Rich heat exchanger	10 °C

Murphree efficiencies for CO_2 in the absorption column stages are specified; efficiency is constant at 0.21 for the first five stages and then decreases linearly down to 0.11 for stage 15 (Øi, 2012). Murphree efficiency for CO_2 in the desorption column is constant at 0.5. The Murphree efficiency for a stage is defined by the change in mole fraction CO_2 from a stage to another divided by the change on the assumption of equilibrium. Pumps were simulated with an adiabatic efficiency of 0.75.

Figure 3 shows the representation of the standard amine-based absorption desorption process in the simulation program Aspen HYSYS. The calculation sequence is similar to earlier works (Øi, 2007; Aromada and Øi, 2015). First the absorption column T-100 is calculated from the inlet gas and the lean amine (which is first guessed). The rich amine from the bottom of the absorption column passes through the pump P-100 and the main rich/lean heat exchanger E-102 and gains heat from the lean amine from the desorption column. The heated rich amine is entering the desorption column T-101 which calculates the hot lean amine leaving the desorption column. The hot lean amine leaving from bottom of desorber is being pumped to a higher pressure via lean amine pump P101 and passes through the lean/rich heat exchanger E-102 and is then further cooled in the lean cooler E-101. Then this lean amine is checked in a recycle block RCY-1. It is checked whether the recycled lean amine is sufficiently close to the earlier guessed lean amine stream, which may be changed by iteration. This is completing the loop.

2.2 Dimensioning and cost estimation calculations

2.2.1 Scope analysis

The cost analysis is limited to the equipment listed in the flow-sheet Figure 3 excluding the flue gas cooler. No pre-treatment like inlet gas purification or cooling is considered. And no treatment after stripping like compression, transport or storage of CO_2 is considered. The cost estimate is limited to installed cost of listed equipment. It does not include e.g. land procurement, preparation, service buildings or owners cost.

2.2.2 Dimensioning of equipment

The dimensions of the process equipment are estimated based on typical dimension factors. The absorption column diameter is based on a gas velocity of 2.5 m/s and the desorption column is based on a gas velocity of 1 m/s (Park and Øi, 2017). The packing height of the absorption and desorption column is 1 meter per stage with a specified stage efficiency. The total height of the absorption column and desorption column is assumed to be 40 m and 22 m respectively. The calculation of absorber height includes packing, liquid distributors, water wash, demister, gas inlet & outlet and sump while calculation of desorber height includes inlet for condenser, packing, liquid distributor, gas inlet and sump.

The heat transfer areas of the heat exchangers are calculated based on duties and temperature conditions obtained from simulations. Overall heat transfer coefficient values have been assumed, for lean/rich heat exchanger 500 W/(m^{2}K), lean amine cooler 800 W/(m^{2}K), reboiler 800 W/(m^{2}K) and condenser 1000 W/(m^{2}K) (Øi, 2012). Shell and tube heat exchangers were mainly considered for case studies but for one alternative study plate & frame heat exchanger was also considered.

