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Modification of postcombustion CO_2 capture process: A techno-economic analysis

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Abstract: Postcombustion CO_2 separation using aqueous amine solution has a great potential to minimize CO_2 emissions but is expensive due to huge energy requirement for solvent regeneration. Several structural modifications can be implemented to minimize the regeneration energy requirement and optimize the energy efficiency. In this study, we evaluated the techno-economic benefits of three stripping modifications, namely lean vapor compression (LVC), stripper overhead exchanger (SOE), and an advanced hybrid configuration (LVCSOE) in a single flowsheet aimed at reducing the energy consumption of CO_2 separation process using 30 wt.% monoethanolamine (MEA) solution. All the configurations were simulated using Aspen Plus® rate-based modeling, while capital investment was evaluated using Aspen Economic Analyzer®. All the modifications reduced the energy consumption and showed economic benefit compared to the base case. The optimal configuration was LVCSOE, which reduced the energy requirement for solvent regeneration and the CO_2 capture cost by 18% and 5.3%, respectively, and can save 5.4 million USD annually. Additionally, sensitivity analysis of economic variables suggested that CO_2 capture cost is more sensitive to regeneration steam cost than any other economic parameter. Furthermore, the effect of regeneration steam cost revealed that the implementation of an advanced process configuration can result in higher net savings as compared to the base case at high fuel price. © 2020 Society of Chemical Industry and John Wiley & Sons, Ltd.

Keywords: advance process configuration; aspen simulation; CO_2 capture cost; heat integration; postcombustion CO_2 capture; techno-economic analysis

Introduction

Due to intermittent nature of renewable energy, energy-mix from fossil fuels and renewable energy sources is a feasible option to meet the future energy requirement. However, the extensive emission of carbon dioxide (CO_2), mainly from the combustion of fossil fuels to meet the energy demand, is considered a primary reason of global warming. In the coming decades, the statistics indicate an increase in combustion of fossil fuels for power generation resulting in higher level of CO_2 emission.[1] With an aim of clean power generation from fossil fuels and controlling the CO_2 emissions, extensive work is being done to develop energy-efficient and cost-effective CO_2 capture techniques. Among three major approaches established to capture CO_2, that is, pre-combustion, postcombustion, and oxy-fuel, the postcombustion

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CO_2 capture using aqueous amine solutions are the most promising because it is reliable, technically mature, and can be easily retrofitted to existing plants[2]. There are, however, several shortcomings of this approach including high capital cost, large heat duty, and facility corrosiveness.[3]

In an amine-based CO_2 capture process, the aqueous amine solution absorbs CO_2 in an absorber column at 40–60 °C. This CO_2-rich solution is then thermally treated in a stripper column at 110–130 °C to collect a pure stream of CO_2 from the top of the column. Unfortunately, the stripping operation requires substantial amount of thermal energy that accounts for around 70% of the total operating cost (TOC) of the process. The heat duty of amine solvent regeneration ($Q_{regeneration}$) consists of three main parts: (i) energy required for breaking the chemical bond between solvent and carbon dioxide ($Q_{desorption}$), (ii) energy required to increase the temperature of the aqueous amine solution ($Q_{sensible}$), and (iii) energy required to produce water vapors for CO_2 stripping ($Q_{vaporization}$).[4] Generally, $Q_{desorption}$ accounts for half of the total required energy ($Q_{regeneration}$), whereas $Q_{sensible}$ and $Q_{vaporization}$ together are responsible for rest of the $Q_{regeneration}$. Therefore, a decrease in any of the three parts can reduce the overall energy requirement.

In the past few decades, extensive research efforts have been devoted to optimize the solvent formulation and the process configuration. The main focus has been on screening or synthesizing advanced absorbents that can be readily regenerated. For instance, several new classes of absorbents including ionic liquids,[5] biphasic solvent,[6] and nonaqueous absorbents[7] have been developed to overcome the large energy penalty drawback. In addition, many absorbents including ammonia,[8] diethylenetriamine,[9] 2-amino-2-methyl-1-propanol,[10] and piperazine/ammonia,[11] have been identified as potential candidates because they have low CO_2 reaction enthalpies. However, most of the developed solvents are limited to laboratory-scale testing or have not progressed beyond the pilot plant scale because their synthesis involves a laborious process and are expensive.[12] Huge efforts are required to make them economically feasible.

On the other hand, improvement in process configuration is effective and easy because it can be done by minimum retrofitting in any existing facility. Various innovative process modifications like rich vapor configuration,[13] absorber inter-cooling,[14] stripper interheating,[15] rich solvent split,[16] stripper overhead compression,[17] and many other have been reported to reduce regeneration energy. Erik *et al.*[18] and Amrollahi *et al.*[19] compared different configurations on the basis of process's energy consumption and concluded that process modifications have significant potential to reduce the energy requirement of the process. A combination of multiple process modifications has also been reported to be effective in lowering the heat requirement.[20] Haider *et al.* studied the heat integration within postcombustion CO_2 capture process and reduced the energy consumption by 14%.[21] Perevertaylenko *et al.*[22] and Zhang *et al.*[23] sought opportunities for heat integration by applying pinch analysis and reported a reduction of 13% and 19%, respectively. Ahn *et al.* evaluated ten different configurations for CO_2 capture process and reported a reduction of 37% in steam consumption and 14.1% in net energy consumption by combining multiple configurations in a single flowsheet.[24] Hafiz *et al.* reported a reduction of 21% by combining different stripper modifications integrated with CO_2 compression process.[25] Jin *et al.* modeled a combination of five different modifications including membrane separation in a single flowsheet and reported a reduction of 28% in regeneration energy.[26] Oh *et al.* developed a superstructure by integrating UNISIM® and MATLAB® to evaluate different process modifications and reduced the energy consumption by 7%.[27] Recently, Oh *et al.* reported the reduction in energy consumption by stripper modification and concluded that increase in capital cost by process modification is inevitable, however, the energy consumption and operating cost can be saved by optimizing the process configurations.[28]

To move one step close to commercialization, the impact of various process modifications on the heat duty reduction was studied on pilot-plant scale. Commonwealth Scientific and Industrial Research Organization (CSIRO) pilot plant integrated rich solvent split modification and found that the solvent regeneration energy can be reduced by up to 7% while condenser duty can be minimized by 60%.[29] Danish Oil and Natural Gas (DONG) Energy's pilot plant integrated lean vapor compression (LVC) configuration, which resulted in a 20% reduction in the regeneration energy using MEA solvent.[30] The results of these pilot plant studies showcase the promise of process modifications in improving the energy efficiency of CO_2 capture process. However, energy

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efficiency is not the only decisive parameter. Process modifications require additional equipment, which adds to the capital cost. In some cases, the cost saved in terms of heat duty can be offset by the additional cost of equipment. This reduces the certainty of economic benefit of process modification. The additional cost might overweigh the energy saving and make the process more expensive.

Given the scenario, it is crucial to perform a techno-economic analysis of any process modification to estimate the impact on process cost and energy saving. A better index to investigate the process modifications is "CO_2 Capture Cost." Nwaoha *et al*. introduced an advanced configuration by splitting the CO_2-rich stream into four substreams and performed its techno-economic analysis for CO_2 capture from paper industry.[31] The results for MEA 30 wt.% solvent showed that the advance configuration has 17.4% lower regeneration energy than conventional process but is 7% more expensive than the conventional process. To make a holistic evaluation of post combustion CO_2 capture process integrated with power plant, Li *et al*. calculated the CO_2 capture cost for different process modifications and reported a reduction of 13.5% in regeneration energy and 11.3 \$/tonCO_2 in CO_2 avoided cost.[32]

Various configurations with techno-economic analysis have been published to reduce the CO_2 capture cost but a knowledge gap still exists for development of an optimum process configuration with lower energy requirement and capture cost along with a lower capital expenditure increment with modification. Therefore, the prime objective of this study is to propose a novel configuration, which not only reduces the energy requirement and CO_2 capture cost of the process for clean production of energy from coal fired power plant but also copes with the issue of increment in capital expenditure. The configurations studied in this work are LVC, modified stripper overhead exchanger (SOE), and a hybrid configuration (lean vapor compression stripper overhead exchanger [LVCSOE]) incorporating heat pump (LVC) and heat integration in a single flowsheet. To the best of authors' knowledge, LVCSOE has not been reported in the literature. All the configurations were simulated using Aspen Plus V10.1® rigorous rate-based modeling. A model for techno-economic analysis was presented using Aspen plus and Process Economic Analyzer in \$/2018. All the process configurations were compared with the conventional process in terms of energy requirement, capital expenditure, and capture cost. In addition to economic analysis, sensitivity analysis was also carried out to understand the impact of economic variables on CO_2 capture cost (USD/tonneCO_2).

