



Techno-Economic Assessment for CO₂ Capture From Air Using a Conventional Liquid-Based Absorption Process

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In this study, the process of carbon dioxide (CO₂) capture directly from ambient air in a conventional monoethanolamine (MEA) absorption process was simulated and optimized using a rate-based model in Aspen Plus. The process aimed to capture a specific amount (148.25 Nm³/h) of CO₂ from the air, which was determined by a potential application aiming to produce synthetic methane from the output of a 2.7 MW electrolyser (593 Nm³/h H₂). We investigated the technical performance of the process by conducting a sensitivity analysis around different parameters such as air humidity, capture rate defined as a ratio of moles of CO₂ captured during the process to the total mole of CO₂ in the feed stream, CO₂ loading of lean and rich absorption liquids and reboiler temperature, and evaluated the energy consumption and overall cost in this system. In order to meet the design requirement for standard packed columns, the rich absorption liquid was circulated to the top of the absorber. A capture rate of 50% was selected in this process as a baseline. At higher capture rates, the required energy per ton of captured CO₂ increases due to a higher steam stripping rate, required in the desorber, and at lower capture rates, the size of equipment, in particular, absorber and blowers increases due to the need for processing a significantly larger volume of air at the given CO₂ production volume. At the base case scenario, a reboiler duty of 10.7 GJ/tCO₂ and an electrical energy requirement of 1.4 MWh/tCO₂ were obtained. The absorber diameter and height obtained were 10.4 and 4.4 m, respectively. The desorber is found to be relatively small at 0.54 m in diameter and 3.0 m in height. A wash water section installed at top of the absorber decreased the MEA loss to 0.28 kg/ton CO₂. However, this increased capital cost by around 60% resulting in CO₂ capture costs of \$1,691 per ton CO₂ for the MEA base scenario. Based on the techno-economic analysis, assuming a non-volatile absorbent rather than MEA thereby avoiding a wash water section, and using an absorption column built from cheaper materials, the estimated cost per ton of CO₂ produced was reduced to \$676/tCO₂. The overall cost range was between \$273 and \$1,227 per ton of CO₂ depending on different economic parameters such as electricity (\$20–\$200/MWh) and heat price (\$2–\$20/GJ), plant life (15–25 years) and capital expenditure (±30%). In order to reduce the cost further, the use of innovative cheap gas-liquid contactors that operate at lower liquid to gas ratios is crucial.

Keywords: CO₂ capture, techno—economical assessment, chemical absorbents, amine, air capture

INTRODUCTION

Ongoing use of fossil fuel over the past one and half century led to an increase in the concentration of CO₂ in atmosphere from around 280 ppm to just above 400 ppm (Lindsey, 2019). Concerns about the effect of this increase on the global climate resulted in increasing efforts for the development of technologies that enable to remove CO₂ directly from atmosphere referred to as Negative Emissions Technologies (McGlashan et al., 2012; McLaren, 2012; Pritchard et al., 2015). Recent studies indicate that in order to prevent the increase of global temperature to about 2°C above the pre-industrial era by the end of this century, the large scale deployment of negative emission technologies is probably required (Gasser et al., 2015).

Among all proposed negative emission technologies (McLaren, 2012; Gasser et al., 2015), the direct capture of CO₂ from air (DAC) by physical or chemical processes attracted more attention due to the advantages that this method offers. For instance, DAC technology requires a relatively small area of land and can be located close to the storage/utilization sites or even can be deployed in remote areas where the land is unusable, or on the roofs of buildings in populated cities. It can also provide a larger removal capacity compared to other methods of CO₂ removal from atmosphere (Bacocchi et al., 2006). Moreover, DAC was claimed to provide a means for a permanent decrease of CO₂ concentration in atmosphere (in fact it can capture 100% of CO₂ emission to the atmosphere), for capturing dispersed fugitive emissions, and for direct use in different industries such as beverage, greenhouse, and synthetic fuels production industries (Lackner et al., 1999; Keith, 2009; Lackner, 2009; Krekel et al., 2018).

Several technologies are proposed and investigated in the literature for the direct capture of CO₂ from ambient air. Carbon Engineering, Climeworks and Global Thermostat are the three major companies that developed technologies for the large scale capturing of CO₂ from air. Carbon Engineering uses a potassium hydroxide solution to capture CO₂ and the energy intensive calcination process to regenerate the solvent. On the other hand, the other two companies, Climeworks and Global Thermostat use solid sorbents to capture CO₂. They use high temperature and low pressure to regenerate the sorbent. A summary of these studies is shown in **Table 1**. Based on these studies, DAC is currently considered as an energy-intensive and costly approach for removing CO₂ from the atmosphere. However, researchers believe that the high cost and energy consumption in such systems are mostly due to the immaturity of the technology, and hence DAC may be eventually considered as a viable option for removing CO₂ from the atmosphere (Field and Mach, 2017; Senftle and Carter, 2017). Further research is under way on different aspects of the process such as the chemistry and stability of liquid and solid sorbents, their effective contact with gas and process modifications in order to reduce the energy requirements and cost of such systems.

The use of aqueous alkaline sorbents such as NaOH, Ca(OH)₂, and KOH, based on the concepts of the conventional MEA absorption process, for the DAC application, was claimed to provide a simpler contact between air and sorbents. It was also

anticipated that the system can operate continuously with a very long contactor's lifetime as the absorbing liquid in such systems is not subject to degradation reactions. The high cost for the regeneration of the aqueous solutions and water loss are the main disadvantages of such liquid-based absorption systems (Lackner, 2009; Wang et al., 2011; Keith et al., 2018).

While testing CO₂ capture systems at large scale is expensive, process simulation programs such as Aspen Plus have been widely used to evaluate the process configurations and identify the optimum operating conditions. There are a number of studies that used Aspen Plus to model CO₂ capture from flue gases in MEA-based absorption processes (Desideri and Paolucci, 1999; Freguia and Rochelle, 2003; Alie et al., 2005; Svendsen and Hoff, 2005; Øi, 2007). However to the best of our knowledge, no such simulation study has been conducted for CO₂ capture from air. Such a study is believed to be important as it provides a baseline for the technical and economic performance of DAC technologies that use chemical absorption processes.

In this work, using a rate-based model in Aspen Plus, we conducted a comprehensive analysis on the performance of a conventional MEA-based absorption process for capturing CO₂ from air, with particular focus on reducing the thermal and electrical energy consumption and, ultimately, the overall cost. The process performance is evaluated against the various parameters such as air humidity, the CO₂ loadings of lean and rich solutions, capture rate and reboiler temperature. As a result, a benchmark condition for DAC technologies that use a chemical absorption process is determined and a parametric techno-economic assessment is conducted for this baseline case study. Further analysis of the study results has resulted in the identification of areas for efficiency improvement and cost reduction.

METHODOLOGY

Process Description

An MEA-based air capture process was designed to capture CO₂ from air. **Figure 1** shows the representation of the air capture process using a standard stripping process (a) and a cold rich-split process (b).

A conventional packed column was utilized for CO₂ absorption using a 30 wt% MEA solution, which is typically used for CO₂ capture from flue gas. It is important to note that the CO₂ concentration in air is ~300 times lower than that in the flue gas from coal-fired power station. Hence the absorption liquid flow could be much lower than that in a flue gas capture system. However, this would give rise to an extremely low liquid to gas ratio (L/G) that could not be achieved in a traditional packed column. In terms of operability, the process design used a liquid recirculation process in which most of the rich CO₂ absorption liquid exiting absorber is pumped to the top of absorber in order to achieve a standard L/G of about 2.5 in the absorber, at a 70% flooding ratio. A small portion of the rich absorption liquid is sent to the desorber for CO₂ removal. The regenerated lean absorption liquid from the desorber was cooled and mixed with the absorption liquid exiting the absorber, and fed back to the

TABLE 1 | Major technologies on direct air capture.

