Simulation and Cost Optimization of different Heat Exchangers for CO_2 Capture

Solomon Aforkoghene Aromada[1], Nils Henrik Eldrup[1,2], Fredrik Normann[3], Lars Erik Øi[1]

[1]Department of Process, Energy and Environmental Technology, University of South-Eastern Norway
[2]SINTEF Tel-Tek, SINTEF Industri, Kjølnes Ring 30, 3918, Porsgrunn, Norway
[3]Chalmers University of Technology, 412 96, Göteborg, Sweden
Corresponding author's email address: solomon.a.aromada@usn.no; saromada@gmail.com

Abstract

The industrial deployment of amine-based CO_2 capture technology requires large investments as well as extensive energy supply for desorption. Therefore, the need for efficient cost and economic analysis aimed at CO_2 capture investment and operating costs is imperative. Aspen HYSYS simulations of an 85% CO_2 absorption and desorption process for flue gas from cement industry, followed by cost estimation have been performed. This is to study the cost implications of different plants options. Each plant option has a different lean/rich heat exchanger type. Cost optimisation of the different heat exchangers is also done in this work. Three different shell and tube and two plate and frame heat exchangers have been examined. The minimum CO_2 capture cost of €57.9/ton CO_2 is obtained for a capture plant option having a gasketed-plate heat exchanger with ΔT_{min} of 5 ℃ as the lean/rich heat exchanger. The use of plate and frame heat exchangers will result in considerable CO_2 capture cost reduction.

Key words: simulation, CO2, CCS, heat exchanger, shell and tube, Aspen HYSYS, plate heat exchanger

1 Introduction

There has been increased public concern for mitigation of global warming, which is largely caused by emissions of carbon dioxide (CO_2). Carbon capture and storage (CCS) is generally recognised as an urgent mitigation measure (Rubin et al., 2013). The amine-based post-combustion CO_2 capture technology is the most matured and promising technology option **(**Nwaoha, 2018). However, its industrial deployment requires large investments as well as enormous energy supply for desorption (Lim et al., 2013; Aromada and Øi, 2017). Therefore, the need for efficient cost and economic analysis aimed at reduced CO_2 capture investment and operating costs is imperative.

The lean/rich heat exchanger is one of the most expensive equipment in an amine-based CO_2 capture plant, and it has a considerable cost implication on the investment (Ali et al., 2019).

In preliminary cost estimation of heat exchangers, the important design parameter is the heat transfer area needed. That is evaluated from the heat duty (heat transfer from hot to cold stream), overall heat transfer coefficient, and the log-mean temperature difference (LMTD) (van der Spek et al., 2019). However, the required heat duty depends on the minimum approach temperature (ΔT_{min}).

In post-combustion solvent-based CO_2 capture, studies on cost optimisation of the lean/rich MEA heat exchanger have been based on ΔT_{min} of the shell and tube heat exchanger (STHX) types (Kallevik, 2010; Øi et al., 2014; Aromada and Øi, 2017; Ali et al., 2019). None of such studies has been found for the plate and frame heat exchanger (PHE). Thus, this study is conducted on cost optimisation of the PHE based on ΔT_{min}. This is carried out by performing process simulations of CO_2 absorption and desorption process. Cost estimation and optimisation to find the most cost effective and technically suitable type of heat exchanger for the lean/rich heat exchanger is then carried out.

1.1 Process Description and Scope

The process comprises a flue gas fan for transporting the flue gas through the direct contact cooler (DCC) where the temperature is reduced. The DCC pump and DCC cooler help in circulation and cooling of the water respectively. The main capture process consists of an absorber, a desorber with a reboiler at the bottom and a condenser, lean/rich heat exchanger, pumps and a cooler. *Figure 1* shows the flowsheet of the standard capture process.

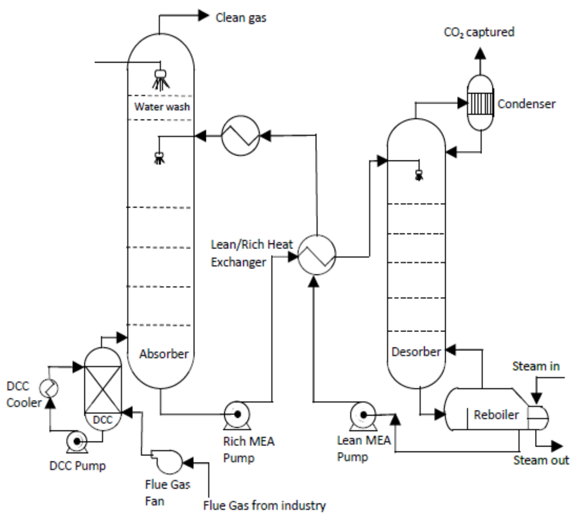

Figure 1. . Flowsheet of the standard process

2 Simulation, Specifications and Assumptions

2.1 Specifications for Simulation

Table 1 presents the specifications used for the base case simulations. The flue gas data are from a cement industry and are taken from (Onarheim et al., 2015; Ali et al., 2019).