Table 1. Flue gas specifications.

Component	Molar composition
CO_2	13%
H_2O	10%
N_2	72%
O_2	5%
Temperature	40 °C
Pressure	1.5 bar

The originality and novelty of this work includes (i) developing advanced process configuration and evaluating its performance for postcombustion CO_2 capture from coal fired power plants, (ii) performing techno-economic analysis of advanced process configurations and comparing their performance with the conventional process, (iii) studying the impact of various economic parameters, that is, regeneration energy cost, electric energy cost, discount rate, project lifespan, and capital expenditure (CAPEX) requirement on CO_2 capture cost, and (iv) studying the effect of regeneration steam cost on CO_2 capture cost for advanced and conventional configurations.

MEA based CO_2 capture process simulation

Simulation basis

The CO_2 capture process was designed for a 300 MW coal-based power plant located in the United States. The plant has a designed capacity of 1.5 million tonne CO_2 captured per annum. The flue gas composition was taken from Department of Energy (DOE) guidelines for CO_2 capture process for coal-fired power plant. The composition and the operating conditions of flue gas are summarized in Table 1. The ENRTL property package and Redlich–Kwong (RK) equation of state were used for modeling the liquid phase and vapor phase, respectively. Rate-based models were used for the absorber and stripper. FLEXIPAC 250Y was used as packing material in the absorber and stripper columns. With a capture efficiency of 90%, the CO_2 removal rate was set at 190 tonne/h. At the top of the stripper, the collected CO_2 had a purity of 99 wt.% and was then compressed to 150 bar. The absorber and

Table 2. Specification of CO_2 capture process.

Component	Specification
Absorber height	24 m
Absorber diameter	13 m
Stripper height	11 m
Stripper diameter	8 m
Flue gas flow rate	300 kg/s
Amine flow rate	2900 tonne/h
CO_2 capture rate	190 tonne/h
CO_2 capture efficiency	90%
CO_2 product purity	99 wt.%

Table 3. Reaction kinetics.

Reaction No.	*k*	*E* (cal/mol)
4	4.32e+13	13 249
5	2.38e+17	29 451
6	9.77e+10	9855.8
7	3.23e+19	15 6554

stripper are operated at 80% flooding to calculate the diameter of columns. The amine flow rate and absorber height are optimized to achieve 90% CO_2 removal efficiency from flue gas. The specifications mentioned in Table 2 were optimized and kept constant in all configurations.

Process modeling description

Thermodynamically H_2O-MEA-CO_2 mixture is represented by ENRTL and is implemented in Aspen Plus®.[33] The reactions that take place in the process are given in Eqn 1–7. The kinetic parameters of these reactions are given in Table 3. Henry's constant for CO_2 in water are obtained by regressing with binary VLE data,[34] and for CO_2 in MEA are obtained from Wang *et al.*[35] The NRTL interaction parameters between MEA and H_2O are obtained from Austgen *et al.*[36] NRTL interaction parameters between CO_2 and H_2O are set zero. The detailed modeling parameters of the absorber and stripper are given in Table 4.

$$H_2O + MEAH^+ \leftrightarrow MEA + H_3O^+ \rightarrow \text{Equilibrium} \quad (1)$$

$$2H_2O \leftrightarrow H_3O^+ + OH^- \rightarrow \text{Equilibrium} \quad (2)$$

$$HCO_3^- + H_2O \leftrightarrow CO_3^{2-} + H_3O^+ \rightarrow \text{Equilibrium} \quad (3)$$

Table 4. Parameters used for the modeling of absorber and stripper.

Parameter	Absorber	Stripper
Calculation types	Rate based	Rate based
No. of stages	20	20
Packing type	FLEXIPAC	FLEXIPAC
Reaction condition factor	0.9	0.9
Film resistance	Discretized film	Discretized film
Film discretization ratio	5	5
Mass transfer coefficient method	Brf-85	Brf-85
Heat transfer coefficient method	Chilton and Colburn	Chilton and Colburn
Interfacial area method	Brf-85	Brf-85
Holdup correlation	Stichlmair	Stichlmair
Flow model	Mixed	Mixed
Condenser	None	Partial condenser
Reboiler	None	Kettle type reboiler

$$CO_2 + OH^- \leftrightarrow HCO_3^- \rightarrow \text{Kinetic} \quad (4)$$

$$HCO_3^- \leftrightarrow CO_2 + OH^- \rightarrow \text{Kinetic} \quad (5)$$

$$MEA + CO_2 + H_2O \leftrightarrow MEACOO^- + H_3O^+ \rightarrow \text{Kinetic} \quad (6)$$

$$MEACOO^- + H_3O^+ \leftrightarrow MEA + CO_2 + H_2O \rightarrow \text{Kinetic} \quad (7)$$

The CO_2 capture model developed on Aspen Plus® has been validated by comparing the temperature profile of the absorber with experimental data from CESAR pilot plant test campaign.[37] Figure 1 shows the comparison of temperature profile from simulation results and experimental data.

Process configurations

Base case

The flowsheet of the conventional CO_2 capture process is shown in Fig. 2. MEA 30 wt.% enters from top while the flue gas enters from the bottom of the absorber

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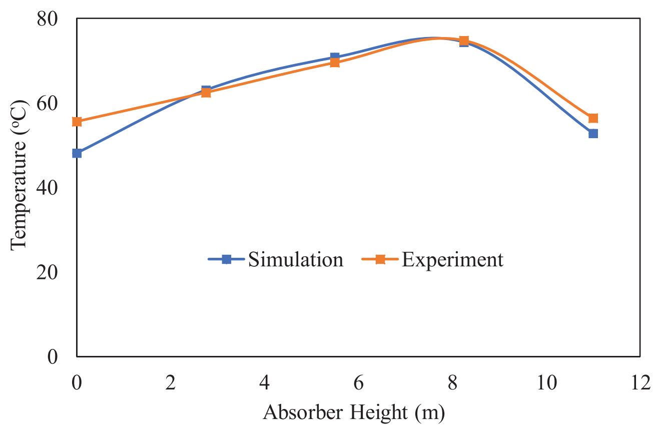

Figure 1. Validation of CO_2 capture unit.

column. The MEA solvent absorbs CO_2 and the CO_2-rich solution is pumped to Rich-Lean heat exchanger before entering the stripper column. In the stripper column, the CO_2-rich solution is heated using a reboiler and the CO_2 along with the water vapors leaves from the top of the column. A condensing unit is used to cool the gas stream and a flash column is used to separate the condensable species. From the flash column, pure stream of CO_2 is compressed to around 150 bar. The hot CO_2-Lean solution that leaves from the bottom of the stripper column exchanges heat with the CO_2-rich solution in the Rich-Lean heat exchanger and is further cooled before going back to the absorber. CO_2 compression unit was integrated with all the configuration to deliver compressed CO_2 at 150 bar pressure and 99.5 wt.% purity. The CO_2 compression unit is shown in Fig. 3.

Stripper overhead exchanger

The SOE modification proposed by Gelowitz *et al.* enhances the heat integration of the process.[38] The configuration is a modified form of CO_2-rich solvent split. In contrast to base case, the CO_2-rich stream coming from the bottom of the absorber column is divided into two streams. One of the streams exchanges heat in the SOE with the H_2O/CO_2 vapors leaving at high temperature from the top of stripper, while the other stream enters the lean-rich heat exchanger (LR-HX). The solvent from SOE enters the stripper and the H_2O/CO_2 mixture moves to condenser. A part of the heat that was being consumed in condenser is now coming back to stripper. This modification reduces the heat duty of reboiler, condenser, and LR-HX. The configuration is shown in Fig. 4. The added streams are highlighted in red color while added equipment are colored yellow.