Investigator	Capturing method	Sorbent	Regeneration	Application	Maximum capacity (ton/year)	References
Carbon Engineering Company	Absorption	KOH solution	Calcination process (High temperature)	Synthetic fuels	1,000,000****	Keith et al., 2018
Climeworks Company*	Adsorption	amine based porous sorbents	High temperature	Greenhouse, Synthetic fuels	900	Climeworks press release, 2017; Marshall, 2017
Global Thermostat Company	Adsorption	amine based monoliths	High temperature and low pressure (vacuum)	Beverage and food	4,000****	Li et al., 2016; Chichilniski, 2018
Arizona State University and Columbia University***	Adsorption	Anionic exchange resin	High moisture	–	Not declared	Lackner, 2009; Wang et al., 2011
Others**: VTT, the Oak Ridge Laboratory, The US Naval research lab, X and PARC, European Space Agency, infinitree	Adsorption/ Absorption	Amine-functionized polymer/Aqueous amino acid solutions/ Ion exchange	High temperature/low pressure/precipitation/	Fuels/longer space mission/Greenhouse	<1	Sandalow et al., 2018

*First company that commercially captured CO₂ from air, producing around 900 tCO₂/year for a greenhouse.

**Produced CO₂ at very small scale.

***Produced CO₂ at low concentration (5%).

****They claimed this as their predicted production capacity.

top of absorber for continuous CO₂ absorption. The desorption process can be modified using a cold rich-split configuration, i.e., sending a small portion of cold rich absorption liquid to the top of desorber, as shown in **Figure 1B**, to reduce the reboiler duty.

Process Simulation

A rigorous, rate-based MEA model developed in Aspen Plus was used to simulate the air capture process. The MEA model has been well-validated in previous work at CSIRO (Li et al., 2016; Jiang et al., 2018) against the pilot plant experimental results and laboratory data and is expected to achieve reliable simulation results for the air capture process. Please refer to the **Table 1** in the paper published by Li et al. (2016) for further detail about the MEA model.

A series of simulation activities were undertaken to examine the effect of various technical variables on the reboiler duty and equipment size to determine the optimized conditions for a standard MEA-based air capture process. In this optimization, the aim is to minimize the reboiler duty at any given conditions, for example different capture rates etc. The generic simulation conditions used for the CO₂ absorption and desorption are provided in **Table 2**. Technical parameters, i.e., CO₂ capture rate, rich CO₂ loading, lean CO₂ loading and air relative humidity were varied in the simulations, within the ranges shown in **Table 2**. The process was simulated based on the production of 148.25 Nm³/h of CO₂ from the air (0.291 ton/h at standard conditions). This production size determined by a potential application aiming to produce synthetic methane from the output of a 2.7 MW electrolyser that has around 78% efficiency (593 Nm³/h H₂).

Once the optimized conditions were determined using the standard absorbent regeneration process, the technical performance was further optimized using the rich-split configuration and the addition of a wash water column (**Figure 1C**) to the absorber, providing a technical basis for the economic evaluation of the MEA-based air capture process.

Economic Evaluation

With the determination of technical parameters and equipment size of the air capture process, the Aspen Capital Cost Estimator (ACCE) was utilized to estimate the direct costs for the equipment, materials and labor for construction of the air capture plant. The other costings associated with the supportive materials and labor, facilities, engineering, contractors and contingencies for plant construction were estimated based on the direct costs using multiplicative factors. These multiplicative factors were derived from a 2007 report from the Department of Energy, USA (DOE/NETL, 2007) for estimating the capital costs of an early post combustion carbon capture (PCC) plant. The assumptions for capital costs estimation and input data for the air capture process are shown in **Table S1**. For the costings of plant operation and maintenance, the annual maintenance cost was assumed to be 3% of total capital costs, and 30% of maintenance cost was assumed to be the labor cost (Socolow et al., 2011). The other plant and financial assumptions and economic information for the base case, such as plant capacity, plant life, discount rate, energy price etc. are also provided in **Table S1**.

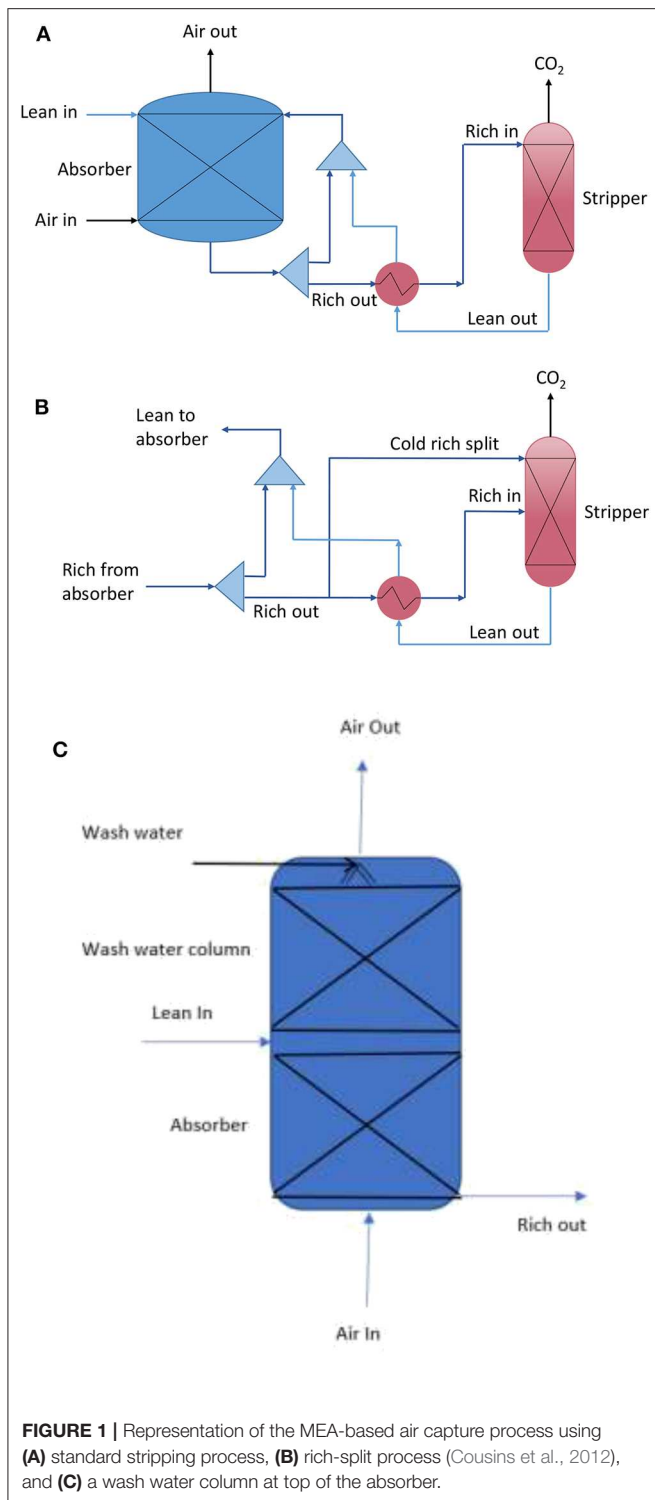
RESULTS AND DISCUSSION

Technical Performance

In this section, the effects of various parameters on the air capture process are reported. The heat of regeneration and the size of absorber are mainly discussed as they are often the two major costs in CO₂ capture process. It should be mentioned again that this process aims to capture a specific amount of CO₂ from air (around 0.291 t/h).