2.2 Process Simulation

Aspen HYSYS Version 10 is used for the simulations with the same calculation approach as in (Øi, 2007; Aromada and Øi, 2015). The difference is that in version 10, the acid gas property package replaces the Amine property package in previous versions.

Table 1. Specifications for simulation (Onarheim et al., 2015; Ali et al., 2019)

Specifications	
Flue gas	
Temperature [°C]	80
Pressure [kPa]	121
CO_2 mole-fraction	0.2520
H_2O mole-fraction	0.0910
N_2 mole-fraction	0.5865
O_2 mole-fraction	0.0705
Molar flow rate [kmol/h]	11472
Flue gas from from DCC to absorber	
Temperature [°C]	40
Pressure [kPa]	121
Lean MEA	
Temperature	40
Pressure [kPa]	121
Molar flow rate [kmol/h]	96850
Mass fraction of MEA [%]	29
Mass fraction of CO_2 [%]	5.30
Absorber	
No. of absorber stages	15
Absorber Murphree efficiency [%]	11- 21
ΔT_{min}, lean/rich heat exchanger [°C]	10
Desorber	
Number of stages	10
Desorber Murphree efficiency [%]	100
Pressure [kPa]	200
Reboiler temperature [°C]	120
Reflux ratio in the desorber	0.3
Temperature into desorber [°C]	104.6

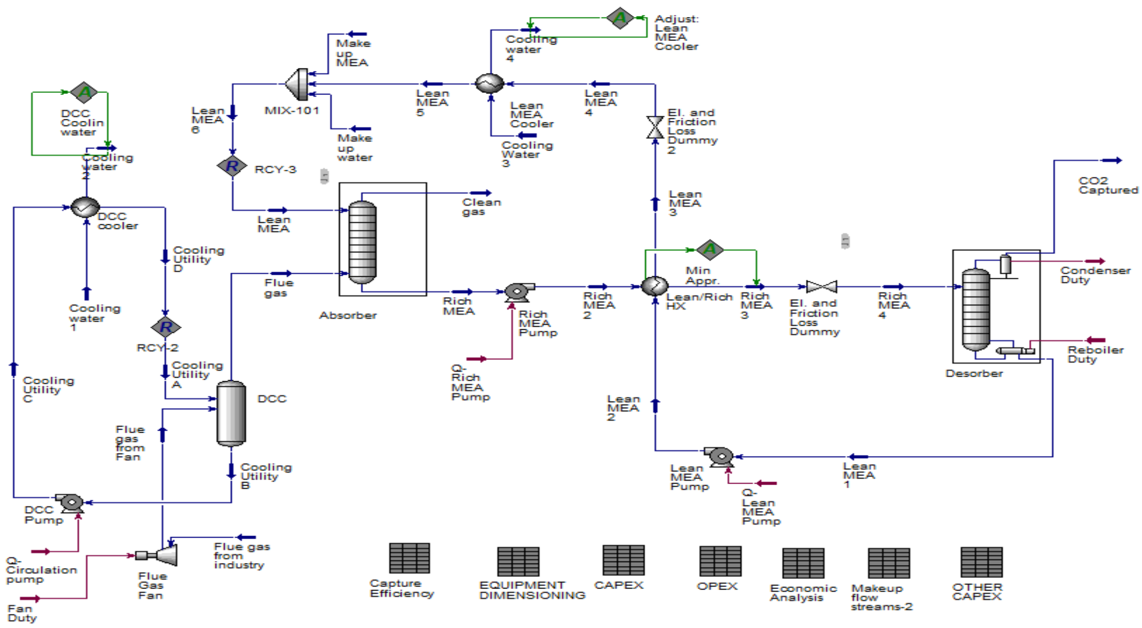

Figure 2. Aspen HYSYS flowsheet

The absorption column as well as the desorption columns are both simulated as equilibrium stages with stage efficiencies. The absorber is simulated with 15 packing stages, while it is 10 packing stages for the desorber. Murphree efficiencies for CO_2 are specified in the simulation. For more details on Murphree efficiency, see (Øi, 2007). Equilibrium stages of 1 m height each for both columns are assumed. Murphree efficiencies of 11 – 21% were specified from bottom to the top of the absorption column (Ali et al., 2019). A constant Murphree efficiency of 100% is specified for all the stages of the desorption column. The Modified HYSIM Inside-Out algorithm was selected in the columns because it helps to improve convergence (Aromada and Øi, 2015).