Lean vapor compression

Lean vapor configuration is an application of heat pump. A heat pump improves the thermal energy by adding a small amount of high-quality mechanical energy into the system. In the conventional base case process, the hot CO_2-lean solvent leaves from the bottom of the stripper and enters to Lean-Rich HX, while in LVC, the lean solution is flashed in a flash column removing additional CO_2 from the lean solvent. The CO_2 and the water vapors leaving from top of the flash column are compressed to stripper pressure and sent back into the stripper column. The compressed vapors are at high temperature. They not only add thermal energy in the column but also act as a stripping agent, thus reducing the reboiler duty. Additional electrical energy is required to drive the compressor. Pumping of lean solvent also requires more power due to pressure loss in flash column, thus electrical energy requirement in LVC is higher. The liquid leaving from the bottom of the flash column contains lower amount of CO_2, thereby increasing the absorption capacity of solvent. The CO_2-lean solvent is pumped to LR-HX as in conventional process. The modification is shown in Fig. 5. The added streams are highlighted in red color while added equipment are colored yellow.

Lean vapor compression with stripper overhead exchanger

This advanced hybrid configuration incorporates both heat pump (LVC) and process heat integration (SOE) in a single flowsheet to exploit the benefits of both the LVC and the rich split flow with overhead heat exchanger. Although this combination slightly increases the complexity of the process, but the heat duties of the LR-HX, condenser, and reboiler decrease significantly, which reduces the cost of heat exchangers. The configuration is shown in Fig. 6. The added streams are highlighted in red color while added equipment are colored yellow.

Economic analysis approach

Capital expenditure

The cost of the equipment has major contribution in the capital expenditure of any plant. Among various methods to estimate the equipment cost, the most

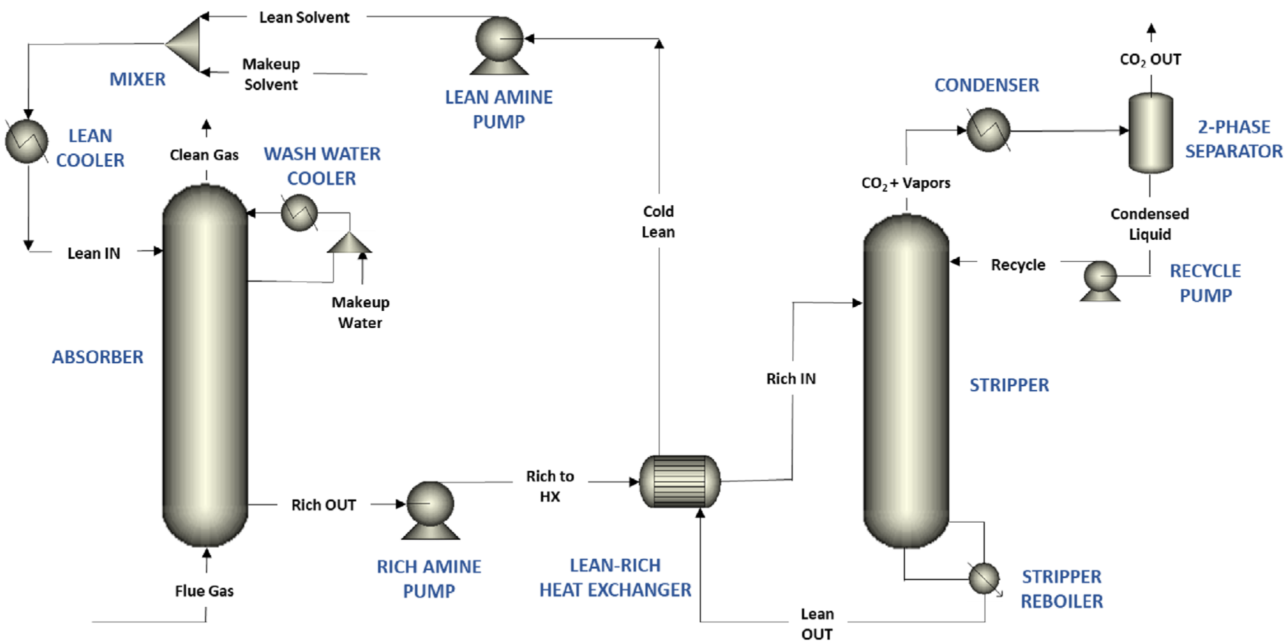

Figure 2. Base case flowsheet for MEA-based CO_2 capture process.

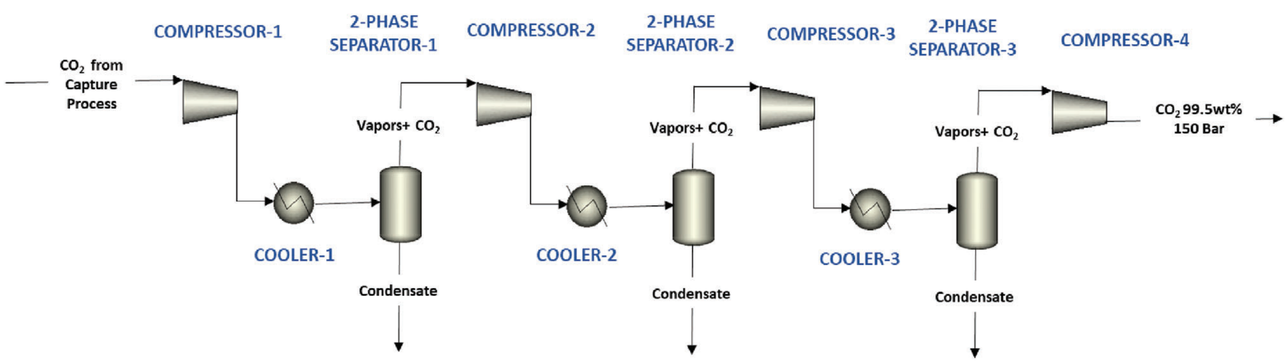

Figure 3. CO_2 compression unit (multistage compression up to 150 bars).

reliable method is quotation from vendors, but it requires detailed engineering drawing and data. Another method for estimating the cost of equipment is applying six-tenth rule on data acquired from engineering and procurement companies. However, such data are mostly unavailable in open literature. Another method of equipment cost estimation is using correlation and graph published in process design handbooks.[39–44]

A more reliable method to estimate equipment cost is by using cost estimation software. Aspen Tech has developed Aspen Process Economic Analyzer® that is integrated with Aspen Plus®. Aspen Process Economic Analyzer® was used in this study to estimate the cost of equipment. Li *et al*. confirmed the authenticity of Aspen Process Economic Analyzer results by comparing the results of AspenTech® software with real built plant.[32]

The methodology to estimate the capital expenditure is shown in Fig. 7. Total equipment cost (TEC) was calculated using Aspen Process Economic Analyzer. Various factors used for calculating CAPEX were taken from the reports published by US Department of Energy.[45] The bare erected cost (BEC) includes the cost of equipment, supporting facilities required for operation of equipment, and the labor cost for construction/installation of equipment and supporting facilities. Engineering procurement construction (EPC) cost is the sum of BEC and engineering/construction management cost. According to U.S. department of energy, engineering/construction management cost is 8–10% of BEC.[45] The factors for process contingency