Capture Rate

Here it is assumed that a constant amount of CO₂ was produced from ambient air using the MEA-based absorption process as the capture rate of the process varied from 20 to 90%. The simulation results revealed that at a capture rate of 90% the reboiler duty was 21.9 GJ/tCO₂, almost double the one required in a process with



a capture rate of 20% (Figure 2A). This is due to the fact that at lower capture rates, the rich loading is higher, and this results in a lower relative vaporization heat required in the desorber. The decrease in reboiler duty would reduce the operating cost of the air capture process.

TABLE 2 | Conditions for the MEA-based CO₂ capture process simulation.

Parameters	Value
Generic conditions	
CO ₂ concentration in inlet air, ppm	400
Inlet air temperature, °C	25
CO ₂ capture rate, ton/h	0.291
MEA concentration, wt%	30
L/G for absorber with circulation, ton/ton	2.54
Packing materials in absorber and desorber	Mellapak 250X
Flooding capacity of absorber/desorber (%)	70/65
Desorber pressure (bar)	2
Temperature approach of cross heat exchanger, K	10
Variable conditions	
Capture rate, %	20, 35, 50, 70, 90
Rich loading, mol/mol	0.27, 0.30, 0.33, 0.36
Lean loading, mol/mol	0.15, 0.20, 0.25
Air relative humidity, %	30, 50, 70, 90

Figure 2B shows that the absorption column height required for capturing the specific amount of CO₂ in a system with 90% capture rate was almost 6 times greater than that required in a system with 20% capture rate. This is due to the longer contact time between liquid and air required in the system with 90% capture rate. It should be considered that in order to produce a specific amount of CO₂ with an air capture system at low capture rate, a significantly larger volume of air needs to be processed. This results in a larger diameter absorption column (Figure 2). This suggests that considering a capture rate of around 50% may be more feasible in terms of cost effectiveness of the air capture process. It should be noted that the optimum capture rate can vary with the economic parameters such as electricity and heat prices. For example, if electricity and heat are available at low cost, the optimum capture rate can be >50%. A summary of operating conditions and results at different capture rates are indicated in Table S2.

An air capture system operating at a low capture rate can also result in an extra number of energy and cost components, such as much higher MEA loss, a more significant circulation rate of rich absorption liquid to the absorber, and a significantly high energy requirement for air movement through the absorption column.

CO₂ Loading

The effect of the CO₂ loading of lean and rich absorption liquids on the reboiler duty and the size of the absorber were examined at a 50% capture rate. Figure 3A shows that the reboiler duty decreased by about 50% with increasing the lean loading from 0.15 to 0.25. With increasing lean loading, the loading in the recirculation of the absorber increased as well, leading to a lower regeneration duty. Higher rich loadings led to a lower ratio for water vapor to CO₂ in the desorber, hence it reduces the heat requirement for vaporization in the desorber. However, as the lean loading and consequently the liquid loading in the absorber recirculation increased, the lower mass transfer driving

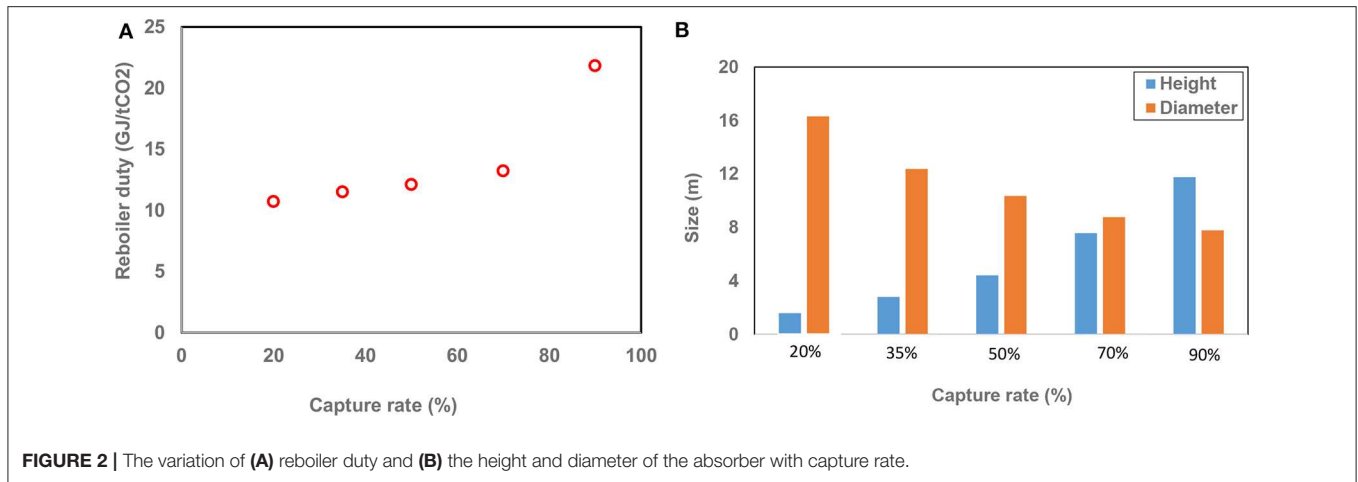


FIGURE 2 | The variation of (A) reboiler duty and (B) the height and diameter of the absorber with capture rate.

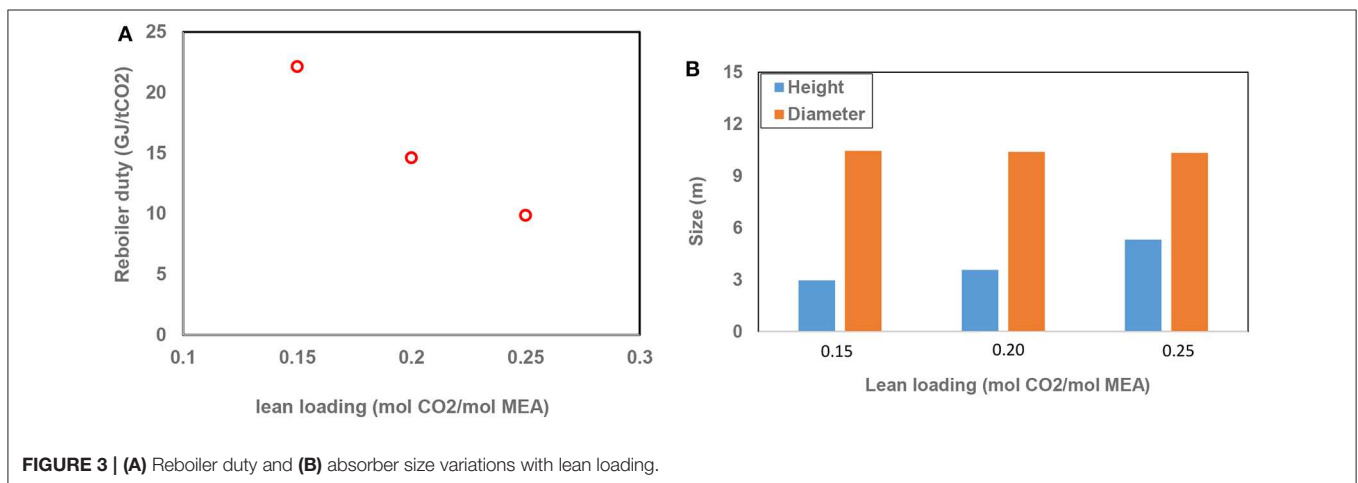


FIGURE 3 | (A) Reboiler duty and (B) absorber size variations with lean loading.

force led to a taller absorption column. **Figure 3B** also shows that the column diameter which is mainly dependent on the hydrodynamics of the system expectedly remained unchanged with lean loading. Again, the loss of MEA was significantly higher at higher loadings due to the large volume of air processed. **Table S3** lists the simulation's parameter and results for this case study. It is noted that the air humidity of 90% considered in this case changed the results slightly compared to the case when air humidity is 100%.

