Techno-Economic Analysis of CO_2 Capture from Pulp and Paper Mill Limekiln

Amod Parkhi*,, Selen Cremaschi*, Zhihua Jiang*,****

**Department of Chemical Engineering, Auburn University, 212 Ross Hall, Auburn, AL, 36849 USA (e-mail: azp0105@auburn.edu;szc0113@auburn.edu)*

***Alabama Center for Paper and Bioresource Engineering, Auburn University, 356 Ross Hall, Auburn, Al, 36849 USA (e-mail: zzj0012@auburn.edu)*

Abstract: This paper presents a techno-economic analysis of an absorption-based CO_2 capture process using Monoethanolamine (MEA) as the solvent for pulp and paper mill limekiln. The flue gas specifications were obtained from published limekiln data of a theoretical pulp and paper mill. The process was simulated in Aspen Plus and linked to CAPCOST using a python script for the cost calculations. The CO_2 capture cost estimates were compared to the only CO_2 capture costs data available in the literature for limekiln flue gas. Comparing the cost breakdown between the published data and this study, the capital cost difference was found to be highest for the stripper and the compression and dehydration sections. Further, the capture cost sensitivity analysis, evaluating the impacts of key parameters, including flue gas CO_2 mol%, MEA, electricity, and steam, showed that the capture costs varied from \$70 to \$82 per tonne of CO_2 captured.

Keywords: Carbon dioxide capture, Aspen Plus, Techno-economic analysis, Pulp & paper mill, Limekiln

1. INTRODUCTION

Of all the greenhouse gases (GHGs), carbon dioxide (CO_2) is the primary GHG emitted through human activities. In the U.S., industrial processes accounted for 16% of the CO_2 emissions, and the pulp and paper industry accounted for 1.2% (30 MMT/y) of these industrial emissions in 2019 (EPA, 2020). The primary sources of CO_2 emissions in pulp and paper manufacturing are from the recovery boiler, bark boiler, and limekiln (Onarheim *et al.*, 2017b). The economic feasibility of CO_2 capture from the pulp and paper sector has not been studied extensively as the CO_2 emissions are primarily of biomass origin. Because all the existing regulations and policy instruments consider only fossil-fuel CO_2, there are limited incentives for CO_2 capture (Jönsson, Kjärstad and Odenberger, 2014). While the MonoethanolAmine (MEA) solvent-based absorption desorption process for CO_2 capture has been widely studied for oil, gas, iron and steel, cement, and chemicals production (Leeson *et al.*, 2017), it is neither fully explored nor a mature technology for the pulp and paper industry. Also, there is a wide range of capture costs in the literature for this process (e.g., (Hektor and Berntsson, 2007), (Onarheim *et al.*, 2017a), (Garðarsdóttir *et al.*, 2018), (Nwaoha and Tontiwachwuthikul, 2019), (Yang, Meerman and Faaij, 2021)) making it difficult to assess its potential for the pulp and paper industry.

To the best of our knowledge, four studies provide estimates for CO_2 capture costs from pulp and paper industry emissions. Onarheim et al. (2017b) studied CO_2 capture and storage from various sources of a hypothetical Kraft pulp mill and a pulp and board mill. The capture process was MEA-based absorption desorption. The result showed that it is technically feasible to retrofit post-combustion CO_2 capture to an existing pulp mill or pulp and board mill. Another study (Onarheim *et al.*, 2017a) performed the techno-economic analysis of retrofitting post-combustion amine scrubbing to a pulp mill and an integrated pulp mill. The economic evaluations included estimating capital expences, total installed cost, total plant cost, and the discounted cash flow calculations. They utilized in-house cost databases of ÅF- consult Oy and Techincal Center of Finland (VTT) and quotes from vendors if required for the CO_2 capture plant. Also, a higher exponent value (0.7) for the "rule of six-tenths" was selected when estimating equipment costs. For 90% CO_2 capture from the limekiln flue gas, the capture cost was calculated as \$91 per tonne CO_2, which is the only cost estimate for CO_2 capture from pulp and paper mill limekiln available in the literature. We will refer to this study (Onarheim *et al.*, 2017a) as the reference study in this paper.

Gardarsdottir et al. (2014) evaluated the technical performance of post-combustion CO_2 capture integrated with three different industries, including a kraft pulp mill recovery boiler. The study concludes that a pulp mill can become a negative net contributor to global CO_2 emissions. However, the study didn't provide any techno-economic evaluations for the CO_2 capture process. Recently, Gardarsdottir et al. (2018) carried out an economic evaluation of 90% CO_2 capture from the recovery boiler flue gas. The study calculated the equipment costs using Aspen In-Plant Cost Estimator and assumed a generic cost level for all the sources. The estimated cost was \$61.8 per tonne CO_2. Nwaoha et al. (2019) also evaluated the cost of CO_2 capture from the recovery boiler flue gas. The paper considered two different solvents, MEA and 2-amino-2-methly-1-propanol and MEA blend, and three process configurations for each solvent. The equipment costs were calculated using reference guesstimate/ballpark estimates from the literature and scaling up to the equipment sizes suggested by the process simulation. The CO_2 capture costs varied from $129.05 to $147.23 per tonne of CO_2.