Adiabatic efficiency of 75% was specified for all the pumps and the flue gas fan. The Aspen HYSYS simulation process flow diagram (PFD) is given in *Figure 2*.

3 Methods

3.1 Scope of the Cost Estimates

The equipment included in this cost analysis are for cooling the flue gas before it enters the absorption column, and for the absorption and desorption process as can be seen in *Figure 1* and *Figure 2*. The study does not include equipment for pre-treatment unit of the flue gas and water-wash section. The equipment for CO_2 compression are not considered because the focus is on the lean/rich heat exchanger

The total investment cost in this study is limited to the sum of the installed costs of the equipment considered. The cost of acquiring the site (land), preparing the site and for service buildings are not included.

The operating and maintenance costs (OPEX) include the cost of electricity, steam, cooling and process water, solvent (MEA), salaries of 6 operators and 1 engineer, and annual maintenance cost set at 4 % of the installed cost of the equipment as given in *Table 2*.

Table 2. Operating cost data

	Unit	Value/unit*	Reference
Steam	€/kWh	0.032	Husebye et al. (2012)
Electricity	€/kWh	0.132	Ali et al. (2019)
Water	€/m^3	0.022	Ali et al. (2019)
MEA	€/m^3	2 069	Ali et al. (2019)
Maintenance	€	4% of CAPEX	Ali et al. (2019)
Operator	€	85 350 (x6)	Ali et al. (2019)
Engineer	€	166 400	Ali et al. (2019)

**The costs have been escalated to January 2020*

Costs for CO_2 transport and storage, pre-production costs, insurance, taxes, first fill cost and administrative costs are not included in the OPEX.

3.2 Equipment Dimensioning and Assumptions

Dimensioning of equipment in this study follows the approach used in Ali *et al.* (2019) based on mass conservation and energy balances of the system. *Table 3* summarises the dimensioning factors and assumptions used in this work.

Table 3. Equipment dimensioning factors and assumptions

Equipment	Sizing factors	Basis/Assumptions
DCC Unit	Tangent-to-tangent height (TT), Packing height, internal and external diameters (all in [m])	Velocity using Souders-Brown equation with a k-factor of 0.15 m/s (Yu, 2014, pp. 97). TT =15 m, 1 m packing height/stage (4 stages)
Absorber		Superficial velocity of 2 m/s, TT=40 m, 1 m packing height/stage (15 stages)
Desorber		Superficial velocity of 2 m/s, TT=22 m, 1 m packing height/stage (10 stages)
Lean/rich heat exchanger	Heat transfer area, A [m^2]	U = 0.5 kW/m^2.K (Ali et al., 2019)
Reboiler		U = 0.8 kW/m^2.K (Ali et al., 2019)
Condenser		U = 1.0 kW/m^2.K (Ali et al., 2019)
Coolers		U = 0.8 kW/m^2.K (Ali et al., 2019)
Pumps	Flow rate [l/s]	Centrifugal
Flue gas fans	Flow rate [m^3/h]	Centrifugal

3.3 Cost Estimation and Assumptions

The Enhanced Detailed Factor (EDF) method is used for estimation of all the equipment costs and overall plant investment cost. Readers are referred to Ali *et al.* (2019) for the details and application of the EDF method.

The purchased costs of the equipment are obtained from Aspen In-plant Cost Estimator Version 11 with a cost year of 2018 (January). The costs are then escalated to January 2020 using the SSB (Norwegian Statistisk sentralbyrå, webpage) industrial cost index (2018 = 106; 2020 = 111.3). The currency conversion rate for Euro to NOK is 10.13, taken from (NorgesBank, 2020 webpage) on January 25, 2020. Conversion to NOK is necessary to use the enhanced factors developed by Nils Eldrup (Ali et al., 2019). The default location is Rotterdam in Netherlands.

All equipment is assumed to be made from stainless steel (SS316), except the Flue gas fan, which is from carbon steel (CS). Material factor to convert costs in SS316 to CS is 1.75 and 1.30 for seamless and welded equipment respectively.

A brownfield, and an Nth-of-a-kind (NOAK) project are assumed. 25 years of project, of which 2 years are for plant construction, and 7.5% interest rate is also assumed (Ali et al., 2019).

4 Results and Discussion

4.1 Simulation Results

Table 4 presents the process simulation results for the base case and sensitivity analysis of $\Delta T_{\min}$. Lower $\Delta T_{\min}$ give lower reboiler heat and lower lean MEA cooler duty (more heat has been transferred from the lean stream to the rich stream). Therefore, less steam and less cooling water are required in the reboiler and lean MEA cooler respectively.