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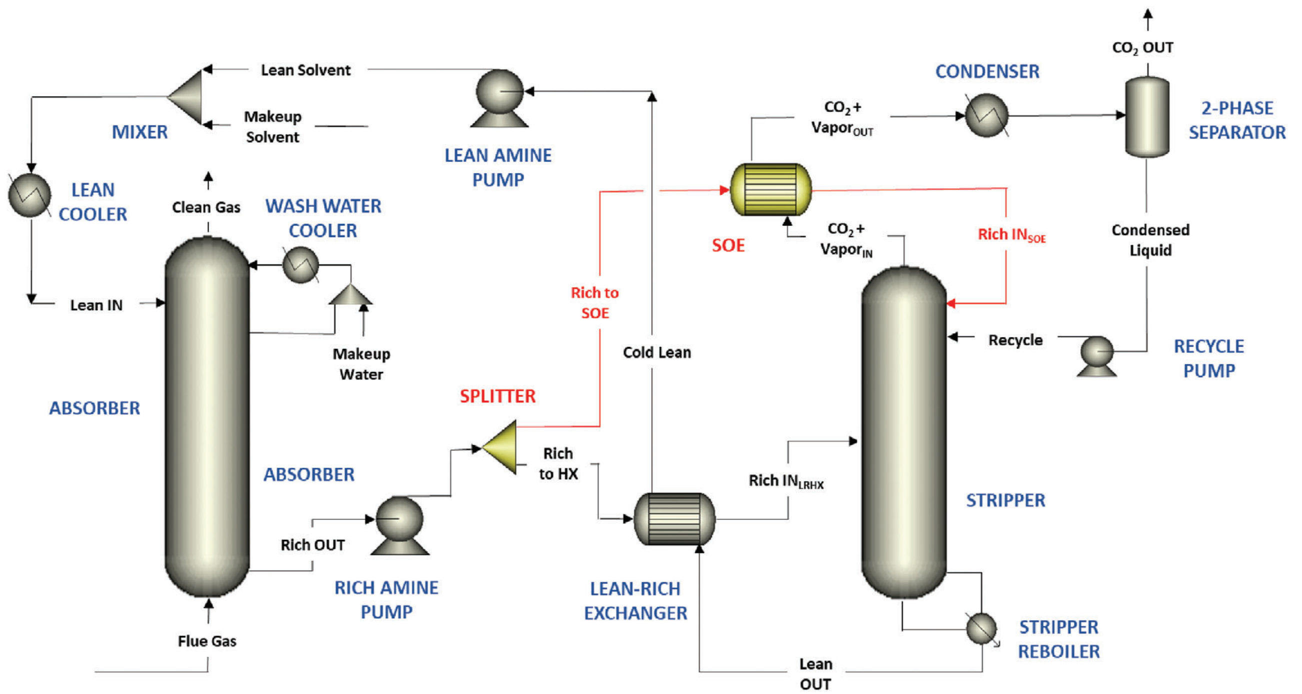

Figure 4. Stripper overhead exchanger flowsheet for MEA-based CO_2 capture process.

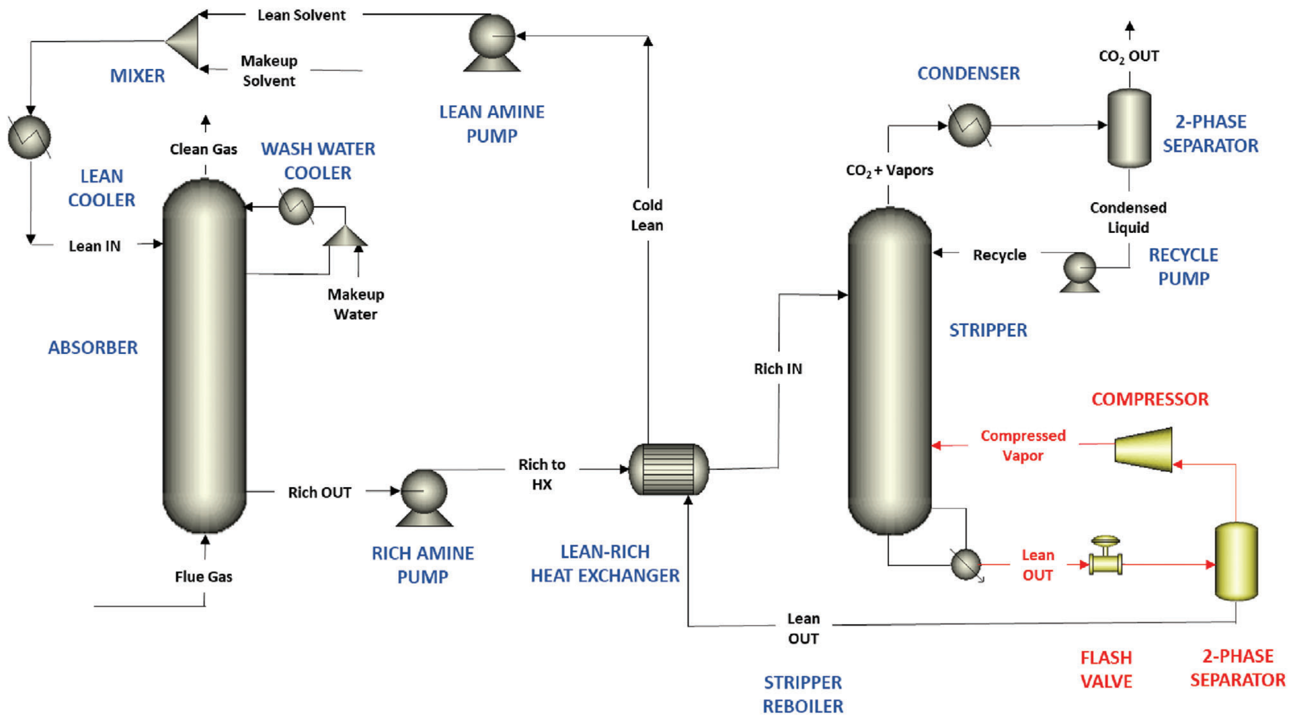

Figure 5. Lean vapor compression flowsheet for MEA-based CO_2 capture process.

and project contingency were also taken from the report of US Department of Energy.[45] Global CCS Institute published a factor of 15% of total plant cost for owner's cost estimation. The analysis is based on $2018 and all the costing data are converted in $2018.

Initial amine solution is the amount of amine solution required for startup, filling the sump of

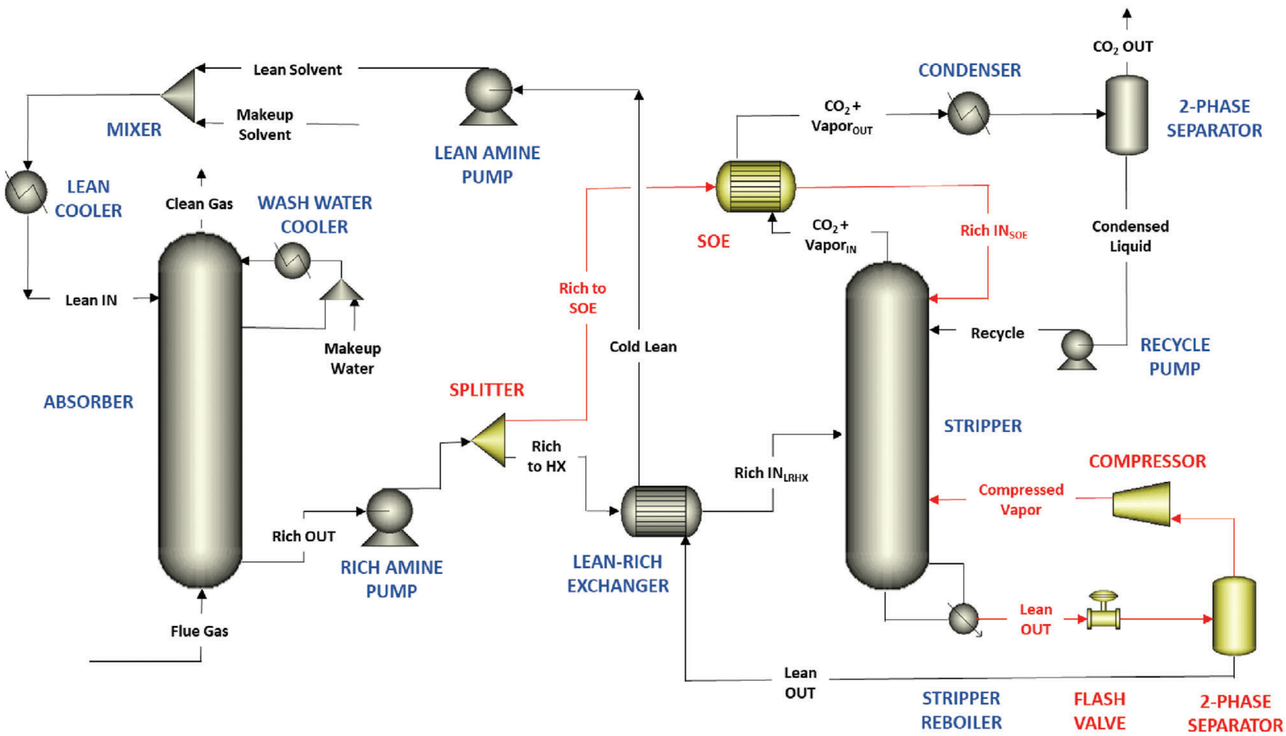

Figure 6. Lean vapor compression with stripper overhead exchanger flowsheet for MEA-based CO_2 capture process.