Centrifugal pumps are selected for the rich amine and lean amine pump. Volumetric flow rate and pump power are required in order to calculate equipment cost for pumps, which is available from the simulations.
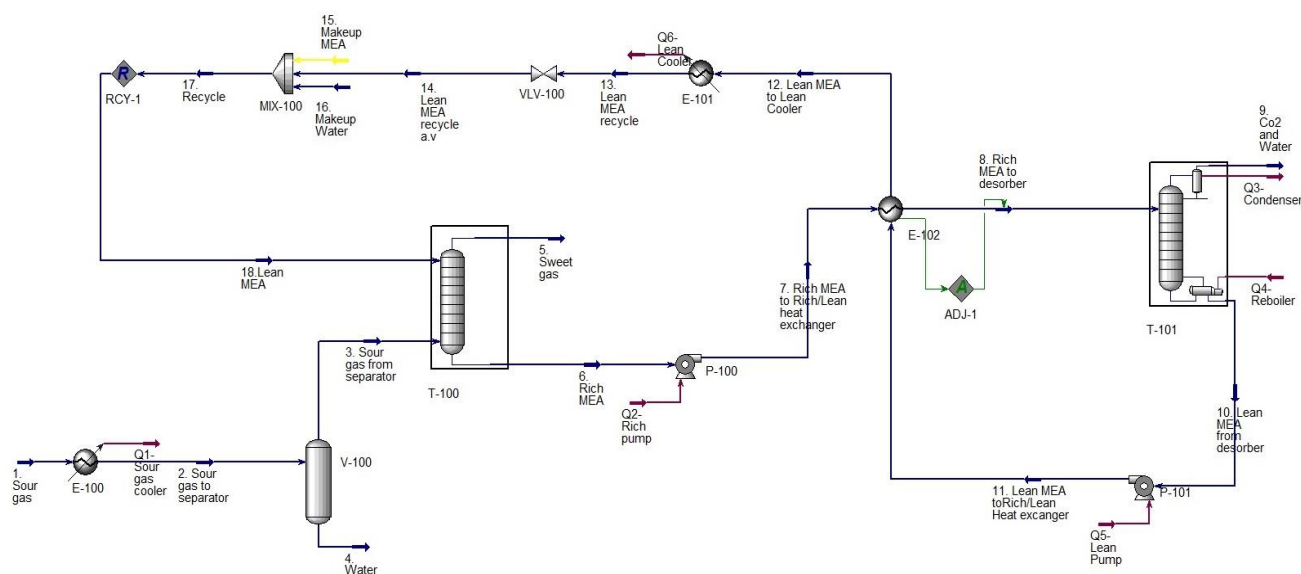

Figure 3. Aspen HYSYS flow-sheet of a standard amine-based CO_2 capture process

2.2.3 Capital cost estimation methods used

The equipment costs are taken from the Aspen In-plant Cost Estimator (v.10), which gives the cost in Euro (€) for Year 2016 (1st Quarter). A generic location that has good infrastructure and easy access to a workforce and materials, e.g. Rotterdam, is assumed. Stainless steel (SS316) with a material factor of 1.75 was assumed for all equipment units. To calculate capital cost, two methods were used.

In the detailed factor method, each equipment cost (in carbon steel) was multiplied with its individual installation factor to get equipment installed cost, as in earlier works (Øi, 2012; Park, 2016). The total capital cost was then calculated by adding all the individual equipment installed costs. The detailed installation factor is a function of the site description, equipment type, materials, size of equipment and includes direct costs (such as the costs for erection, instruments, civil, piping, electrical, insulation, steel and concrete), engineering costs, administration costs and the costs for commissioning and contingency. The updated installation factors for year 2016 (Eldrup, 2016) were used that decreases with increasing equipment cost. This cost estimate is expected to have an accuracy of ±40%.

In the Lang factor method (named after Hans J. Lang in 1947) the idea is to have overall installation factors, called Lang factors, depending upon the type of process plant. In this study, a Lang factor for a fluid process plant which is 4.74 (Turton et al, 2013) has been multiplied with the sum of all equipment costs to estimate the total capital cost.

2.2.4 Operational cost calculation

The electricity cost is set to 0.12 €/kWh. The cooling water cost is set to 0.02 €/m³, and the excess heat is specified to be free although the excess/waste heat always comes with a cost. The annual maintenance cost was set to 4 % of the equipment installed cost. Annual operator cost is added on basis of shift work (6 operators). One operator is assumed to cost 77000 €/year which includes salary as well as employer's expenses. The yearly operating time was 8000 hours, the calculation time was set to 25 years (2 years construction) and the interest was set to 7.5 %.