The effects of CO₂ loading of rich absorption liquid on the performance of the process were also investigated. In this set of conditions, a lean loading of 0.2 and a humidity of 100% were used. As evident in **Figure 4**, the heat of regeneration in the reboiler decreased from 24.1 GJ/tCO₂ to about 12.1 GJ/t CO₂ with increasing the rich loading from 0.26 to 0.35 mol CO₂/mol MEA, which is due to the lower relative vaporization in the desorber at higher rich loadings. As expected, the height of the absorber increased with increasing rich loading.

Air Humidity

The effect of humidity of inlet air on the air capture process was also evaluated. This is important as the very large volumes of air that pass through the absorber may lead to high rates

of water vaporization, depending on humidity of the inlet air. The results indicate that the reboiler duty decreased from 14.4 GJ/tCO₂ to around 9.9 GJ/tCO₂ when the relative humidity of air increased from 30 to 90% (**Figure 5**). In the low humidity cases, a massive amount of water was evaporated, which resulted in an absorption liquid containing a higher concentration of MEA. At the fixed pressure in the desorber, the reboiler temperature and the required heat of regeneration consequently increased. **Figure 5** also shows that the water loss was around 13 t/CO₂ when the air relative humidity was 30% while it decreased to 2.5 t/CO₂ when a 90% humid air was used. This suggests that preventing water losses would be beneficial when processing relatively dry air. This could be done by applying appropriate water separation system to recover the lost water. Otherwise, there would be a substantial water supplement for the air capture system, in turn increasing the costs of the air capture process.

The humidity of the inlet air had no effect on the height and diameter of the absorber.

Reboiler Temperature

Next, the effect of reboiler temperature on the performance of the air capture process was examined. A lean loading of 0.2, a rich loading of 0.35 and a capture rate of 50% were considered for this

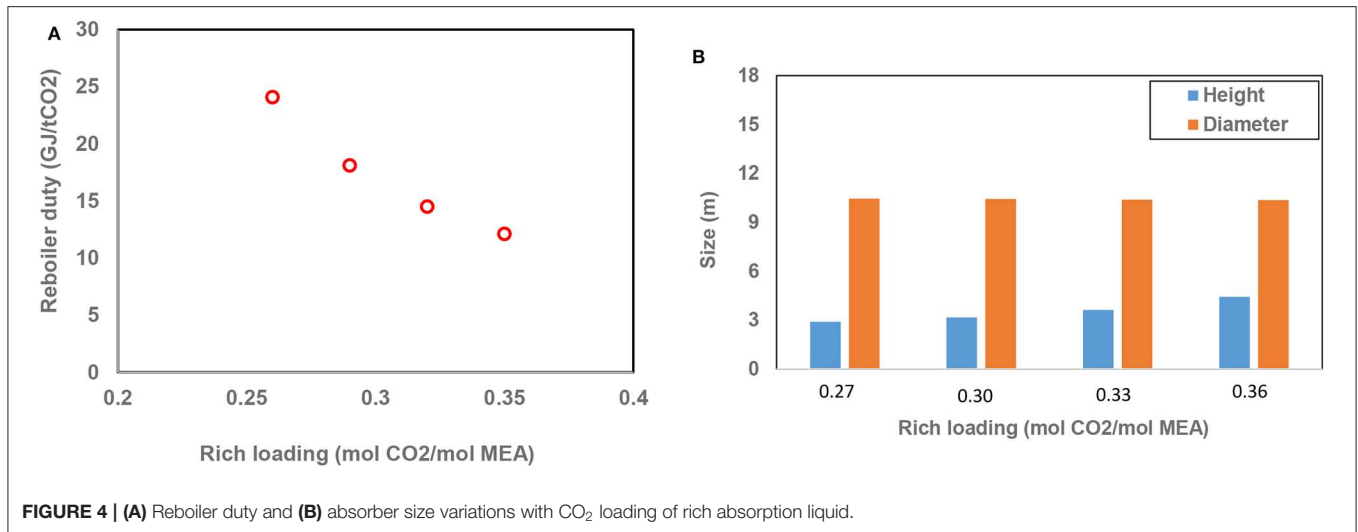


FIGURE 4 | (A) Reboiler duty and **(B)** absorber size variations with CO₂ loading of rich absorption liquid.

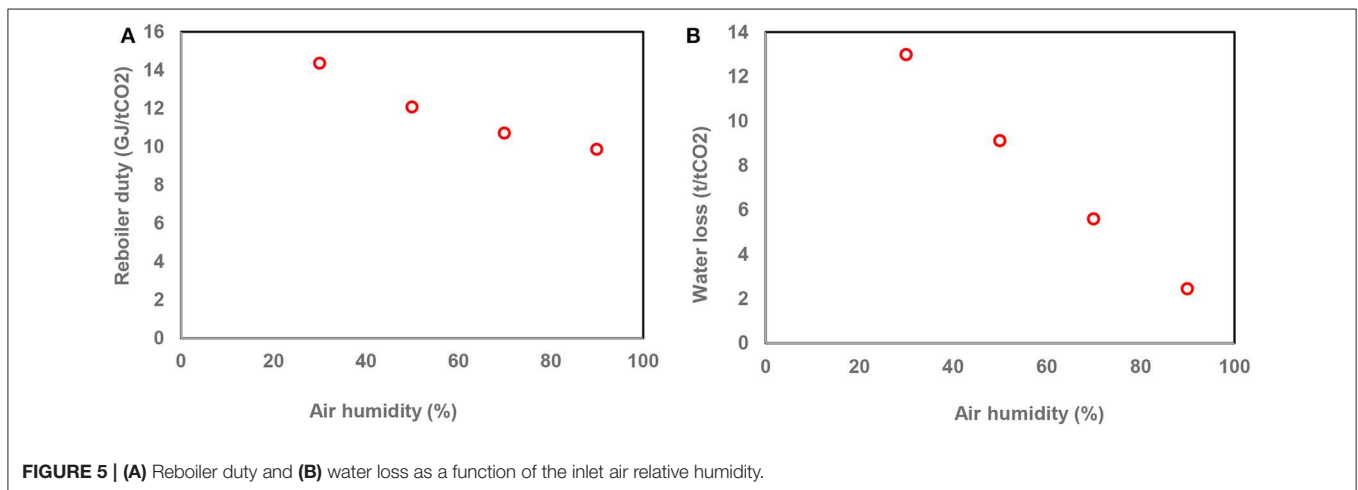


FIGURE 5 | (A) Reboiler duty and **(B)** water loss as a function of the inlet air relative humidity.

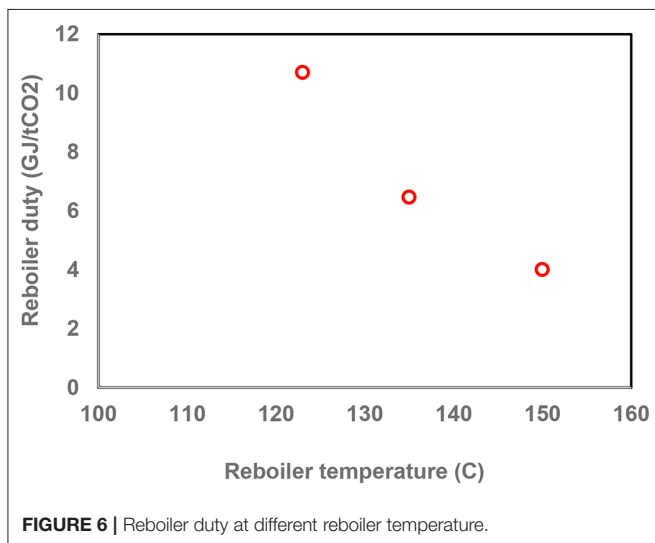


FIGURE 6 | Reboiler duty at different reboiler temperature.

case. **Figure 6** shows that the reboiler duty decreased significantly with increasing the reboiler temperature. By increasing the reboiler temperature from 123 to 150°C, the reboiler duty

decreased by more than 50%, to around 4.0 GJ/tCO₂. The operating conditions and results are given in **Table S4**. The results indicate the importance of alternative absorption liquids to have a higher resistance to thermal degradation. MEA does not lend itself for regenerator temperatures much higher than higher than 120°C. However, there are other amines and amine formulations such as piperazine that could be used at temperature up to 160°C (Rochelle, 2016).