In this study, we perform a techno-economic analysis for CO_2 capture using the MEA-based absorption desorption process for pulp and paper mill limekiln and compare the capture cost to the only available CO_2 capture cost estimate in literature (Onarheim *et al.*, 2017a). The CO_2 capture from the limekiln section of a pulp and paper mill considers the system in addition to the existing pulp and paper mill. For the comparison, the same limekiln flue gas specifications (Onarheim *et al.*, 2017b) are utilized in the analysis. Detailed process and techno-economical analysis for these two studies are provided in the report sponsored by the International Energy Agency Greenhouse gas R&D Programme (IEAGHG) and prepared by ÅF- consult Oy and Techincal Center of Finland (VTT) (IEAGHG, 2016). A variant of split-flow configuration was modeled using Aspen Rate-Based Distillation to reduce the reboiler heat duty. Further, the CO_2 capture plant is equipped with an amine reclaimer recovering the solvent lost to heat stable salts. Unlike the report, a conventional 30 Wt.% MEA CO_2 absorption desorption process is simulated in our study using Rate-Based Distillation in Aspen Plus to reduce convergence issues in Aspen Plus. The equipment costs are calculated using CAPCOST (Turton *et al.*, 2018). The present study applies, for the first time, the CAPCOST factor-based modular program for techno-economic analysis of a solvent-based CO_2 capture process using MEA as the solvent for pulp and paper mill limekiln. CAPCOST uses custom equations to perform the equipment cost calculations which provides transparency and editability while dealing with various equipment, different process configurations, and flowsheet optimization. Further, a sensitivity analysis was carried out to understand the impact on capture cost of varying flue gas CO_2 compositions and the changes in the MEA solvent, electricity, and steam costs.

2. PROCESS SIMULATION AND ECONOMIC ANALYSIS

2.1 Kraft pulp and paper mill

The process by which the fibrous raw material is reduced to a fibrous mass is called pulping. The Kraft pulping process involves cooking the wood chips in 'white liquor,' a sodium hydroxide and sodium sulfide solution. As in Fig. 1, wood chips from the wood preparation section are cooked in 'white liquor' at around 170°C for up to 2 hours. After cooking, the washed pulp is screened, bleached if needed, and formed into paper in the paper machine section. The weak black liquor, after washing, is concentrated and burned in the recovery boiler for chemical and energy recovery. The recovery boiler smelt is dissolved in water to form green liquor and causticized with the reburned lime (CaO) to form white liquor used in the cooking process, completing the cycle (Gary A.Smook, 2015).

CO_2 emissions from the bark boiler and recovery boiler are considered biogenic as the CO_2 is released from burning wood-derived fuels. In the limekiln (Fig. 1), CO_2 is released during the calcination of limestone (Onarheim *et al.*, 2017b).

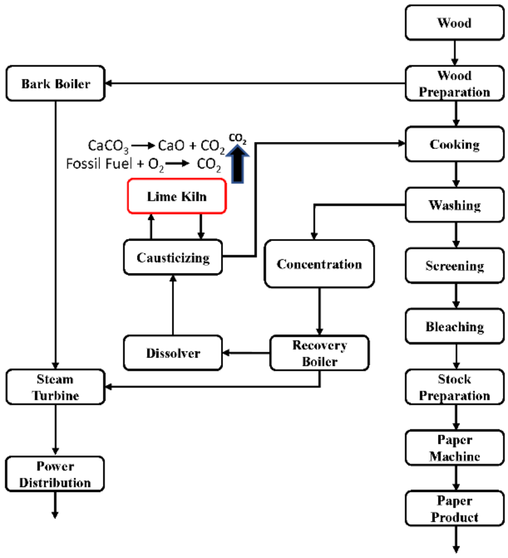

Fig. 1. Overview of a Kraft pulp and paper mill

2.2 Aspen Plus simulation of the MEA-based absorption process for CO_2 capture from limekiln

Aspen Plus simulation model was constructed utilizing the main design parameters listed in the reference study to simulate the CO_2 capture process from the same limekiln flue gas. The economic analysis was carried out using the equations and data in CAPCOST. Table 1 summarizes the assumptions for calculating the capture costs and the utility and raw material costs used in the calculations. The MEA cost is taken from Kohl and Nielsen (1997) and inflated using the producer price index for the year 2016.