Table 4. Simulation results

ΔT_{min}	Reboiler heat	Typical results	Lean MEA cooler duty
[°C]	[GJ/ton CO_2]		[kW]
5	3.83		66 389
10	4.08	3.3 - 5.0	81 333
15	4.27	(Nwaoha	89 333
20	4.67	et al.,2018)	117 778

4.2 Base Case Plant Investment Cost

The base case in this study has a U-tube shell and tube heat exchanger. The total investment cost (CAPEX) which is the sum of the installed costs of all the equipment is €97.5 million. The cost estimation results obtained show the same trends with similar studies by Ali et al. (2018) and Ali et al. (2019). The lean/rich heat exchanger contributes most to the total investment cost compared to other equipment as in *Figure 3*.

The heat exchanger accounts for 41% of the total capital cost (*Figure 4*). Ali et al. (2019) also calculated the lean/rich heat exchanger to have the highest installed cost for the same scope as in this study. It accounts for 37% of the CO_2 capture plant. They obtained their cost data from Aspen In-plant Cost Estimator V10 with a cost data year of 2016. Aspen In-plant Cost Estimator V11 with a cost data year 2018 is used in this study. In addition, the cost in this study are escalated from 2018 to 2020. This explain the 3% difference from a similar process. In the work of Nwaoha *et al.* (2018), for a process with an absorber packing height and diameter of 21.95 m and 10.07 m respectively, the lean/rich heat exchanger has the second highest cost for both MEA and AMP-PZ-MEA systems. The absorber in their case has the highest cost. The diameter is almost twice and the packing height is approximately 7 m higher than in this work. The study was for a 90% CO_2 capture process from a cement plant flue gas with 0.115 mole of CO_2. In this study, capture efficiency is 85% and the CO_2 molar composition is 0.252.

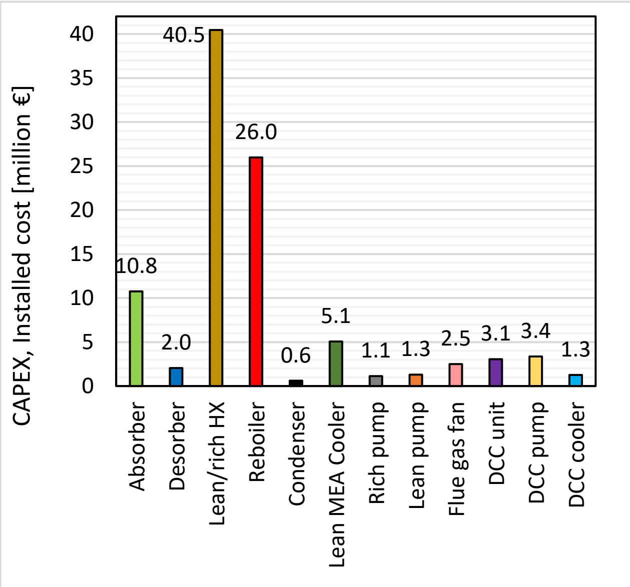

Figure 3. Equipment installed costs of the base case 85% CO_2 capture plant

4.3 Operation and Maintenance Costs

The annual operation and maintenance (O&M) cost for the base case is €44.5 million. Only the steam consumption costs €31.7 million and annual maintenance cost is €3.9 million.

4.4 Annualised CAPEX, Total Annual Cost and Capture Cost

Annualised capital cost is obtained from the following relation:

$$Annualised\ CAPEX = \frac{CAPEX}{Annualised\ factor} \qquad (1)$$

The annualised factor is calculated as follows:

$$Annualised\ factor = \sum_{i=1}^{23} \left[\frac{1}{(1+r)^n}\right] \quad (2)$$

Where n represents operational years and r is discount/interest rate. The annualised CAPEX for the base case is evaluated to be €9 million (CO_2 compression equipment not included). Thus, the total annual cost, which is the sum of the annualised CAPEX and the yearly OPEX, is €53.6 million. Figures 4 presents the annual cost distribution. The CO_2 capture cost is estimated from:

$$CO_2\ capture\ cost = \frac{Total\ annual\ cost}{Mass\ of\ CO_2\ Captured} \quad (3)$$

The CO_2 capture cost for the base case is 61.9 €/ton CO_2 (2020). In the literature, it is between €50/ton CO_2 – 128/ton CO_2 (Ali et al., 2019). (Ali et al., 2019) calculated this cost for a similar process but with the compression section to be €62.5/ton CO_2 for a cost year of 2016. For a full process that include compression, Nwaoha et al. (2018) calculated this cost for 90% CO_2 capture from a cement plant flue gas with CO_2 compression to be US$93.2/ton CO_2 (i.e., €74.5/ton CO_2). According to Irlam (2017), for a first-of-a-kind (FOAK) CSS complete technology, the CO_2 avoided cost for the cement industry is US$188 (€164.4) and US$130 (€113.7) per ton CO_2 for Germany and Poland respectively. FOAK technologies usually cost between 15 – 55% more than NOAK (Boldon & Sabharwall, 2014).