Benchmark Conditions for the Capture Process

According to the results obtained from the sensitivity analysis shown above, a benchmark condition was selected for the capture of CO₂ from air using the MEA absorption process. This point was based on a capture efficiency of 50%, a lean loading and a rich loading of 0.2 and 0.35, respectively, and a reboiler temperature of 123°C. The rich split configuration was used in this case study due to its superior energy performance. A detailed analysis was conducted for this base case scenario, which included determination of thermal and electrical energy requirements in this system. The total electricity consumption for air blowers and liquid pumps was calculated to be 1.452 MWh/tCO₂, and the reboiler duty was calculated to be 10.7 GJ/tCO₂. The absorber

TABLE 3 | The operating conditions and process simulation results for a base case scenario in the air capture process.

Operating conditions	
Inlet air temperature (C)	25
Air relative humidity (%)	100
CO ₂ production (t/h)	0.291
CO ₂ concentration in inlet air (ppm)	400
CO ₂ capture efficiency (%)	50
Flow rate of air (t/h)	943
MEA concentration (%)	30
Flooding capacity of absorber/desorber (%)	70/65
L/G in absorber	2.54
Type of packing in the absorber and desorber	M250X
CO ₂ rich loading	0.35
CO ₂ lean loading	0.2
Desorber pressure (bar)	2
Temperature approach on cold side (K)	10
Simulation results	
Absorption liquid flow to desorber (t/h)	9.113
Absorber recirculation liquid flow (t/h)	2396.5
Specific liquid flow to desorber (t/tCO ₂)	31.3
Specific absorber recirculation flow (t/tCO ₂)	8247.3
Condenser temperature (°C)	31.2
Reboiler temperature (°C)	123.1
Reboiler duty (GJ/tCO ₂)	10.70
Total electricity (MWh/tCO ₂)	1.452
MEA evaporation in absorber (kg/tCO ₂)	53.4
Wash water flow (kg/s)	612
MEA loss after water wash (kg/tCO ₂)	0.28
Equipment dimensions	
Absorber packing diameter (m)	10.36
Absorber packing height (m)	4.43
Absorber volume (m ³)	373.2
Absorber pressure drop (kPa)	0.73
Washing column pressure drop (kPa)	0.26
Washing column packing diameter (m)	10.36
Washing column packing height (m)	5
Cross heat exchanger surface area (m ²)	102.59
Desorber diameter (m)	0.544
Desorber packing height (m)	3
Desorber (m ³)	0.70
Electrical Energy Requirements	
Recirculation pump (MWh/tCO ₂)	0.504
Washing pump work (MWh/tCO ₂)	0.238
Blower work (MWh/tCO ₂)	0.703
Pumps to/from desorber (MWh/tCO ₂)	0.007
Total (MWh/tCO ₂)	1.452

height and diameter were calculated to be 6.3 and 10.4 m, respectively. A wash water section, 10.4 m in diameter and 5 m high, was also added to decrease the evaporative loss of MEA to around 0.28 kg/tCO₂ (0.04 ppm in exhaust air) during the absorption process. The desorber dimensions are significantly smaller than the absorber dimensions which reflects the large

TABLE 4 | Economic performance of MEA-based air capture. All the costs are on a basis of 1st Qtr 2016 US\$.

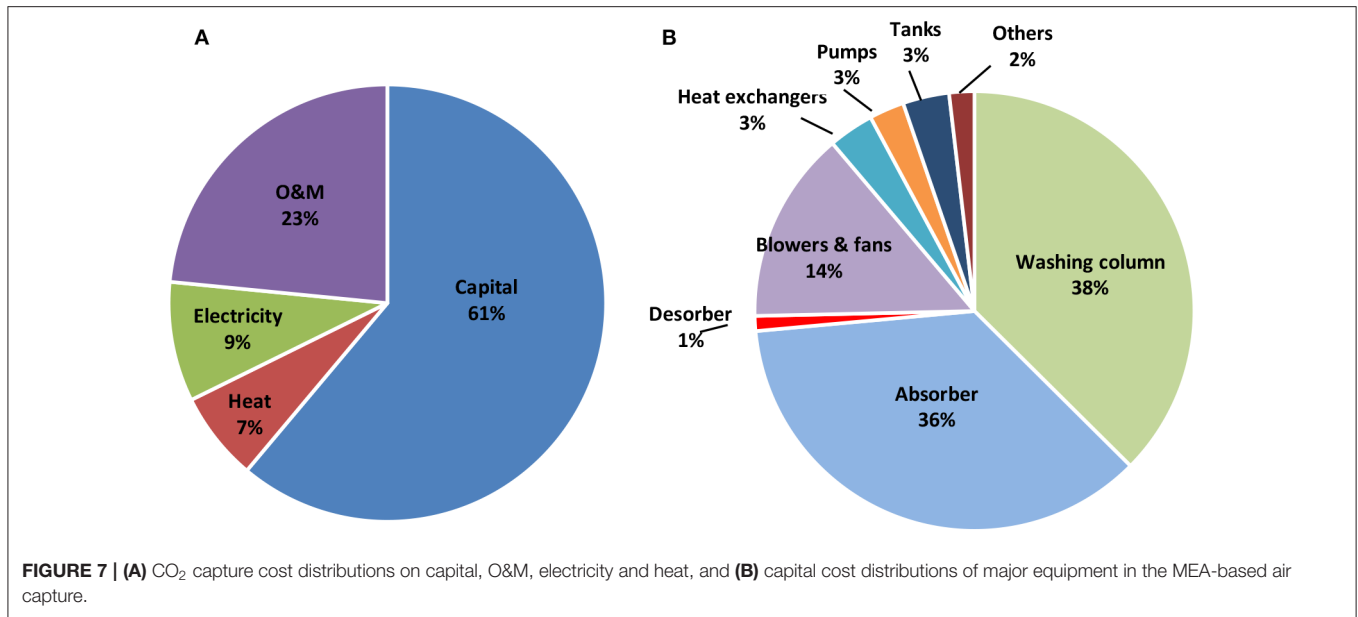
Major equipment and cost element	Cost, Million US\$
Washing column	4.38
Absorber	4.22
Desorber	0.13
Blowers and fans	1.66
Heat exchangers	0.39
Pumps	0.30
Tanks	0.40
Other equipment	0.22
Total direct costs	11.70
Total indirect costs	2.27
Engineering	1.40
Contractor fees	0.42
Contingencies	3.49
Total plant costs	19.27
Spare parts	0.096
Total investment costs	19.37
Operating expenses	
Annual O&M costs	0.757
Annual heat costs	0.213
Annual electricity costs	0.286
Capture costs	
	\$/ton CO ₂
Capital	1,033
O&M	396
Heat	111
Electricity	150
Total	1,691

difference in gas and liquid flow rates. The final simulation results for this base case scenario is indicated in **Table 3**. The break-down of required electrical energy in this process are also indicated in **Table 3**.