Table 1 Economic analysis assumptions, utility costs, and raw material prices (Turton et al., 2018)

Parameter	Value	units
Operation hours	8400	Hours (h)
Plant operation	20	Years (m)
Interest rate	20	% (i)
Maintenance & repair costs	6	% of $C_{Capital}$
Insurance & taxes	2	% of $C_{Capital}$
Makeup-water	0.177	$/tonne
Operators	13	-
Supervisory level costs	18	% of C_{Labor}*
Start-up & MEA costs	10	% of $C_{Capital}$
General & Administrative costs	18 % C_{Labor} + 0.9 % $C_{Capital}$	
Plant overhead	70.8 % C_{Labor} + 3.6 % $C_{Capital}$	
Contingency	15	% C_{BM}**
Contractor Fee	3	% C_{BM}
Auxiliary facility cost	50	% C_{BM}
Steam (5 barg)	9.45	$/tonne
Cooling water	0.0157	$/tonne
Electricity	0.0674	$/kWh
Make-up MEA	2,100	$/tonne

* Labor cost
** Bare module costs

Fig. 2 shows the Aspen Plus flowsheet developed for the solvent-based CO_2 capture process, excluding the CO_2 compression train (Fig. 3) in the reference study. The absorption cycle is a temperature-dependent acid-base reaction where the flue gas CO_2 (weak acid) reacts with a solvent (a weak base) (1). The "CO_2 loaded" solution (rich MEA) is stripped off of its CO_2 by reverse reaction (Eq. 1), regenerating the lean solvent and a gaseous CO_2 product (Bhown and Freeman, 2011). The absorption-desorption reactions for primary amines like MEA can be represented in (1) (Kohl and Nielsen, 1997).

$$RNH_2 + CO_2 \leftrightarrow RNHCOO^- + H^+ \quad (1)$$

The flue gas from the limekiln stack is cooled and quenched in the direct contact cooler (DCC). The cooled flue-gas enters the absorber column from the bottom, and the lean MEA regenerated from the stripper column enters from the top. Solvent MEA, after absorbing CO_2 from the flue gas, is pumped and heated in a rich/lean heat exchanger with the stripper column bottoms (lean MEA) before being sent to the stripper column. The CO_2 and some water vapor are recovered as the top product of the stripper column, the CO2OUT stream in Fig. 2. The lean MEA, recovered as the stripper bottoms, passes through the rich/lean heat exchanger and is mixed with makeup water and amine. The lean MEA is further cooled with cooling water before being sent back to the absorber.

The flue gas specifications are given in Table 2. We used the specifications given in the reference study for comparison.

Table 2. Limekiln flue gas data (Onarheim et al., 2017b)

Parameter	Units	Value
Temperature	°C	250.0
Mass flow	MTPY	684 000
CO_2	mol %	20.4
N_2	mol %	47.4
O_2	mol %	1.2
H_2O	mol%	30.9
SO_x	ppm	50.0
NO_x	ppm	175.0
TRS	ppm	15.0
Particulates	ppm	30.0

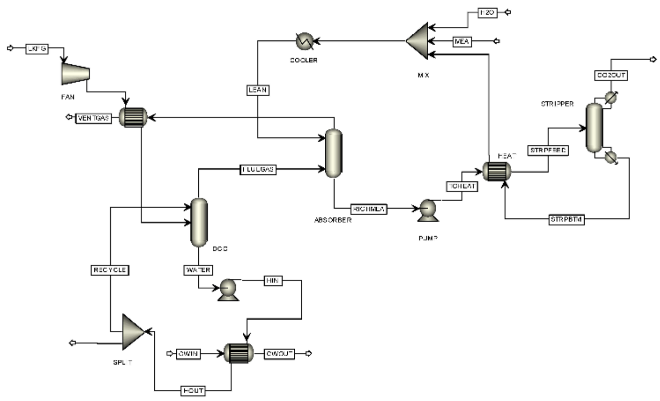

Fig. 2. Aspen Plus model of CO_2 capture

Aspen Plus version, the property package, the model, and the built-in units used for the columns and other equipment are summarized in Table 3. The KEMEA package contains kinetics and rate constants, which allows modeling the MEA system more accurately with an electrolyte-NRTL model. RadFrac unit operation is used to model the absorber and stripper columns. The absorber does not have a condenser or a reboiler. The stripper has a condenser at the top and a reboiler at the bottom. Both absorber and stripper columns employ the rate-based model. Pressure drops in the piping and equipment are neglected. The stripper column operates at 1.8 bar(a) to prevent potential solvent degradation due to high pressure and higher stripper bottom temperatures (Abu-Zahra *et al.*, 2007).