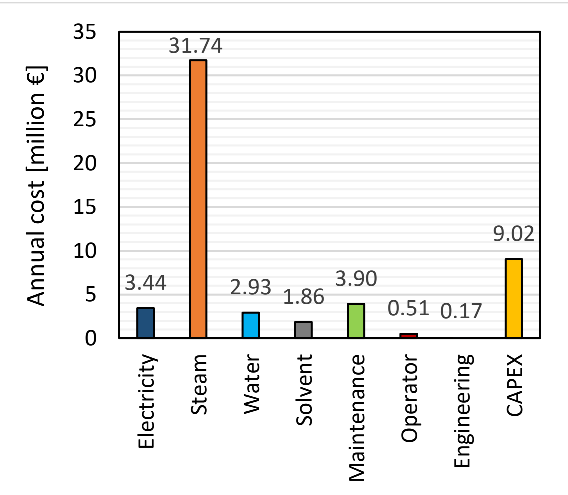

Figure 4. Cost distribution of the base case total annual cost

(Carbon Capture & Storage Association, 2011-2020) states that the capture cost range is €60/ton CO_2 – €90/ton CO_2 for the power industry. They projected that it will reduce considerably to €35 – 50/ton CO_2 in the beginning of 2020. Based on *Figure 4*, this reduction will have to come from reducing mainly the cost of steam. This can be achieved using available waste heat to generate steam or very cheap steam for desorption (Ali et al., 2018). Electricity cost is low in this study compared to Nwaoha et al. (2018) and Ali et al. (2019). This is because CO_2 compression is not considered in this work. The compressors require much more electrical energy compared to pumps and fan/blower.

4.5 CAPEX Based on Different Heat Exchangers

Figure 5 presents the total installed cost of CO_2 capture plant options of using the different types of heat exchangers. The compact heat exchangers offer considerable lower total investment cost compared to the conventional shell and tube heat exchangers. Using the gasketed-plate heat exchanger (G-PHE) will give the lowest plant investment cost. The purchase cost of the welded-plate heat exchanger (W-PHE) was assumed to be 30 % more expensive than the G-PHE based on information from Peters et al. (2004).

The reference case, which has U-tube shell and tube heat exchanger (UT-STHX) has investment cost of €97.5 million. The case with fixed tube sheet heat exchanger (FTS-STHX) has a CAPEX of €102.4 million. The installed cost of the plant with G-PHE is €72.6 million. The plant option with floating-head shell and tube heat exchanger (FH-STHX) gives the highest installed cost of €103.8 million.

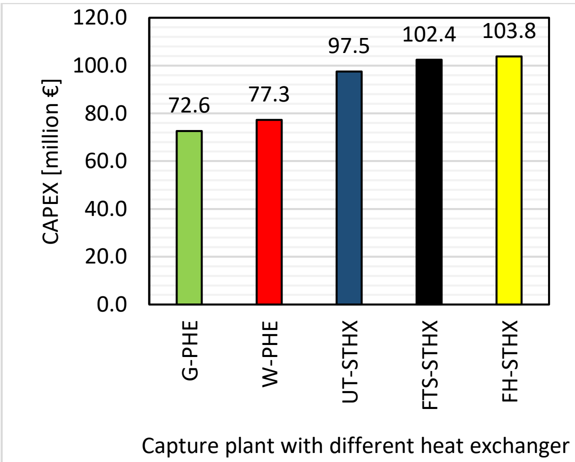

Figure 5. Total plant installed costs for different heat exchangers

4.6 Optimisation: Minimum Approach Temperature

Cost optimisation of the lean/rich heat exchanger in this study is done by finding the cost optimum minimum approach temperature (ΔT_{min}).

The plants with G-PHE and welded-plate heat exchanger (W-PHE) have their minimum CAPEX at 15°C, while it is 20°C for the 4 STHXs. As the ΔT_{min} increases, the heat transfer area is reduced, thereby reducing the CAPEX since the lean/rich heat exchanger with STHXs account for 41 – 45 % of the CAPEX in this study. The slight increase of CAPEX from 15 – 20°C as can be observed in Figure 6 for the PHEs is caused by increase in the cost of other equipment like the lean MEA cooler and the reboiler. This will also result in higher OPEX, especially from higher steam consumption as can be seen in *Table 3* and *Figure 7*. More cooling water is also needed. However, increase in OPEX is slight from 5 – 15°C for the STHXs but becomes significantly steep from 15 – 20°C. That is the same for the PHEs except that the OPEX is considerably lower at 5°C compared to 10°C.

In order to find the optimum design ΔT_{min} we evaluated the CO_2 capture cost at the different ΔT_{min} for the different heat exchanger options. **Figure *8*** presents the results.