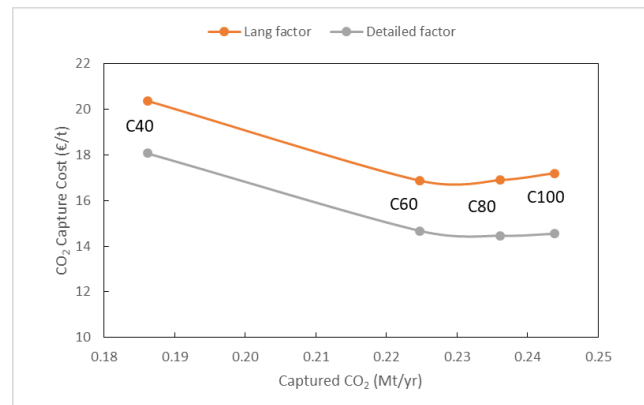### 2.2.5 Capture efficiency and cost calculation

The CO_2 capture efficiency is calculated using equation (1) and the CO_2 capture cost is calculated using equation (2) shown below.

$$Capture\ efficiency = \frac{CO_2\ sour\ gas - CO_2\ sweet\ gas}{CO_2\ sour\ gas} \times 100 \quad (1)$$

$$CO_2\ capture\ cost \left(\frac{€}{ton\ CO_2}\right) = \frac{CAPEX + OPEX\ (€/yr)}{CO_2\ captured\ (ton/yr)} \quad (2)$$
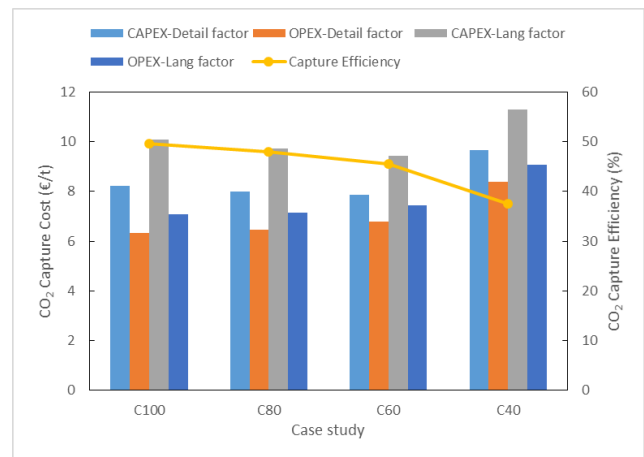
3 Results and Discussion

For the main four case studies of partial CO_2 capture, Figure 4 shows the plot between captured CO_2 from full flow (C100) to 40% flow (C40) and the cost of capture per ton CO_2. The lowest cost is obtained for C80 with the detailed factor method. The cost results for the Lang factor method has a higher cost per ton CO_2 captured than with the detailed factor method for all the cases. The reason for this is the fact that in the detailed factor method, each equipment gets different installation factor and when the installation factors for all the equipment are combined, that was found to be less than the Lang factor (4.74) used for this study.

Figure 4. CO_2 capture cost plotted against captured CO_2 for full-flow and part-flow case studies
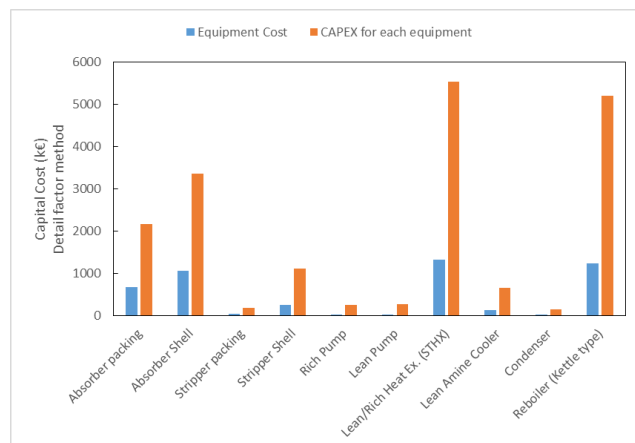
The curve in Figure 4 also indicates that the cost of CO_2 capture initially goes down when the amount of CO_2 capture decreases from 0.245 Mt/yr to around 0.23 Mt/yr but then the cost increases sharply as the captured amount decreases further.