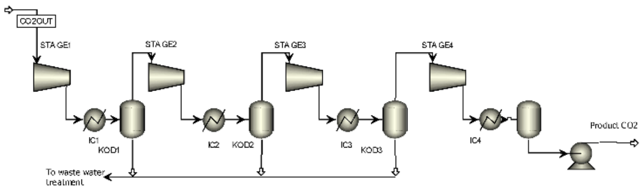

Fig. 3. CO_2 compression train

To replicate the process conditions for the temperature and pressure from the reference work, a four-stage compressor train is used to compress the product CO_2 to 110 bar at 33°C. The Aspen Plus flowsheet for the compressor train is presented in Fig. 3. The property packages and the unit operations used in Aspen Plus simulation are given in Table 3.

Table 3. Property package and the equipment model used

Parameter	Package/units
Aspen Plus	V10.0
Property Package	KEMEA (Kent-Eisenberg)
Absorber Column	RadFrac (Rate-based model)
Stripper Column	RadFrac (Rate-based model)
Lean/rich heat exchanger	HeatX
MEA cooler	HeatX
Compressor	Compr
Intercoolers	Heater

2.3 Economic analysis of the MEA-based absorption process

The CO_2 capture costs include capital and operating costs. CAPCOST is used to evaluate the capital costs for the columns (DCC, absorber, and stripper), heat exchangers (condenser, reboiler, rich/lean heat exchanger, intercoolers, and the solvent cooler), compressors/fan, and drivers for compressors and pumps. The costs are adjusted for inflation using the 2016 Chemical Engineering Plant Cost Index (CEPCI) value of 542 (Turton *et al.*, 2018). The total annualized cost (TAC) of the process, including the compression train, is shown in (2),

$$TAC\left(\frac{\$}{y}\right) = AF \times C_{Capital} + C_{Operating} \quad (2)$$

where, $C_{Capital}$ and $C_{Operating}$ are capital and operating costs, and AF is the annualization factor.

The equipment installation costs for estimating the capital cost are calculated using the equipment bare module costs from CAPCOST. The absorber and stripper columns are modeled as packed columns. The column diameters are taken from the converged Aspen plus file and the packed height per stage (HETP) in meters is calculated using (3) (Wankat, 2004):

$$HETP = \frac{100}{a_p} + 0.1 \quad (3)$$

where, a_p is the surface area per volume of the packing. The capital cost includes the bare module costs, contingency, contractor fee, plant start-up costs, and auxiliary facilities costs. The operating costs comprise the labor costs, the process utility, and the raw materials costs.

The annualization factor, AF, is calculated in (4),

$$AF = \frac{i(1+i)^m}{(1+i)^m - 1} \quad (4)$$

where, i is the interest rate, and m is the plant operation year.

The total CO_2 capture cost is calculated by using (5):

$$CO_2\ capture\ costs\ \left(\frac{\$/yr}{t\text{-}CO_2/yr}\right) = \frac{TAC}{Captured\ CO_2} \quad (5)$$

The degradation of amine by the presence of SO_x in the flue gas is accounted for by considering MEA degradation losses due to SO_x in the economic evaluation. Separate treatment for the NO_x removal was not considered as NO_x present in the flue gas is regarded as inert. The amount of NO_2 is below the recommended limit, and a separate NO_2 removal setup is not required (IEAGHG, 2016).

3. OVERVIEW OF THE REFERENCE STUDY ECONOMIC ANALYSIS

3.1 CO_2 capture setup of the reference study

A 30% MEA based CO_2 capture process was simulated in Aspen Plus using the Aspen Rate-Based Distillation model for processing the flue gas given in Table 2. A solvent split-flow configuration was used to reduce the stripper column reboiler duty. The overall CO_2 capture rate was set at 90%.

3.2 Economic analysis of the reference study

Economic analysis results were reported in terms of Earning Before Interest, Taxes, Depreciation, and Amortization (EBITDA). For the mill and the CO_2 capture plant, the cost calculations included the capital investments (CAPEX) and the operating costs (OPEX).

Equipment costs were estimated using in-house cost databases of ÅF consult Oy and VTT, and quotes from vendors when necessary. When the capacity of the equipment was different from the quoted capacity, a scaling factor of 0.6 was applied to estimate the equipment cost using (6). In (6), the exponent n was assumed to be 0.7 for the CO_2 capture plant.

$$\frac{C}{C_0} = \left(\frac{S}{S_0}\right)^n \quad (6)$$

In (6), C is the scaled capital cost, C_0 is the actual purchased cost, S is the target capacity, and S_0 is the design capacity. IEAGHG economic assessment model developed in-house was applied to calculate the CO_2 avoided costs using a discounted cash flow (DCF) analysis. For CO_2 capture from the limekiln section, the amount of CO_2 avoided is equal to the amount of CO_2 captured. Based on the DCF calculations, the cost of CO_2 capture from the limekiln section was \$91 per tonne CO_2, which, at a CO_2 capture rate of 0.24 ton CO_2 per air dried tonne pulp (adt_{pulp}) produced translates to \$22 per adt_{pulp}. The cost estimates were developed in EUR (Quarter 2 of 2015). An exchange rate of 1.0 EUR = 1.1 USD was used for currency conversion.