The STHXs and W-PHE have their optimum CO_2 capture costs at 15°C. While the G-PHE optimum cost is at 5°C, which is due to its relative lower cost per heat transfer area and lower maintenance cost. Cost savings of €1.6/tCO_2, €1.1/tCO_2 and €1.0/tCO_2 are achieved by the cost optimum cases with U-tube, fixed tube-sheets and floating-head shell and tube heat exchangers when compared with the base case. The cost optimum cases with gasketed and welded plate heat exchangers have a cost savings of €4.0/tCO_2 and €3.4/tCO_2 respectively.

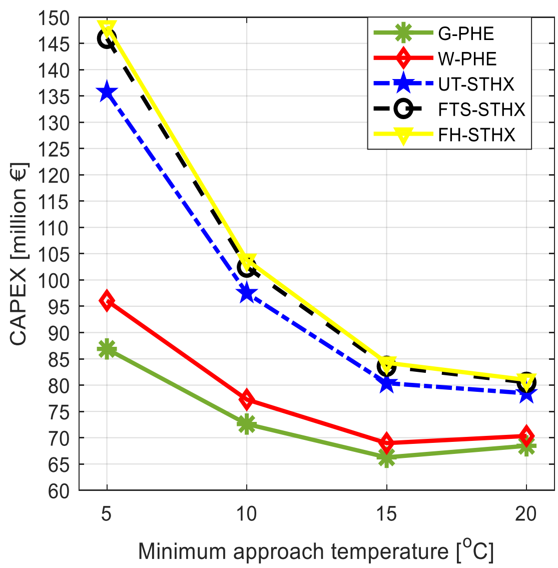

Figure 6. CAPEX of the different heat exchangers at different ΔT_{min}

All the studies of optimum ΔT_{min} we found of solvent-based CO_2 capture used STHXs (Kallevik, 2010; Øi et al. (2014), Li et al., 2016; Aromada & Øi, 2017; Nwaoha et al., 2018; Ali et al., 2019). None of such studies was found for other types of heat exchangers like PHE. This is because in the chemical industry, about 60% of heat exchangers in use are STHX (Peters et al., 2004). They are more robust, they can be applied in all types of processes, they can withstand higher pressures, higher temperatures and thermal stresses, and higher pressure difference between the hot and cold streams.

Additional advantage of the STHX is that they have well-established design codes, standards and specifications, especially by TEMA (Tubular Exchanger Manufacturers Association) and American Society for Mechanical Engineers (ASME). The PHEs do not have such established or accepted design standards. Therefore, higher design uncertainties are expected for the PHEs.

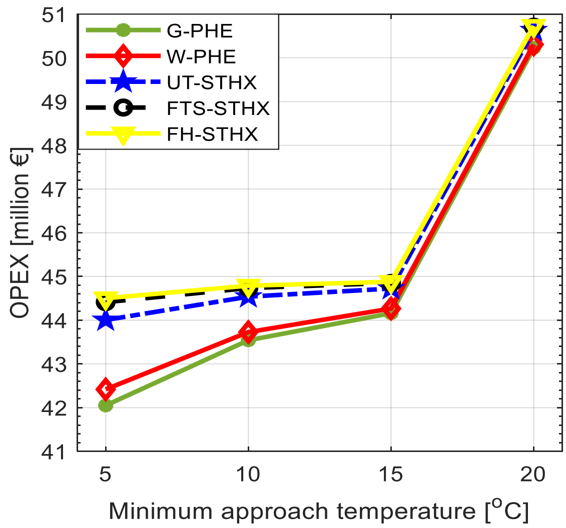

Figure 7. OPEX of the different heat exchangers at different ΔT_{min}

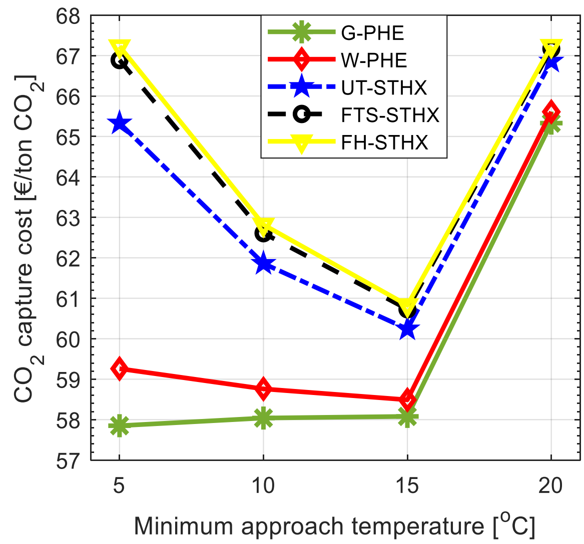

Figure 8. Cost optimum ΔT_{min} of the different heat exchangers

Nonetheless, the plate and frame heat exchangers are increasingly being considered for application in the process industry (Peters et al., 2004). This is because capital-intensive processes need heat exchangers that can achieve higher thermal efficiencies and simultaneously reducing equipment/investment costs (Peters et al., 2004). The PHEs also occupy less space and have less weight for the same heat transfer area as STHX.