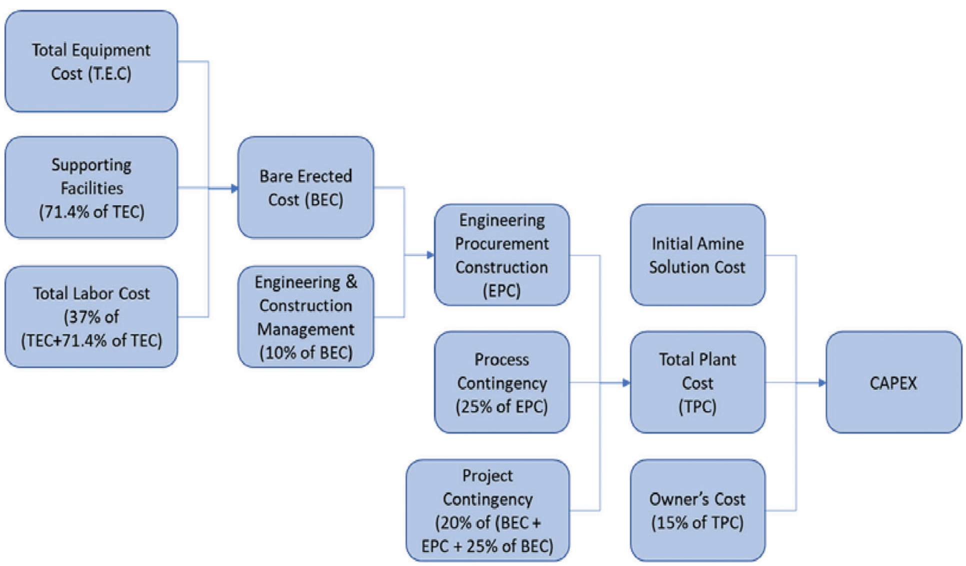

Figure 7. Economic model for capital expenditure calculation.

absorber/stripper column, flash column, amine make-up drum, heat exchangers, and pipes. The volume of amine in heat exchanger and absorber/stripper sump was calculated by multiplying the flow rate of amine solution with the retention time (from Aspen Plus Process Simulator). In order to accommodate the amount of amine solution in pipes and drums, the cost for amine solution required in sump and heat exchanger was multiplied by a factor of 1.7.[30]

Total annual cost

The total annual cost (TAC) is defined as the sum of the TOC and annualized capital expenditure

Table 5. Factors for estimating fixed operating cost calculation.

Fixed operating cost	Estimation factor
Operating labor cost	OLC
Supervision	18% of OLC
Laboratory charges	15% of OLC
Maintenance and repair	6% of FCI
Operating supplies	0.9% of FCI
Local taxes insurance	3.25 of FCI
Plant overhead expenditure	(70.8% of OLC) + (3.6% of FCI)
Administrative expenditures	(17.7% of OLC) + (0.9% of FCI)

($CAPEX_{Annual}$). TOC consists of fixed operating cost (FOC) and variable operating cost (VOC).

Fixed operating cost

FOC includes the expenditures that does not change with the rate of production. Estimation factors for calculating the FOC are given in Table 5, where FIC is fixed capital investment. Operating labor cost (OLC) was calculated using Eqn (8), where N_O is no of operators per shift and N_{np} is the total number of equipment.

$$N_O = \left(6.29 + 0.23N_{np}\right)^{0.5} \quad (8)$$

$$N_{np} = \sum_{\substack{Compressors\\ Towers\\ Heat\ Exchangers\\ Reactors}} \text{Equipment} \quad (9)$$

The calculation procedure for the OLC and estimation factors for fixed operating expenditure were taken from the work of Turton *et al.*[39]

Variable operating cost

VOC includes the expenditures that vary with the rate of production, that is, utilities and make-up chemicals. The main utilities in the CO_2 capture plant are cooling water, regeneration steam, and electric energy. Amine make-up and water make-up expenditures are also included in VOC. The unit cost of utilities is 0.212 \$/GJ cooling water, 0.0775 \$/kWh electric energy, and 8 \$/GJ of regeneration steam. MEA is purchased at 2200 \$/tonne[31] while makeup waster cost is 1.25 \$/tonne.

Annualized capital expenditure

$CAPEX_{Annual}$ is the annual payback of CAPEX. CAPEX is converted into fixed annual payment series

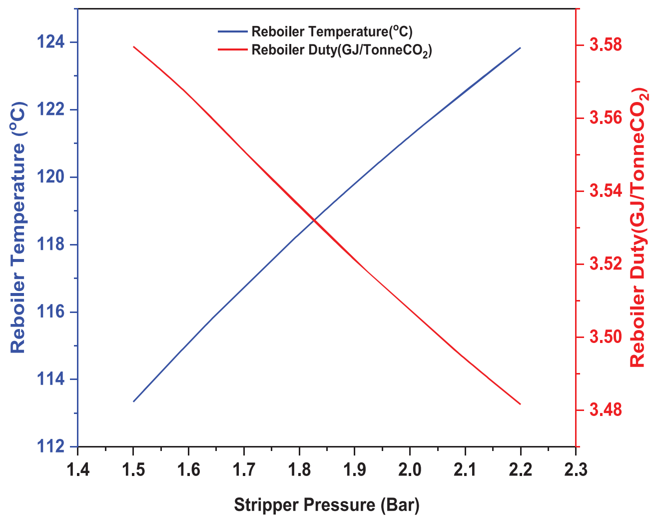

Figure 8. Effect of stripper pressure on reboiler temperature and reboiler duty.

($CAPEX_{Annual}$ USD/year) during the project life using Eqn (10). The project life (n) is 30 years and discount rate (i) is 8.5%.

$$CAPEX_{Annual} = CAPEX\left(\frac{i(1+i)^n}{(1+i)^n - 1}\right) \quad (10)$$

TAC is the sum of TOC and $CAPEX_{Annual}$ (Eqn (11)). The cost of CO_2 capture is calculated as TAC for capturing CO_2 divided by total CO_2 captured (Eqn (12)).

$$TAC = CAPEX_{Annual} + TOC \quad (11)$$

$$CO_2\ CaptureCost = \frac{TAC}{CO_2 Captured} \quad (12))$$

Result and discussion

For the analysis, the base case was first optimized to establish a baseline. Following this, the SOE and LVC were optimized using the optimized baseline parameters. Lastly, LVCSOE was optimized using optimized parameters from base case, SOE and LVC.

Optimization of base case

Effect of stripper pressure

The reboiler heat duty is significantly influenced by the stripper pressure. Figure 8 shows the effect of stripper pressure from 1.5 to 2.2 bar on reboiler duty and reboiler temperature. Reboiler duty decreases with increase in stripper operating pressure but reboiler temperature increases due to increase in pressure of the

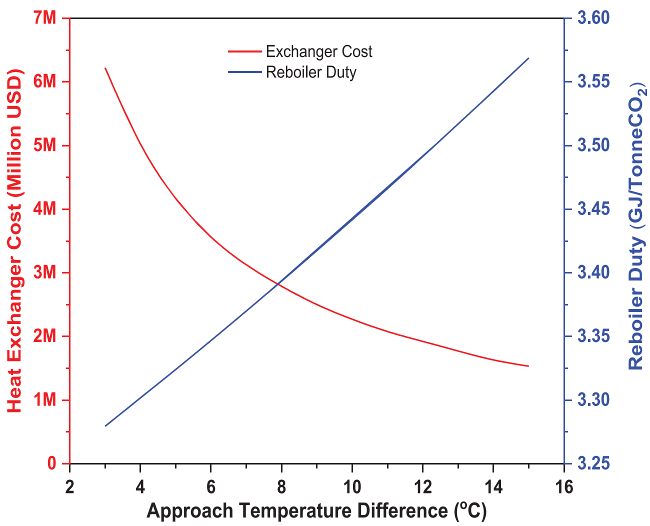

Figure 9. Effect of approach temperature on reboiler duty and HX cost.