Economic Performance of the Base Air Capture Process

CO₂ Capture Cost—Standard Chemical Engineering Design

Using the Aspen Capital Cost Estimator with the base case design specified in **Table 3** and standard chemical engineering design specifications, the CO₂ capture cost was calculated to be \$1,691/ton CO₂ for the MEA-based air capture. It should be noted here that this estimation is based on particular capture amount of 0.291 tCO₂/h (~2,300 tCO₂/year). Considering the economy of scale (Blok and Nieuwlaar, 2017), the cost per ton of CO₂ captured will be lower for larger scale systems. **Table 4** provides the specific costs of major equipment and total investment costs, the operating expenses and the break-down of CO₂-capture costs. **Figure 7A** provides the distribution of capture costs over capital, operating expenses and energy



expenses. The capital constitutes 61% of the CO₂ capture cost, followed by 23% for O&M, 9% for electricity consumption with heat consumption being the smallest contribution at 7%. The contribution of electricity and heat to the total cost can be significantly reduced if there would be cheaper sources of electricity and heat available for the process. The sensitivity analysis around different economic parameters including the heat and electricity price will be discussed in the next section.

The total capital costs of the MEA-based air capture process were estimated to be \$19.37 million for our benchmark MEA-based air capture process. The washing column (\$4.38 million) and absorber (\$4.22 million) constituted the largest and the second largest capital item among the total capital investment and accounted for 74% of total capital costs, as shown in **Figure 7B**. Considering that a water cooling system could be difficult to implement on locations with shortage of water, air cooling was used in the present study for the cooling requirement. Due to the large amount of air requirement to meet the high cooling duty of stripper condenser and the large air flow through absorber for CO₂ absorption, the costs of blowers and fans accounted for 14% of total capital costs, being the third largest contribution.

By comparison, the annual costs of O&M and energy, i.e., electricity for blowers and pumps and heat for absorbent regeneration, are lower than the capital costs. The electricity costs (\$0.286 million/year) were higher than the heating costs (\$0.213 million/year) owing to the large electricity consumption for air transfer through the columns and solution pumping and the high circulation for CO₂-absorption and washing. It should be noted here that, in the base case, if less humid air than 100% was used, the added cost due to water evaporation in this process would be up to \$4/tCO₂, which is still negligible in comparison with other cost components shown in **Table 4**.

For the base case, a plant life of 20 years, a discount rate of 8%, a capacity factor of 90%, heat price of \$10/GJ and electricity

price of \$100/MWh were considered. A sensitivity analysis shows that with variation of equipment costs in the range of 30%, the CO₂ capture cost varied between \$1,262 and \$2,120/ton CO₂, a 34% decrease/increase with the 30% decrease/increase in capital costs. This again suggests that decreasing the process equipment costs would significantly reduce the cost of CO₂ capture from air. This can be achieved by three aspects: (1) enhancing mass transfer between absorbent and CO₂ through more efficient gas-liquid contacting in order to reduce the equipment size of CO₂ absorption, (2) absorbents that have lower evaporation of absorbent (water and amine) to the air in order to minimize the size of equipment for emission control, and (3) seeking cheaper materials for equipment manufacture in order to reduce the equipment costs. In the following section, these elements for reducing the cost of the air capture process will be explored qualitatively.

CO₂ Capture Cost—New Design

As described, the wash water section and absorber constitute 45% of total cost in the process, hence using an alternative absorption liquid with low vapor pressure and replacing the packings and absorber materials with cheaper materials can drastically reduce the total cost of the process.

– Use of alternative absorption liquid

Owing to the contact between the large volume of air and the small amount of liquid in the air capture process, MEA is evaporated during the absorption process. As discussed in section Benchmark Conditions for the Capture Process, this necessitates the use of a large wash water section after the CO₂ absorber, and hence creates an additional cost to the process. As shown in **Figure 7B**, 38% of capital cost is related to this wash water section, which in fact constitutes around 25% of total cost of the process.

Amino acid salts solutions have been proven to have a potential for absorbing CO₂ with an effectiveness that matches that of MEA (Aronu et al., 2010; Jockenhövel and Schneider, 2011; Ciftja et al., 2013; Guo et al., 2013; Wei et al., 2014; He et al., 2017). These salts do not impose any amine vapor. Based on the assumption that they show the same performance as MEA in the CO₂ capture process, if they are used in the air capture process, the need for the wash water section can be eliminated from the process. This will reduce the capital cost of the CO₂ capture process from \$1,033 to around \$640 per ton of CO₂. This will also decrease the electrical energy requirement from 1.452 MWh/tCO₂ to around 1.03 MWh/tCO₂, as the washing water pump is no longer required, and less energy will be required for blowing the air into the absorber. This will reduce the electricity cost from \$150 to around \$126 per ton of CO₂.

– Use of cheap plastic packings

Based on the techno-economic analysis in APS report (Socolow et al., 2011) and details by Carbon Engineering (Holmes and Keith, 2012; Keith et al., 2018), cost of packing materials can be up to one third of the total cost of major equipment. Using the cheap plastic packings as proposed by Carbon Engineering, the cost was reported to be 1/6 of standard stainless steel packings such as Sulzer Mellachevron 250X. Assuming the same cost for Mellapak 250X and Sulzer Mellachevron 250X, this reduces the cost of packings in this process from \$1,190,000 to around \$198,000. The absorber cost reduction is around 23% from \$372/ton of CO₂ to around \$285/ton of CO₂ (another \$87 reduction in capital cost to \$554 per ton of CO₂).

– Use of cheap materials in absorber structure

The cost of the absorber structure was reported to be around \$2,300 and \$3,700 per inlet area of absorbers (m²) for cooling tower and contactors used in Carbon Engineering's process (Holmes and Keith, 2012), respectively, while based on the APS report (Socolow et al., 2011), the cost for a conventional absorption tower is around \$15,800 per inlet area of absorber (m²). This shows that the cost of the absorber using a standard chemical engineering design specification is around 6 times more than the structure used in the cooling towers/Carbon Engineering systems. This means that if the absorption column used in this work is replaced by a cooling tower system, the cost of absorber structure will be reduced by 83% from \$30,32,800 to around \$505,460. This reduces the absorber cost by another 60% to \$62 per ton of CO₂ which is close to the number reported by Carbon Engineering. The total capital cost contribution of capturing CO₂ from air will be decreased to \$317 per ton of CO₂. It is assumed that the cheaper absorber equipment materials used are resistant toward to the amino-acid based absorption liquids.

The details of techno economic result based on the new design are given in **Table 5**. The final cost of capturing CO₂ from air is \$676/ton of CO₂, which is close to the cost reported by Climeworks (Climeworks press release, 2017; Marshall, 2017) (\$600/ton of CO₂) and in the APS report (Socolow et al., 2011) (\$610/ton of CO₂).

Figure 8 shows different cost components of the new liquid-based absorption process that was designed based on new

TABLE 5 | Economic performance of an improved amine-based air capture process.