4. RESULTS AND DISCUSSIONS

4.1 CO_2 capture costs evaluation and comparison with the reference study

The base case CO_2 capture cost for the process simulated using design and operation values from the reference study, the theoretical mill limekiln, are given in Table 4. The costs are rounded off to the nearest 1000. For the operating capacity of the mill and the CO_2 capture rate of 23.3 t/h, the amount of CO_2 capture is 0.24 ton CO_2 per adt_{pulp}, translating the capture cost of \$76 per tonne CO_2 to \$18 per adt_{pulp}.

Table 4. Summary of CO_2 capture costs for the base case

Parameter	Value
Total capital cost [\$]	19,606,000
Total annualized capital cost [\$/y]	4,019,000
Total operating cost [\$/y]	6,105,000
Total raw material cost [\$/y]	461,000
Total utilities cost [\$/y]	5,380,000
CO_2 capture [t/y]	196,000
CO_2 capture cost [\$ per tonne CO_2]	76
CO_2 capture cost [\$ per tonne pulp produced]	18

For the economic analysis of this study, the equipment costs given in Table 5 are calculated using the equations given in CAPCOST. These equations require the equipment sizes, which are directly obtained from the Aspen Plus simulation. Various factors contributing to the estimation of equipment grassroot costs (C_{GR}) are given in Fig. 4. Fig. 4 also includes the calculations for estimating the bare module cost (C_{BM}), which is defined as the direct and indirect expenses incurred for equipment purchase and installation. Adding contingency and contractor fees to C_{BM} gives the total module cost (C_{TM}). Finally, including the auxiliary facilities costs in the C_{TM} provides the C_{GR} for each piece of equipment.

Table 5. Equipment-wise breakdown of the capital costs

CO_2 capture plant section	Reference study (\$ million)	This study (\$ million)
Direct Contact Cooler	0.7	2.4
Amine absorber section	5.3	3.1
Amine circulation system	0.5	0.9
Stripper section	11.6	2.3
Compression & dehydration	1.4	7.1
Auxiliary facility costs	0.2	2.5
Start-up costs	3.8	1.3
CAPEX	23.5	19.6

The total capital cost is annualized for the plant operation time and includes the columns, exchangers, pumps, dehydration and compression, auxiliaries, and the start-up costs. The contributions of individual equipment to the total capital cost are shown in Table 5, which also lists the values from the reference study (IEAGHG, 2016) for comparison.

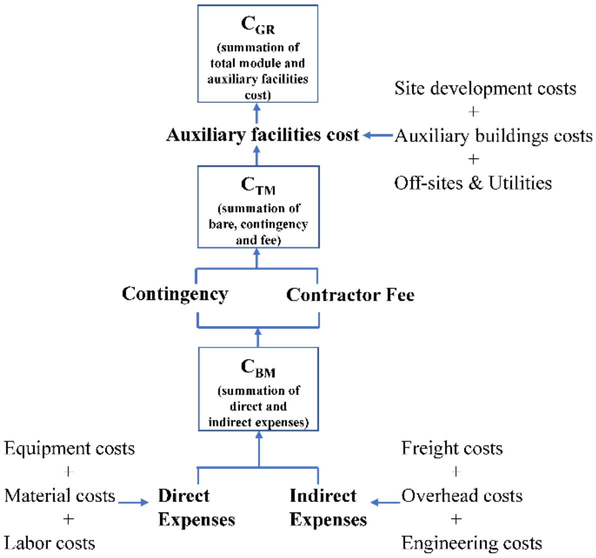

Fig. 4. Factors considered in CAPCOST for estimating equipment costs

Table 5 reveals the differences in equipment costs. For the reference study, the total installed cost, including the project contingency, forms the total plant cost of the CO_2 capture plant. The total installation cost of the CO_2 capture plant includes the installation costs for each piece of equipment individually. The equipment costs for the reference study were adjusted using the 2015 CEPCI value, which is ca. 3% higher than the CEPCI value of 542 considered in this techno-economic analysis. The reference study separately calculated the costs incurred in Engineering, Procurement, and Construction (EPC) and construction for the entire CO_2 capture plant. Further, in the reference study, a project contingency of 10% was added to the total plant installation costs. The estimated value of total installation costs (TIC) with the required mill modification, and other CAPEX of spare parts, start-up costs, owner's cost interest, and working capital forms the total capital requirement (TCR). The CO_2 capture costs were estimated within a ± 50% accuracy, assuming that the implementation of CO_2 capture technologies in the pulp and paper industry is not a mature technology. A 30% increase in the capital costs for the columns and the solvent circulation was applied to account for the solvent split-flow configuration in the reference study (Lars Erik, 2012).