4.7 Maintenance

Maintenance of the G-PHE is easier and far less expensive. The plates are removable and can easily be cleaned mechanically. Thus, it is the most ecological option (Marcano, 2015). The parts can easily be replaced relatively inexpensively.

The plates of the W-PHE are welded and thus are not removable. Consequently, cleaning can only be done by chemical means. The only advantage over G-PHE is that W-PHE can withstand higher pressures and temperatures. This advantage is not relevant in CO_2 solvent-based absorption and desorption systems since the pressures and temperatures are relatively low

FTS-STHX and FH-STHX are cleaned both mechanically (inside the tubes) and chemically (outside surfaces of tube). The UT-STHX normally requires only chemical cleaning because of the U-tube shape of the tubes.

Therefore, the G-PHE which require less space is the most ecologically friendly option, and it is the easiest and the cheapest to maintain among the heat exchangers investigated.

4.8 Maintenance and Operating Cost Discussion

Figure 7 presents the OPEX calculated in this study. At ΔT_{min} of 5 – 10℃, the calculated annual OPEX of the PHEs is considerably less than that of any of the STHXs, even though the difference in this study is only based on maintenance cost. The gap gets closer at 15℃ and and very close at 20℃. This is because the purchase and installed costs of the STHXs reduce drastically at ΔT_{min} 15 and 20℃.

4.9 Comparison with Previous Studies

In this section, comparison of this study is done with some previous studies. All the previous studies used the STHX.

(Øi et al., 2014) calculated the cost optimum ΔT_{min} of a plant with 16 absorber packing stages, and 20 years period with discount rate of 10.5% to be 12°C.

(Aromada and Øi, 2017) estimated it to be 13°C for a system with 15 absorber packing stages with discount rate of 7% for 20 operational years, based on negative-NPV method. When the years of plant operation were reduced to 15 years, cost optimum ΔT_{min} became 14°C.

(Kallevik, 2010), also applied negative NPV for 20 years calculation period, with 7% discount rate, estimated the cost optimum ΔT_{min} to fall within 10 – 14°C, for a 85% CO_2 capture with 15 absorber packing stages.

Most recent is (Ali et al., 2019), for a calculation period of 24 years and interest rate of 7.5%, evaluated the cost optimum ΔT_{min} to be 10°C.

These results suggest that the cost optimum ΔT_{min} for the STHXs is within 10 – 16°C, which are in agreement with this study. The little differences obtained from the different studies occur due to the different sources of cost data and economic assumptions like interest rates and operational years.

Several technical studies have also shown that operating at 5℃ ΔT_{min} will help in reduction of the reboiler heat in CO_2 capture processes. However, the capital cost of achieving this makes it not to be the cost optimum design parameter for the well-established STHX. This study suggests that ΔT_{min} of 5℃ or between 5 – 10℃ can be energy optimum and cost optimum design if G-PHE is used.

5 Conclusion

Simulations of 85% CO_2 absorption and desorption process aimed at cost optimisation of the lean/rich heat exchanger has been performed using Aspen HYSYS Version 10. This was followed by cost estimation and optimisation of the lean/rich heat exchanger by finding the type of heat exchanger and the design optimum ΔT_{min} among 5, 10, 15 and 20℃ ΔT_{min}. Considerable savings in capital and operating costs can be achieved by selecting the plate and frame heat exchanger instead of the conventional shell and tube types, in a CO_2 absorption and desorption plant design. The PHEs require only 30, 15, 9, and 6 number of units for the cases of 5, 10, 15 and 20℃ ΔT_{min} respectively, compared to 44, 23, 13, 9 number of units respectively for the STHXs. The G-PHE gives the lowest total annual cost in all the ΔT_{min}. G-PHE with 5℃ ΔT_{min} is calculated to be the energy optimum and the cost optimum design for the lean/rich heat exchanger.