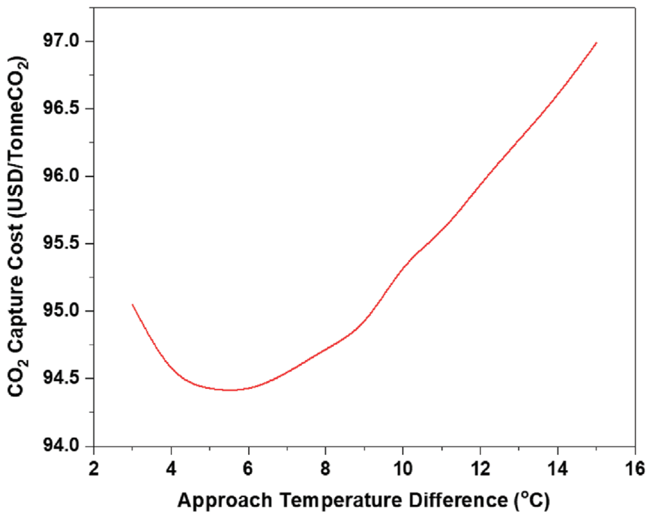

Figure 10. Optimization of LR-HX approach temperature.

boiling liquid. Although boiling temperature of CO_2-amine solution increases with increase in stripper pressure, but at high pressure, water evaporation decreases that reduces Q vaporization part of regeneration energy.[20] The overall impact is the reduction in reboiler duty at higher stripper pressure. Although the reboiler duty decreases by increasing stripper pressure, but it is not desired to operate the stripper at high pressure because higher reboiler temperature increases amine degradation and corrosiveness. Thermal degradation of MEA is minimum below 110 °C and very high above 130 °C.[8] Note that 1.9 bar stripper pressure, corresponding to 120 °C reboiler temperature, was selected as the optimum pressure because higher pressure can cause higher amine degradation, which increases the cost of make-up amine.

LR-HX approach temperature

In order to enhance the heat integration within the CO_2 capture process, the hot CO_2-Lean solvent exchanges heat with cold CO_2-Rich solvent in a countercurrent heat exchanger. As the temperature of rich solvent increases, the $Q_{sensible}$ portion of the heat duty decreases that consequently reduces the reboiler duty. The temperature approach can be reduced up to 5 °C, but lower approach temperature increases the area and cost of the heat exchanger.[46] The cold end temperature approach was varied from 3 to 15 °C to study its effect on the reboiler duty, heat exchanger cost, and cost of CO_2 capture. Figure 9 shows that reducing the approach temperature from 15 to 3 °C can reduce the reboiler duty by 10%. However, the exchanger purchase cost increases fourfold because of large heat transfer area requirement for achieving approach temperature as low as 3 °C. Figure 9 showing the effect of approach temperature on CO_2 capture cost reveals that reducing lean-rich exchanger approach temperature up to 5 °C is economically beneficial in terms of CO_2 capture cost. Although below 5 °C reboiler duty decreases, but the purchase cost of exchanger outweighs the benefits of reducing reboiler duty and thus the CO_2 capture cost increases (Fig. 10). Therefore, the optimum approach temperature is 5 °C at which the capture cost is minimum.

Optimization of SOE

Effect of split fraction

Split fraction dictates the efficiency of heat integration within the system. The target of this modification is to recover maximum heat from the product streams of the stripper column that reduces the reboiler duty. Split fraction (ratio of molar flow rate of $Rich_{To\ SOE}$ to $Rich_{Out}$) was varied from 0.05 to 0.425 to find the optimum split fraction at which the target of modification can be achieved. Figure 11 shows the effect of split fraction on reboiler duty. Rich split fraction of 0.36 shows minimum reboiler duty. The amount of heat a stream can recover from another stream is directly related to its flow rate. Split fraction to SOE higher than 0.36 can recover maximum heat from CO_2/H_2O vapor stream because of high flow rate to SOE but the heat recovery from hot lean stream in

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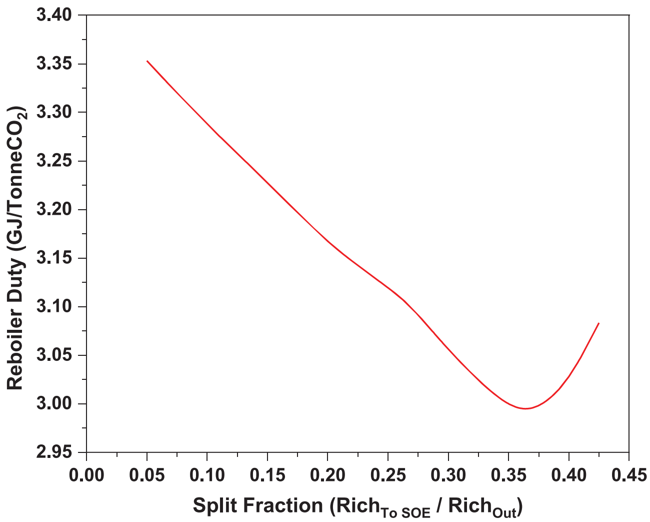

Figure 11. Effect of split fraction ($Rich_{To\ SOE}/Rich_{Out}$) on reboiler duty.

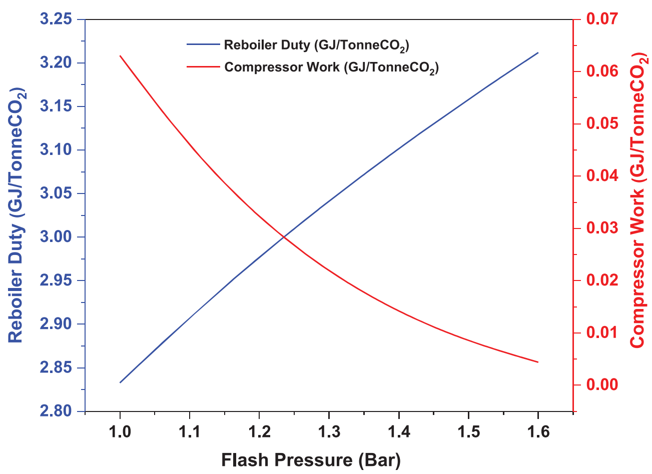

Figure 12. Effect of flash pressure on reboiler duty and compressor work.

LR-HX decreases due to lower flow rate to LR-HX. On the other hand, a split fraction lower than 0.36 to SOE reduces the flow rate to SOE, which cannot extract maximum heat from CO_2/H_2O vapor mixture leaving from stripper top. Note that 0.36 is the optimum split fraction to SOE at which overall maximum heat can be recovered from stripper column product streams.

Optimization of LVC

Effect of flash pressure

The pressure in flash drum affects both compressor work and regeneration energy. Figure 12 shows that reducing the pressure of flash drum decreases regeneration energy but the compressor duty increases.

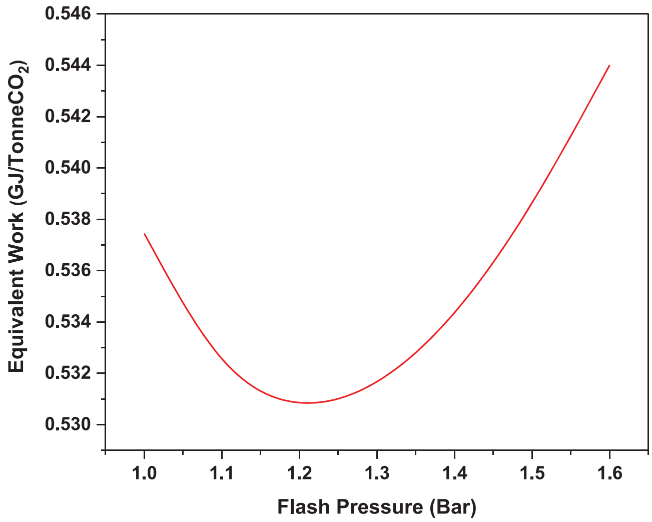

Figure 13. Optimum flash pressure.