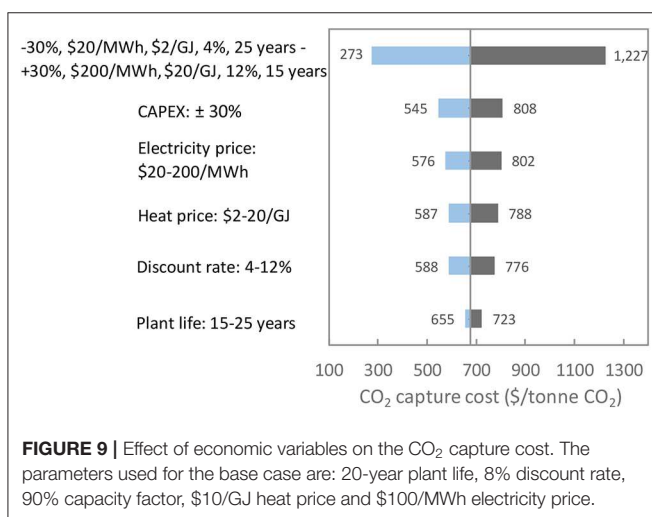
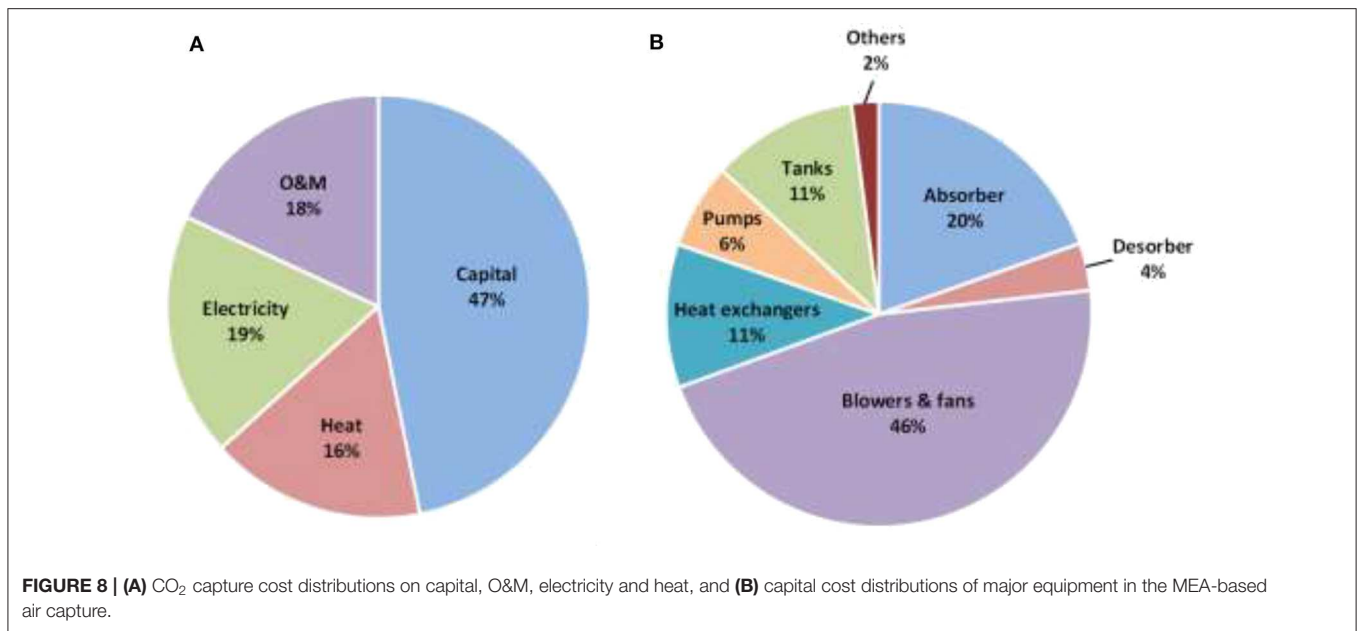
Major equipment and cost element	Cost, Million US\$
Absorber	0.70
Desorber	0.13
Blowers and fans	1.66
Heat exchangers	0.39
Pumps	0.23
Tanks	0.40
Other equipment	0.07
<hr/>	
Total direct costs	3.59
Total indirect costs	0.70
Engineering	0.43
Contractor fees	0.13
Contingencies	1.07
Total plant costs	5.91
Spare parts	0.03
Total investment costs	5.94
<hr/>	
Operating expenses	
<hr/>	
Annual O&M costs	0.233
Annual heat costs	0.213
Annual electricity costs	0.241
<hr/>	
CO₂-capture cost	\$/ton CO₂
Capital	317
O&M	122
Heat	111
Electricity	126
CO ₂ capture costs	676

The parameters used are: 0.291 t/h capture capacity, 20-year plant life, 8% discount rate, 90% capacity factor, \$10/GJ heat price and \$100/MWh electricity price. All costs are on a basis of 1st Qtr 2016 US\$.

absorption liquids, packings and absorber materials. In this new design, capital investment is still the main contributor to the total cost of the process. Among the major equipment, the cost of blowers and fans for moving the massive amount of air has now become the major part in the capital investment.

Sensitivity Analysis

The parameters used for the base case economic analysis are: 20 year plant life, 8% discount rate, 90% availability factor, \$10/GJ heat cost and \$100/MWh electricity cost. The effects of various economic parameters, such as plant life, heat and electricity unit cost, discount rate and equipment costs for the air capture process, on the economic performance were analyzed, with the results summarized in **Figure 9**. With variation of equipment costs in the range of 30%, the CO₂ capture cost varied between \$545 and \$808/ton CO₂. The discount rate also has a significant impact on the CO₂ capture cost. With a discount rate ranging from 4 to 12%, the cost of CO₂ capture ranged from \$588 and \$876/ton CO₂. Here, increasing plant life from 15 to 25 years, the cost of CO₂ captured varied by 10%, from \$655 to \$723/ton CO₂.



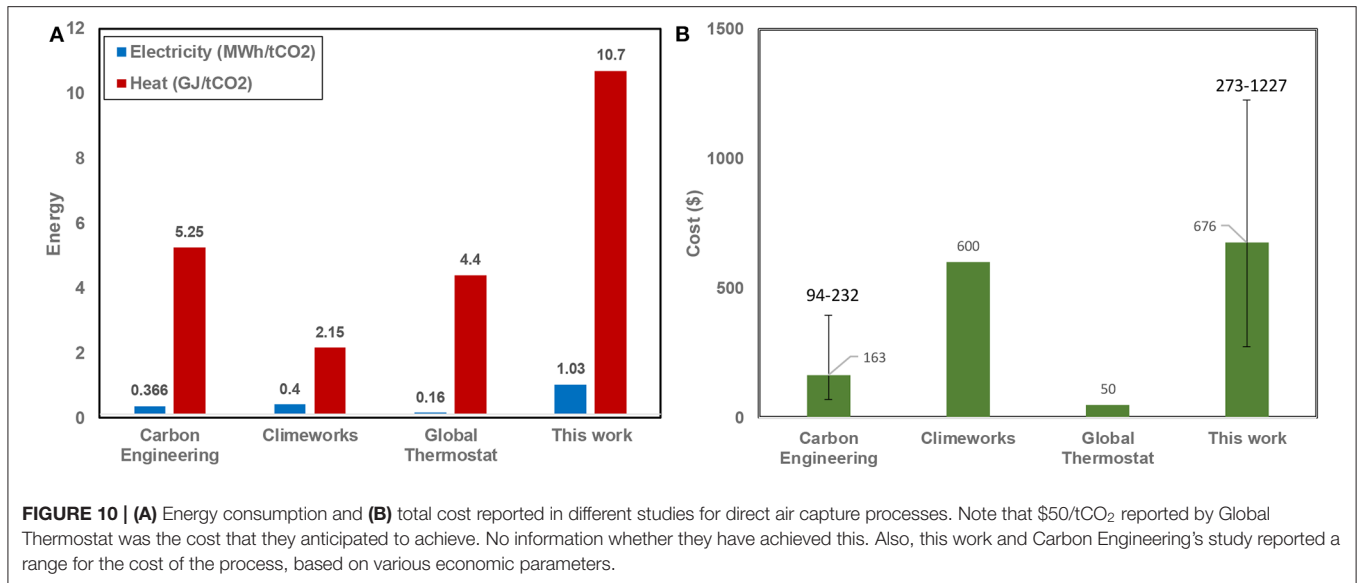
A wide range of electricity prices from \$20 to \$200/MWh and heat prices from \$2 to \$20/GJ were considered in this study, providing various possibilities of electricity and heat sources applicable to the air capture process. The CO₂ capture cost varied by 39% from \$576 to \$802/ton CO₂ and 34% from \$587 to \$788/ton CO₂, as the prices of electricity (\$20–\$200/MWh) and heat (\$2–\$20/GJ) varied by a factor of 10, respectively. The use of zero-carbon energy, e.g., renewables or nuclear, in the air capture process is most likely required to avoid any additional CO₂ emission to air due to the energy consumption for the capture process.