Abbreviations

PHE: Plate and frame heat exchanger
G-PHE: Gasketed-plate or plate and frame heat exchanger
W-PHE: Welded- plate heat exchanger
STHX: Shell and tube heat exchanger
UT-STHX: U-tube shell and tube heat exchanger
FTS-STHX: Fixed-tube sheet Shell and tube heat exchanger
FH-STHX: Floating head shell and tube heat exchanger

References

H. Ali, N.H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi. Cost Estimation of CO_2 Absorption Plants for CO_2 Mitigation–Method and Assumptions. *International Journal of Greenhouse Gas Control*, 88, 10-23, 2019. doi: 10.1016/j.ijggc.2019.05.028

H. Ali, L. E. Øi, and N. H. Eldrup. Simulation and Economic Optimization of Amine-based CO_2 Capture using Excess Heat at a Cement Plant *Linköping University Electronic Press Conference Proceedings (SIMS 59)*, 153:58-64, 2018. doi:10.3384/ecp1815358

S. A. Aromada and L. E. Øi. Simulation of Improved Absorption Configurations for CO_2 Capture. *Linköping Electronic Press Conference Proceedings (SIMS 56)*, 119(2):21-29, 2015. doi: 10.3384/ecp1511921

S. A. Aromada and L. E. Øi. Energy and Economic Analysis of Improved Absorption Configurations for CO_2 Capture. *Energy Procedia*, 114: 1342-1351, 2017. doi: 10.1016/j.egypro.2017.03.1900

L. M. Boldon and P. Sabharwall. *Small modular reactor: First-of-a-Kind (FOAK) and Nth-of-a-Kind (NOAK) Economic Analysis* (No. INL/EXT-14-32616). Idaho National Lab. (INL), Idaho Falls, ID (United States), 2014. doi: 10.2172/1167545

Carbon Capture and Storage Association (Webpage). *Affordability, CCS: Keeping the lights on without costing the earth.* 2011 – 2020. Available: http://www.ccsassociation.org/why-ccs/affordability/ Accessed 07.05.2020

J. Husebye, A. Brunsvold, S. Roussanaly and X. Zhang. Techno economic evaluation of amine based CO_2 capture: impact of CO_2 concentration and steam supply. *Energy Procedia*, 23, 381 – 390, 2012. doi: 10.1016/j.egypro.2012.06.053

L. Irlam. *Global costs of carbon capture and storage*. Global CCS Institute, Melbourne, Australia, 2017.

O. B. Kallevik. *Cost estimation of CO_2 removal in HYSYS*. Master's thesis. Høgskolen i Telemark, 2010.

Y. Lim, J. Kim, J. Jung, C. S. Lee, and C. Han. Modeling and simulation of CO_2 capture process for coal-based power plant using amine solvent in South Korea. *Energy Procedia*, 37:1855-1862, 2013. doi.org/10.1016/j.egypro.2013.06.065

L. A. Marcano. *Design and evaluation of heat exchangers used in post-combustion CO_2 capture plants*. Master's Thesis. Telemark University College, Porsgrunn, 2015.

Norgesbank webpage. Available on: https://www.norges-bank.no Accessed: 25.01.2020.

C. Nwaoha, M. Beaulieu, P. Tontiwachwuthikul and M. D. Gibson. Techno-economic analysis of CO_2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend. *International Journal of Greenhouse Gas Control*, 78:400-412, 2018. doi.org/10.1016/j.ijggc.2018.07.015

K. Onarheim, S.Ò. Garðarsdòttir, A. Mathisen, L. O. Nord and D. Berstad. *Industrial implementation of carbon capture in Nordic industry sectors*. Nordic CCS Competence Centre NORDICCS 2015.

M. S. Peters, K. D. Timmerhaus, and R. E. West. *Plant Design and Economics for Chemical Engineers*, 5th edition, McGraw-Hill Companies, Inc. Singapore 2004.

E. S. Rubin, C. Short, G. Booras, J. Davison, C. Ekstrom, M. Matuszewski and S. McCoy. A proposed methodology for CO_2 capture and storage cost estimates. *International Journal of Greenhouse Gas Control*, 17, 488-503, 2013 doi: 10.1016/j.ijggc.2013.06.004

SSB (Norwegian Statistisk sentralbyrå) webpage. Industrial cost index. Available on: https://www.ssb.no/en Accessed 25.01.2020.

M. van der Spek, S. Roussanaly and E. S: Rubin. Best practices and recent advances in CCS cost engineering and economic analysis. *International Journal of Greenhouse Gas Control*, 83:91-104, 2019. doi: 10.1016/j.ijggc.2019.02.006

F. Yu. *Process design for chemical engineers*. Ten Books, Inc., 2014.

L. E. Øi. Aspen HYSYS simulation of CO2 removal by amine absorption in a gas based power plant. The 48th Scandinavian Conference on Simulation and Modelling (SIMS 2007), Göteborg, Sweden, *Linköping University Electronic Press*, 27:73-81, 2007.

L. E. Øi, T. Bråthen, C. Berg, S. K. Brekne, M. Flatin, R. Johnsen, I. G. Moen, and E. Thomassen. Optimization of configurations for amine based CO_2 absorption using Aspen HYSYS. *Energy Procedia*, 51:224-233, 2014. doi: 10.1016/j.egypro.2014.07.026