By decreasing the pressure in flash drum, more amount of mechanical work is required to compress the gas to stripper pressure. This increases the compressor work as well as the temperature of the compressed gas. Injection of this high temperature gas to stripper reduces the reboiler duty by acting as a mass and energy stripping agent. Total equivalent work is a key parameter that can be used to understand the overall impact of flash pressure on process energy requirement, and is calculated using Eqn (13).[20]

$$W_{eq} = \sum_{i=1}^{n_{reboiler}} 0.756 \times Q_i \left(\frac{T_i + 10K - T_{sink}}{T_i + 10K} \right) + W_{comps} \qquad (13)$$

where Q_i is the reboiler duty (GJ/tonneCO_2), W_{comps} is the compressor work, T_i is the reboiler temperature, and T_{sink} is sink temperature taken as 313 K. Figure 13 shows that 1.2 bar pressure is the optimum flash pressure at which total equivalent work is minimum.

Location of the feed stage

The optimum location of CO_2-rich amine feed and compressed vapor feed was determined by varying the respective feed locations in the packed bed (0 m is the top and 11 m is the bottom of stripper column). CO_2-rich amine feed location was varied from 0.5 to 3.3 m packed height (Fig. 14). Regeneration energy increases significantly when the CO_2-rich amine is fed at the top of the stripper (Fig. 14). This is because the vapors in CO_2-rich feed streams strip CO_2 from the down coming liquid but when the rich solvent is fed to

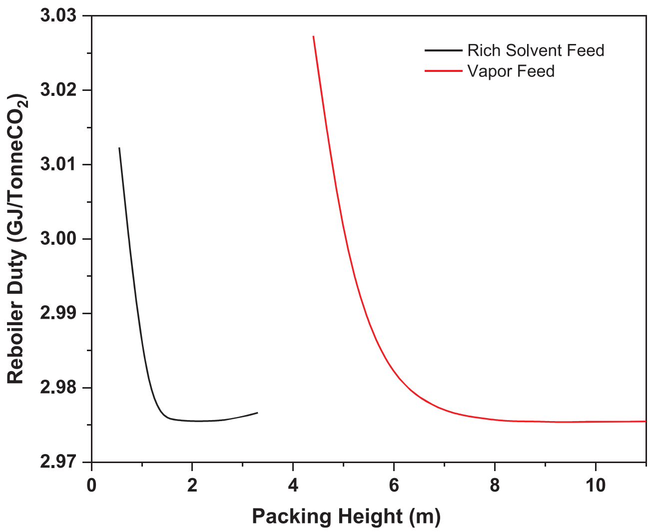

Figure 14. Optimization of feed location.

the top of stripper, this effect is abated. Compressed vapors feed location was varied from 4.4 m to the bottom of packed bed. Reboiler duty increases exponentially, when compressed vapors are fed above the 6.5 m height of packed bed. This can be attributed to the fact that high energy compressed vapors are no longer available to strip CO_2 from the liquid at the lower stages. The optimum location for CO_2-rich amine and compressed vapor feed is found to be at 2.2 and 8.3 m of packed bed, respectively.

Optimization of LVCSOE

LVCSOE is a combination of LVC and SOE modification. The optimized conditions for the aforementioned configurations were used for the optimization of LVCSOE configuration. Table 6 shows the optimum operating conditions for all four configurations.

Regeneration energy

Figure 15 shows the effect of process modification on the reboiler duty. Modified configurations show a reduction of up to 11% in regeneration energy, whereas the advance hybrid configuration, that is, LVCSOE, shows the highest reduction of 19% as compared to the base case. However, it should be noted that integration of LVCSOE increases the process complexity. Although LVC and LVCSOE configurations reduce regeneration energy, they have higher electric energy requirement due to addition of a compressor. The additional electric

Table 6. Optimum parameters of different configurations.

	Base case	SOE	LVC	LVCSOE
Absorber pressure (Bar)	1.5	1.5	1.5	1.5
CO_2 capture efficiency (molar %)	90	90	90	90
CO_2 capture rate (tonne/h)	190	190	190	190
Stripper pressure (Bar)	1.9	1.9	1.9	1.9
Flash pressure (bar)	-	-	1.2	1.2
Cold side ΔT (Rich-Lean HX) (°C)	16.3	14.5	9.4	12.4
Hot Side ΔT (Rich-Lean HX) (°C)	5.0	10.0	5.0	4.4
Rich solvent feed stage	3	6	4	7
Compressed vapors feed stage	-	-	15	15
Splitted stream feed stage	-	1	-	1
Split fraction	-	0.36	-	0.31
Reboiler temperature (°C)	120.6	120	120	120.2
Compressor work (GJ/tonneCO_2)	-	-	0.032	0.031
Reboiler duty (GJ/TonneCO_2)	3.36	2.99	2.98	2.72

energy required for vapor compression in LVC & LVCSOE is 0.03 GJ/tonneCO_2.

Economic analysis

Capital expenditure

The CAPEX required for base case and modified configurations is compared in Fig. 16. Approximately 207.3 million USD are required to install CO_2 capture plant for 300 MW coal-based power plant. A minor modification like SOE can reduce this capital expenditure by nearly 2.1 million USD. This reduction in capital expenditure is due to lower heat exchangers cost, which is directly related to heat transfer area requirement. Heat transfer area required in an exchanger can be calculated using Eqn (14).

$$A = \frac{mC_p\Delta T}{U\Delta T_{LM}} \tag{14}$$

where A is the area of the exchanger (m^2), m is the mass flow rate (kg/s), ΔT is temperature difference at inlet and out of stream (K), C_p is the heat capacity of solution (J/kg·K), U is the overall heat transfer coefficient (W/m^{2}K), and ΔT_{LM} is log mean

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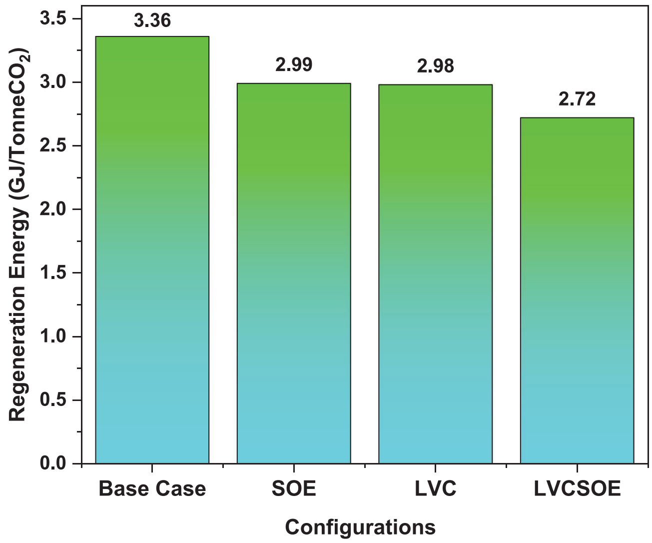

Figure 15. Regeneration energy comparison of different configurations.