Based on a sensitivity analysis on all variables, considering the most desirable and the most undesirable values for various economic parameters, the overall cost of capturing a ton of CO₂ ranges from \$273 to \$1227.

Energy and Cost in Comparison With Other Air Capture Studies

There are only a few studies that reported the required energy and cost for the capture of CO₂ from air. Using different technologies such as absorption and adsorption, the required thermal energy for the direct capture of CO₂ from air was reported to be almost 4–8 times larger than the required electrical energy. The cost of capturing one ton of CO₂ was reported to be between \$100–1,000 (Keith et al., 2005, 2018; Heidel et al., 2011; House et al., 2011; Kulkarni and Sholl, 2012). The difference between the ranges reported for capture cost in this study and in literature originates from the fact that both were estimated using different technologies and economic assumptions around interest rates, plant life, electricity price and heat price etc. **Figure 10** shows the required electrical and thermal energy and total cost estimated in this work in comparison with the other air capture systems. It should be noted that there is insufficient detail on the energy consumption and cost of the other technologies that are still under development (shown in **Table 1**).

For the baseline case chosen in this study, the thermal energy required is 10.7 GJ/t CO₂ which is almost 3 times greater than the required electricity (~1.03 MWh/tCO₂). These are significantly larger than those reported in other studies. This is mainly due to different characteristics of this technology compared to others, for example the large recirculation of rich solvent is required here to prevent loading/flooding in the absorption column (it constitutes around 50% of total electricity requirement). **Figure 10** also shows that the total cost for capturing one ton of CO₂ from air using the new liquid absorption process was estimated to be around \$676, in compared with the cost reported by the other researchers using their technologies (\$93–232 in CE, \$600 in CW, and \$50 in GT). This includes \$317 and \$359 related to capital cost and operating cost, respectively.



As previously indicated in **Table 1**, the only technology in the market that uses a liquid sorbent for air capture is the one developed and demonstrated by Carbon Engineering, based on sodium hydroxide, with subsequent transfer of CO₂ from sodium carbonate to calcium carbonate and CO₂ recovery via a calcination process. To compare, our estimated CO₂ capture cost is higher than the reported results of \$94–232/tCO₂ from Carbon Engineering. This is attributed both to the higher capital costs (\$255–379/tCO₂) and operating cost (\$290–\$430/tCO₂) in our system compared to the costs for Carbon Engineering one (capital: \$64–\$120/tCO₂, the operating cost: \$30–\$110/tCO₂).

The design of the Carbon Engineering gas-liquid contactor reduces the energy requirement and capital costs of blowers and fans. They use a horizontally oriented crossflow cooling tower, resulting in a low pressure drop, while in our study, we considered a vertically counter-current flow packed tower, significantly increasing the energy and capital costs of air movement in our system.

Based on the cost analysis described above, the other areas that can be improved in order to further reduce the capital and operating cost in the air capture system are:

1. If new air-liquid contactor is designed to optimize the contact between the large volume of air and small amount of liquid, this will eliminate the necessity for the large circulation rate/pumping of absorption liquids which consumes 0.54 MWh/tCO₂ electricity (around 50% of total electricity requirement) and hence reduce the capital and operating cost. In this case, the electricity cost reported in **Table 5** can be reduced by 50% to around \$63.
2. Development of new liquid absorbents and applying process modifications. If the reboiler temperature can be increased to 150°C, the absorbent heat requirement will be reduced by 60% to around 4 GJ/tCO₂. The cost for heat reported in **Table 5** will be therefore reduced by 60% to around \$44. This in fact necessitates new absorption

liquids that are more stable at high temperature compared with MEA.

CONCLUSION

The technical and economic aspects of a conventional MEA-based absorption process for capturing CO₂ directly from air was simulated using a rate-based model in Aspen Plus. A benchmark condition was defined and further explored through a sensitivity analysis involving different parameters. It is concluded that:

- In an air capture system with a low capture rate, the heat required for the regeneration of absorption liquid per ton of captured CO₂ is significantly lower. However, a large volume of air needs to be processed and this results in a larger unit and hence a higher capital cost.
- Owing to the very low concentration of CO₂ in air, the amount of liquid required for the contact with air would be extremely low. To meet the operational requirement of a conventional packed column, a large liquid circulation around the absorber is required, resulting in a large electricity requirement.
- At the base case scenario, a capture rate of 50% was selected. The reboiler duty was 10.7 GJ/tCO₂ and the electrical energy requirement was 1.4 MWh/tCO₂. The absorber diameter and height were 10.4 and 4.4 m, respectively and the wash water section was 5 m in height.
- Owing to the large volume of air used in the air capture process, water and MEA losses due to evaporation were quite high. The addition of a wash water section to the absorber can reduce the MEA evaporation significantly down to 0.28 kg/ton of CO₂ (0.04 ppm), at the expense of increased capital costs.
- Using the standard chemical engineering design which may not necessarily be required for the air capture process, the total estimated cost for this process was around \$1,690/ton CO₂. The capital cost and operating cost were \$1031/ton CO₂ and \$659/ton CO₂, respectively.

- The wash water section and absorber accounted for around 74% of total capital cost of this process. Using an alternative absorption liquid with negligible vapor pressure which allows for the removal of the wash water section, the capital cost will be reduced by \$393 per ton of CO₂.
- The replacement of stainless steel packing and absorption column materials by cheaper materials like those used in cooling towers reduced the capital cost by another \$323 per ton of CO₂.
- The techno-economic analysis of the air capture process designed based on the cooling tower technology and non-volatile liquid absorbent showed that the cost for CO₂ capture from air was, in the range of US\$273 to US\$1,227. Around 45% and 55% of this were attributed to the capital cost and operating cost, respectively.

In summary, this study provides an economic baseline for the air capture technologies that aim to use liquid-based capture process. Even though the energy requirement and cost of the baseline for such processes are still high with respect to the current CO₂-price, even for the CO₂-price in the framework of the Californian Low Carbon Fuel Standard (Low Carbon Fuel Standard, 2020), it is believed that the energy and equipment cost of the liquid-based air capture processes can be significantly reduced when this technology further matures and is scaled up to larger plant sizes. The use of cheap materials and more effective ways of contacting air and absorption liquids, using more robust and economical absorption liquids and innovative process modifications is required to further reduce the cost of this process. Also, if the CO₂ captured from atmosphere is utilized to produce synthetic fuels, then the process can be considered more economically favorable.

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DATA AVAILABILITY STATEMENT

The datasets for this article are not publicly available because the data ownership and control must stay with CSIRO for propriety reasons. Requests to access the datasets should be directed to Paul Feron (paul.feron@csiro.au).

AUTHOR CONTRIBUTIONS

AK drafted the manuscript and analyzed and integrated the research results. KJ performed the process modeling and did the techno-economic analysis. PF conceived and designed the research work and analyzed research results. All authors contributed to the review and editing of the manuscript.

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SUPPLEMENTARY MATERIAL

The Supplementary Material for this article can be found online at: <https://www.frontiersin.org/articles/10.3389/fenrg.2020.00092/full#supplementary-material>

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Conflict of Interest: The authors declare that the research was conducted in the absence of any commercial or financial relationships that could be construed as a potential conflict of interest.

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