As the equipment costs calculated using CAPCOST rely on the sizes estimated by Aspen Plus, the contribution of equipment costs to capital costs for different sections are different from those reported in the reference study (Table 5). The EPC and constructions costs have been included in the factors considered while calculating the C_{GR}. On the contrary, in the reference study, the construction and the EPC costs have been calculated separately for the entire facility, and the total plant installation cost was the summation of the equipment, the construction, and the EPC costs. Further, the setup used for the CO_2 capture (a split-flow configuration) in the reference study (Onarheim *et al.*, 2017b) is different from that used in this study. A reclaimer for the degraded MEA is used in stripper column setup; however, a conventional absorption setup is used to simulate the process in this study. A disadvantage of the split flow configuration is the requirement of a higher solvent circulation flowrate to achieve a similar CO_2 recovery compared to the conventional configuration. Corrosion is another issue related to the use of split-flow configuration due to the higher percentage of the MEA components in the solvent. Also, the capital investment for the split flow configuration is higher due to the requirement of a larger absorber column, additional heat exchangers, pumps, and the associated pipings (IEAGHG, 2016). The cost of the amine absorber section in the reference study is around 70% higher than the costs of the absorber section in this study (Table 5).

The second term in (2) is the operating costs, including fixed and variable costs. These costs for the reference study and the present work are provided in Table 6. The total operating costs are similar (Table 6). However, unlike the reference study (Onarheim *et al.*, 2017a), we did not consider the costs incurred for CO_2 transportation and storage in operating costs. The reference study (Onarheim *et al.*, 2017a) considered the fixed costs, variable costs, income from electricity sold to the grid, and the CO_2 storage and transportation costs for calculating operating costs. The fixed costs consisted of direct-indirect labor, insurance, local taxes, and maintenance. Chemical and utility costs, and waste processing and disposal charges were included in the variable costs. Also, for the base case and the case with limekiln CO_2 capture in the reference study, the biomass feedstock costs are the same nullifying the impact of changes in the input biomass costs while calculating the levelized cost of pulp.

Table 6. Operating cost details ($ million/y)

Operating costs	Subsection	Reference study ($ million/y)	This study ($ million/y)
Fixed costs	Operating labor Direct supervisory Plant overhead Insurances and taxes Maintenance	3.9	2.3
Variable costs	Utilities Chemicals	3.6	5.2

The breakdown of the capture costs for the reference study and this work is shown in Fig. 5 in terms of million $/y. A cost difference is observed for the CAPEX and the fixed cost portion of the OPEX. The direct labor cost in the reference study is 1.3% lower than the direct labor cost used in this study. However, in the reference study, the indirect labor costs which include the costs of administration and the general overhead are 40% of the direct labor costs, and for this study, the contribution of the direct labor costs in the administration and the general overhead costs is at 18%. The higher percentage contribution of the labor costs leads to a higher contribution of the indirect labor cost for the fixed costs in the reference study. For the reference study, the loss of electricity exported due to steam use from within the mill for the stripper column reboiler is considered a loss in revenue. The OPEX is similar for both studies; as a result, the difference in the capture costs is attributed to the difference in the total CAPEX, which is higher for the reference study (IEAGHG, 2016).

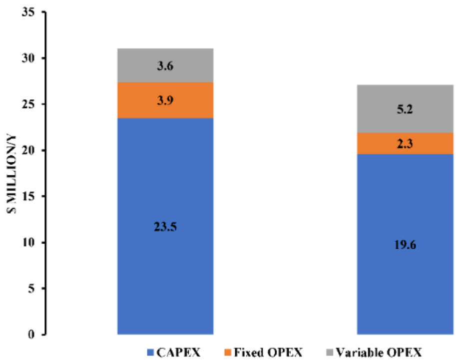

Fig. 5. CAPEX and OPEX comparison

Table 7 provides the energy consumption of the main equipment of the present study. Stripper column reboiler steam alone contributes to around 52% and the electricity costs contribute to 23% of the total operating costs.

Table 7. Energy consumption of the main equipment

Equipment	Energy consumption
Stripper Reboiler	$3.3\ \frac{GJ}{t-CO_2}$
Stage 1 compressor	629 kW
Stage 2 compressor	724 kW
Stage 3 compressor	479 kW
Stage 4 compressor	338 kW

The cost of the stripper section is higher for the reference study; however, the compressor costing is higher for the present study. The stripper section cost in the reference study includes the costs of the stripper column, water wash column, condenser and reflux drums, reboilers, and the associated pumps. The reference study did not include the cost breakdown for the individual equipment and the sizing for the condenser, the reboiler, and the reclaimer for the stripper section, making a one-to-one comparison impossible. Per CAPCOST estimates, the reboiler, condenser, and pump contribute more than 50% of the total stripper section costs. Table 8 summarizes the stripper section equipment and costs of the individual components for this study.