Figure 5. Overall cost analysis of four case studies

Detailed cost analysis and capture efficiency for the main four case studies is shown in Figure 5. The CAPEX dominates in all the case studies. The best capture efficiency is for case study C100 but the capture efficiency does not fall down drastically from C100 to C60 (49.6 to 45.5%). While for C40, the efficiency falls down to 37% and this case study has also the highest
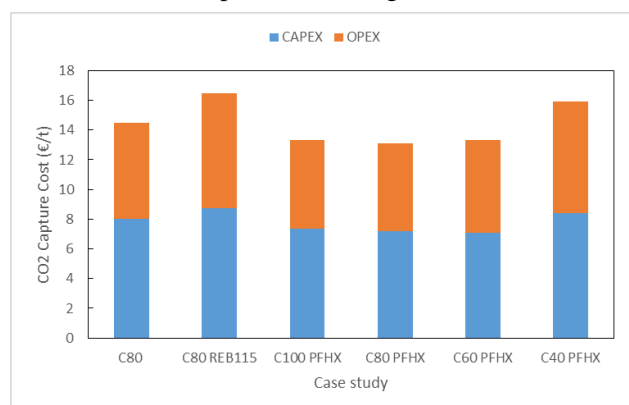
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capture cost as well. The energy optimum case study proves to be C100. The cost optimum case study when it comes to Lang factor is C60 (16.87 €/t) but capture cost of C80 (16.90 €/t) is not far away from the lowest. With the detailed factor method, the lowest capture cost comes for the case study C80 (14.46 €/t) while capture cost for C100 (14.54 €/t) is close to that of C80. Hence, the case study C80 with the detailed factor method is cost optimum and selected for further analysis.

Figure 6. Capital cost overview of case study C80 (STHX, Shell & tube heat exchanger)

Since CAPEX dominates the capture cost, it will be worthwhile to have a detailed look on the capital cost of case study C80 that helps in optimization, which is shown in Figure 6. There are four major equipment, lean/rich heat exchanger, reboiler, absorber shell and packing that are contributing significantly and the efforts should be directed to reduce this cost.

An alternative to reduce the lean/rich heat exchanger capital cost is to replace the shell and tube heat exchanger (STHX) with a plate and frame heat exchanger (PFHX) (Marcano, 2015). That has also been performed for all the case studies, with the name PFHX and the results are presented in Figure 7.

Figure 7. Cost overview of alternatives with PFHX, Plate & frame heat exchanger and with lower reboiler duty

The results clearly indicates that by replacing shell & tube heat exchanger with plate & frame heat exchanger (for lean/rich heat exchanger), the capture cost further

decreases for all the cases. The lowest capture cost in this scenario remains to be case study C80 that decreased from 14.46 €/t (with STHX) to 13.11 €/t (with PFHX).

In another alternative on case study C80, the reboiler temperature has been decreased from 120 °C to 115 °C. By doing this, more excess heat can be available and it might help in reducing the capture cost. The results of this new case study C80 REB115 is also presented in Figure 7. For this case study, excess heat was increased to 25.1 MW since we can utilize further excess heat of 5 °C from hot exhaust gas. Table 3 contains some important input parameters and outputs for case studies C80 and C80 REB115.

Table 3. Input Parameters and results for case study C80 and C80REB115

Parameter	Unit	Case Study	
		C80	C80REB115
Flue gas flow rate	Kmol/h	4630	4630
Excess heat to reboiler	MW	24.5	25.1
Lean MEA flow rate	kg/h	535000	845900
Lean loading		0.26	0.35
Rich loading		0.51	0.50
CO_2 capture efficiency	%	47.9	46.3
CO_2 removed per year		0.236	0.228
Reboiler energy demand	MJ/kg CO_2	3.27	3.47

The results in Figure 7 shows that the capture cost has increased from 14.46 €/t to 16.46 €/t for case study C80 REB115 even though the excess heat has been increased. Besides the capture cost, reboiler energy demand has also increased for this lower reboiler temperature case study, while the capture efficiency and CO_2 removed per year decreases as shown in Table 3.