Table 8. Stripper section costing: CAPCOST

Stripper section	Costs ($ million)
Stripper column (including packings & internals)	1.1
Condenser	0.4
Reboiler & pump	0.8

When the stripper section costs are compared to the absorber section for the reference study, the stripper section cost is more than two times the cost of the absorber section. The absorber column has a larger diameter, 3.9 m, and is taller, 25 m, than the stripper column, which has a diameter of 2.9 m and a height of 20 m in the reference study (IEAGHG, 2016). Considering only the size difference of the absorber and the stripper columns, the pricing of the stripper column should be lower. Table 9 provides the equipment sizing for this study. Also, both absorber and stripper columns have the same packing material, Sulzer mellapack 250Y. However, the stripper column section has a condenser, a reboiler, and an MEA reclaimer, making the stripper column section as a whole expensive as compared to the absorber section.

Table 9. Specification for main equipment in this study

Equipment	Sizing
Direct contact cooler	Ø2.7 m × H10.1 m
Absorber	Ø3.9 m × H25.3 m
Stripper	Ø2.9 m × H19.8 m
Rich/lean heat exchanger	1621 m^2
Reboiler	700 m^2
Condenser	177 m^2
Solvent cooler	574 m^2

A significant difference is observed in the compressor and dehydration section cost between the reference study and the present work. A cost breakdown of the compressor and dehydration section was also not given in the reference study. A breakdown of the individual component costs in each section of the CO_2 capture plant and the equipment sizes would have enabled a more direct comparison.

The CO_2 capture process flow diagram can be divided into three main sections to compare the section-wise contribution to equipment costs. Table 10 lists the three main sections, the major equipment in each section, and the percentage section-wise cost distribution. Table 10 reveals that the major contributor to the reference study's capital costs is the CO_2 capture section, followed by the compressor and dehydration, and then the pretreatment sections. We guestimate that 1) the use of an MEA solvent split configuration for reducing reboiler heat duty, 2) the higher cost of the stripper column and its auxiliary units of the reboiler, condenser, and the MEA solvent reclaimer, and 3) the use of a higher exponent value for estimating equipment costs in (4) by the reference study led to a higher contribution of the CO_2 capture section. However, the capital cost estimated by CAPCOST suggests that the compressor unit is the major contributor to the total capital cost, followed by the CO_2 capture section and then the pretreatment. The pretreatment unit's contribution to the entire capital cost is the least for both approaches. If we did not assume n^{th} plant for our economic analysis (similar to assuming a higher exponent in (6) for the equipment cost calculations) and include a 30% increment in the column costs, the CAPEX for this study would be comparable to the reference study. Considering a 30% increment in the column costs, the CAPEX difference between the reference and this economic analysis reduces from 16.6% to 11.2%.

Table 10. CO_2 capture process section-wise cost breakdown

Section	Major equipment	Reference study (%)	This study (%)
Pretreatment	Flue gas blower Vent gas heater Direct Contact Cooler (column & packing) Pump & cooler	3.5	15.3
CO_2 capture	Absorber section (column & packing) Amine circulation system (pumps & heat exchangers) Stripper section (column & packing) Condenser & reboiler	89.3	39.8
Compression & dehydration	Compressor package Heat exchangers Dehydration unit	7.2	44.9

4.2 Sensitivity of the capture cost for the base case of this study

The variable cost in OPEX includes the utilities, mostly steam and electricity, and chemicals, the makeup MEA. Steam is consumed in the stripper column reboiler, electricity for the compressor unit, and MEA solvent for CO_2 absorption. The capture cost sensitivity is evaluated with varying MEA solvent, electricity, and steam costs. Further, we analyzed the cost sensitivity to changes in flue gas inlet CO_2 mol%.

Fig. 6 (blue) gives the CO_2 capture costs variation with changing flue gas CO_2 composition (mol%) from 18 mol% to 22 mol %. As the CO_2 mol% increases, an increase is observed in the equipment sizes; however, the capture costs decrease because of the relatively higher change in the amount of CO_2 captured. These observations align with previously reported conclusions (Nuchitprasittichai and Cremaschi, 2011).