3.1 Sensitivity analysis

The sensitivity analysis has been performed on capital cost, specifically on the installation factors of the four most costly equipment identified i.e., lean/rich heat exchanger, reboiler, absorber shell and packing. Installation factors for these equipment have been decreased by 50% to see the impact they have on capture cost of main four case studies.

Another analysis has been performed on civil installation sub-factor. This sub-factor of the detailed installation factor is expected to cover additional cost due to equipment cost (and size). This sub-factor has also been decreased by 50% for all the equipment installation factors and its effect on capture cost has been analysed.

The results are presented in Table 4, which shows that by decreasing the installation factors for absorber packing, the full flow case C100 becomes the cost optimum case although the lowest cost 12.82 €/t is achieved for case C80 when installation factor for lean/rich heat exchanger is reduced. For all other scenarios, case C80 continues to give lowest cost per ton

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when the installation factor or civil sub-factor is decreased by 50%. The greatest impact on capture cost is by the lean/rich heat exchanger and the reboiler, the capture cost goes down significantly from 1.4 – 2.5 €/t for all the cases. The lowest impact is by the civil sub-factor where the capture cost decreases by only 0.13 €/t for cases C100 to C60 apart from for the C40 case where the increase is 0.96 €/t.

Table 4. Effect of installation factors (IF) and civil sub-factor (factors decreased by 50%) on capture cost

Case study	C100	C80	C60	C40
Capture Cost, €/t	14.54	14.46	14.67	18.06
IF-Abs. Packing, €/t	13.77	13.82	14.17	17.66
IF-Abs. Shell, €/t	13.47	13.46	13.79	17.15
IF-Reboiler, €/t	13.07	12.92	13.05	16.10
IF-l/r heat exch., €/t	12.86	12.82	13.02	15.48
Civil sub-factor, €/t	14.41	14.33	14.54	17.10

In a more detailed analysis for cost optimization, the number of stages in the absorber should be optimized but this is not included in the scope of this study.

3.2   Comparisons with earlier work

(Dong et al, 2012) calculated that it was possible to capture 78 % CO_2 in a cement case under other conditions. The amount captured was dependent on the degree of integration. (Park, 2016) concluded that the lowest total cost per ton CO_2 captured was calculated for the standard full-flow process with 5 absorption stages. This conclusion was however based on the assumption that transport and treating of the gas before or after CO_2 capture was not considered. (Øi et al, 2017) worked on partial capture from flue gas of cement industry and concluded that the energy optimum case and the lowest total cost per ton CO_2 captured was calculated for the standard full-flow process with a low number of absorption stages.

4   Conclusion

Different case studies from full flow of the flue gas from String 1 to part flow for partial CO_2 capture in a cement industry were simulated with only excess heat using the process simulation tool Aspen HYSYS. These case studies were cost estimated using the Aspen In-plant cost estimator along with two cost estimation methods i.e., detailed factor method and Lang factor method.

The highest CO_2 removal efficiency is obtained for the full flow alternative which is regarded as the energy optimum process with a reboiler energy demand around 3.2 MJ/kg CO_2. The cost optimum case was with 60% of the flue gas flow into the capture plant, when the Lang factor method was used. When using the detailed factor method, the case with 80% of the flue gas flow is the cost optimum alternative. This is valid for all the different case studies performed via detailed factor method with the exception when the installation factor for absorber packing was decreased, the full flow

alternative becomes the cost optimum. This clearly shows that the selection of the cost estimation method and the assumptions made have a great impact on the results.

The greatest impact on capture cost was by the capital cost, specifically by the lean/rich heat exchanger, reboiler, absorber shell and packing. The capture cost can be reduced by selecting a plate and frame heat exchanger as the lean/rich heat exchanger.

Acknowledgments

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