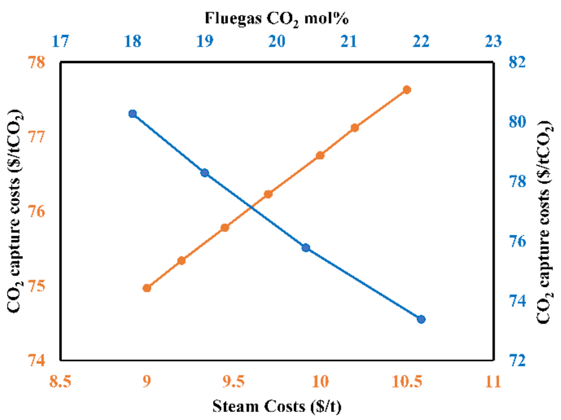

Fig. 6. Capture cost sensitivity with the flue gas CO_2 mol% and steam costs

For this study, the medium pressure steam in the stripper column reboiler contributes to ca. 65% of the total utility costs and 22.5% of the total CO_2 capture costs. The sensitivity of the CO_2 capture costs to the steam cost changes, from \$9 to \$10.5 per tonne of medium pressure steam, is evaluated and shown in Fig. 6 (orange). The steam cost changes significantly impact the OPEX due to its largest share in the utility costs. Hence, there is a linear correlation between the capture cost and the steam cost. For an increase of \$1 per tonne in steam costs, the CO_2 capture costs increase by 2.4% (Fig. 6).

The MEA solvent makeup is needed due to the MEA losses from the absorber and the stripper columns and also to account for the MEA degradation due to the presence of SO_x in the flue gas. The MEA makeup and degradation losses account for 5.5% of the total CO_2 capture costs. The sensitivity of the CO_2 capture costs is evaluated for the MEA cost values of \$1000 and \$5000 per tonne of MEA in Fig. 8 (blue). Almost a 5% increase is observed in the capture cost when the MEA prices increase from \$1000 to \$5000 per tonne.

Electricity is used in the compressor and dehydration section, contributing to almost 26% of the total utility costs. Electricity makes up 8.8% of the CO_2 capture costs. The capture cost as a function of the electricity price is illustrated in Fig. 8 (orange). The literature study (Onarheim *et al.*, 2017a) also observed an increase in the capture costs with increasing electricity costs.

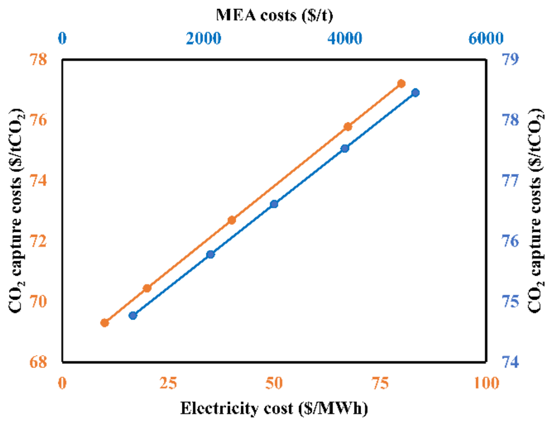

Fig. 7. Capture costs sensitivity with the MEA and electricity costs

5. CONCLUSIONS AND FUTURE DIRECTIONS

This techno-economic analysis evaluated the CO_2 capture costs from pulp and paper mill limekiln and compared the costs to published capture cost data (in the reference study). The capital investment was estimated using CAPCOST in contrast to in-house tools and vendor quotes used in the reference study. A significant difference is observed in the capital cost between the reported value and this study. The difference in the overall capture cost is mainly attributed to the costing equations and the methodology used in evaluating the base equipment, EPC, construction, and fixed operating costs. The literature study estimated the costs considering a high degree of uncertainty due to the lower maturity of the MEA based CO_2 capture process for use in the pulp and paper industry. In contrast, CAPCOST uses the sizing parameters estimated by the Aspen Plus simulation and calculates the equipment costs assuming n^{th} power law. Considering the decrease in the electricity sold to the grid, the reference study took steam integration into account, leading to a lower utility cost than observed in this study. However, OPEX for both approaches was found to be similar to each other.

We conducted a sensitivity analysis on the capture cost by varying the inlet flue gas composition (CO_2 mol%), and steam, electricity, and MEA prices. The results revealed that the capture costs vary from $70 to $82 per tonne of CO_2 captured.

Our study provides a basis for cost calculations, with details on TEA, and helps for future process simulation and optimization studies. The use of CAPCOST modular program provides transparency to replicate cost calculations and utilize this approach to evaluate the CO_2 capture costs in the pulp and paper industry.

Future work will focus on processing flue gas data from two different lime kiln sections, a real integrated paper mill, and a liner board mill, analyzing the capture costs for different input conditions and production rates. Optimization of the CO_2 capture process considering the equipment sizing and the operating conditions as decision variables for capture cost minimization will be performed. Further, steam integration from within the mill will be explored to reduce the total capture costs, and in-mill application of captured CO_2 with an existing federal tax credit for carbon capture and sequestration (Section 45Q - Internal revenue code) will be studied. We also plan to analyze the cost sensitivity with the economic parameters.

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