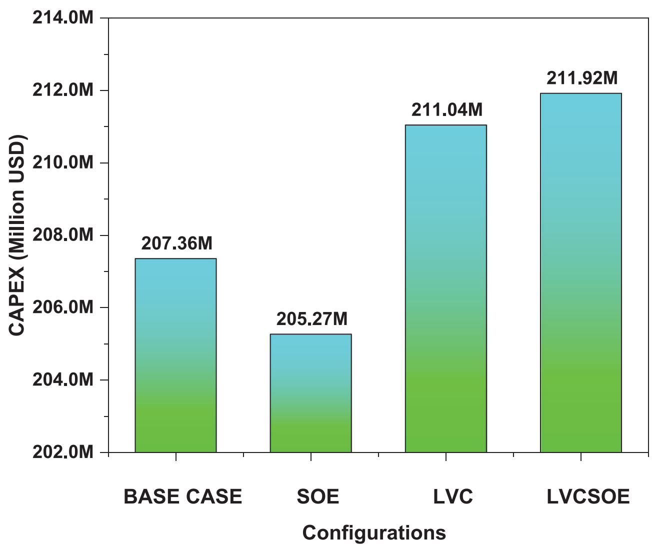

Figure 16. Capital expenditure.

temperature difference across the exchanger (K). Mass flow rate directly influence the required heat transfer area. By splitting CO_2-rich amine, stream mass flow rate to LR-HX decreases that reduces the heat transfer area and cost of LR-HX. SOE reduces the inlet temperature of condenser, which reduces ΔT and the area required. The reduction in the cost of Lean-Rich exchanger and condenser is higher than the cost of SOE. The overall effect is the substantial reduction in heat exchangers cost and CAPEX requirement. On the other hand, modification like LVC can increase the CAPEX requirement by more than 4.1 million USD

Table 7. Cost contribution of process equipment in purchased equipment cost.

	Base case	SOE	LVC	LVCSOE
Compressor (%)	24.2	24.4	26.7	26.6
Separator (%)	0.7	0.7	1.2	1.2
Heat exchanger (%)	17	16.2	15	15.2
Pump (%)	0.3	0.3	0.3	0.3
Stripper (%)	8.7	8.8	8.5	8.5
Absorber (%)	49.1	49.6	48.3	48.1
Total equipment cost (million USD)	45.22	44.76	46.00	46.2

due to the addition of equipment, that is, compressor, flash column, and flash valve. CAPEX calculation of LVCSOE shows the same trend as in LVC modification due to the additional equipment.

Table 7 also shows the contribution of purchase cost of process equipment in different configurations. The absorber has the largest contribution in TEC followed by compressors in all configurations. The columns (absorber/stripper) account for more than 50% of TEC. This is consistent with the published economic analysis of CO_2 capture process at Huaneng Beijing Power Plant.[47] The equipment purchase cost for CO_2 compression unit to deliver CO_2 at 150 bar is almost 25% of the TEC. The number of compressors required and their cost can be reduced by varying the delivery pressure depending on the end use of captured CO_2.

Total operating cost

Figure 17 shows the TOC with the breakdown of fixed and VOC for all configurations in terms of USD/tonne of CO_2 captured. The trend of TOC is as follow: base case > LVC > SOE > LVCSOE. The contribution of FOC in TOC ranges between 31 and 35% while the rest of 61–65% is VOC (Fig. 17). FOC calculation shows very little variation for different configurations. So, the variation in TOC is due to variation in main contributor, that is, VOC.

SOC, LVC, and LVCSOE configurations show a reduction of 6%, 6.6%, and 11.5%, respectively, in VOC relative to the base case. Table 8 shows the cost of utilities contributing to VOC. High VOC is mainly due to high regeneration energy cost, which is 68–73% of VOC. Base case with highest reboiler duty has the highest regeneration energy cost, that is, 26.9 USD/tonneCO_2. LVCSOE shows a reduction of 17.8%

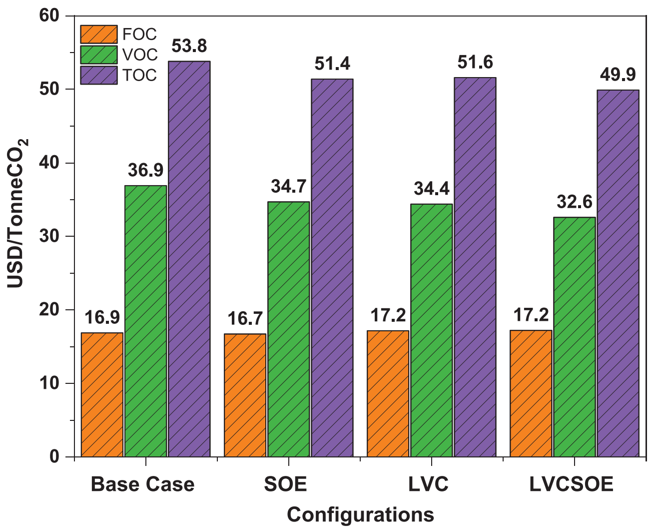

Figure 17. TOC, FOC, and VOC comparison in different configurations.

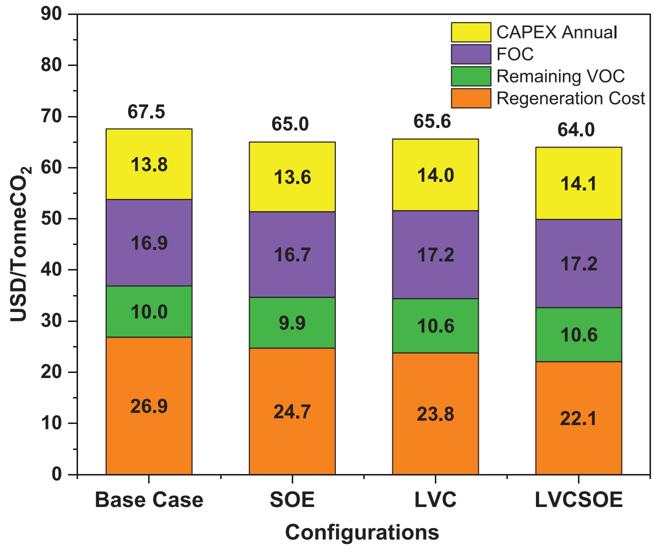

Figure 18. CO_2 capture cost breakdown in different configurations.

Table 8. Variable operating cost breakdown in utilities and chemicals.

	Base case	SOE	LVC	LVCSOE
MEA makeup ($USD/tonneCO_2$)	0.81	0.81	0.81	0.81
Water makeup ($USD/tonneCO_2$)	0.55	0.55	0.55	0.55
Cooling water ($USD/tonneCO_2$)	0.54	0.49	0.47	0.42
Electric energy ($USD/tonneCO_2$)	8.09	8.09	8.77	8.77
Regeneration energy ($USD/tonneCO_2$)	26.9	24.7	23.8	22.1

as compared to base case regeneration energy cost, followed by LVC, 11.4% reduction and SOE, 8% reduction. Due to better heat integration within the system, cooling water requirement also decreases in modified configurations. LVCSOE requires 22% less cooling water as compared to base case. Electric energy cost is almost similar in base case and SOE but with the addition of compressor in LVC and LVCSOE, it increases by 8.4% compared to base case. The lean vapor compressor electric energy expenditure in LVC and LVCSOE is less than 0.7 $USD/tonneCO_2$. Electric energy cost is mainly linked with compression of CO_2, which can be reduced by introducing more efficient CO_2 compression techniques.[25]

A comparison of utilities cost in LVCSOE configuration with base case configuration shows a reduction of 18% (4.79 $USD/tonneCO_2$) in regeneration energy cost and 22% (0.12 $USD/tonneCO_2$) in cooling water utility cost with an increase of 8% (0.7 $USD/tonneCO_2$) in electrical energy cost. On the whole, it can be inferred that reduction in regeneration energy can bring more impact on VOC than any other variable, that is, 18% reduction in regeneration energy reduces cost by 4.79 $USD/tonneCO_2$.

Cost of CO_2 capture

The most important index of this study is CO_2 capture cost that includes TOC and annualized CAPEX. The trend of CO_2 capture cost is as follows: LVCSOE < SOE < LVC < base case. LVC, SOE, and LVCSOE show reduction of 2.9%, 3.7%, and 5.3%, respectively, in CO_2 capture cost relative to base case. LVC, SOE, and LVCSOE save 2.9, 3.7, and 5.4 million USD per annum, respectively. $CAPEX_{Annual}$, which is only 20% of the total capture cost, shows little variation due to comparable CAPEX for different configurations. Operating cost, which is approximately 80% of capture cost, shows much more variation than $CAPEX_{Annual}$. Process modifications reducing TOC can bring more impact on reducing CO_2 capture cost because of higher contribution in CO_2 capture cost.

Figure 18 shows the breakdown of total cost of CO_2 capture as annualized CAPEX, FOC, and VOC. Regeneration energy cost is a major part of CO_2 capture cost so VOC is shown as regeneration energy

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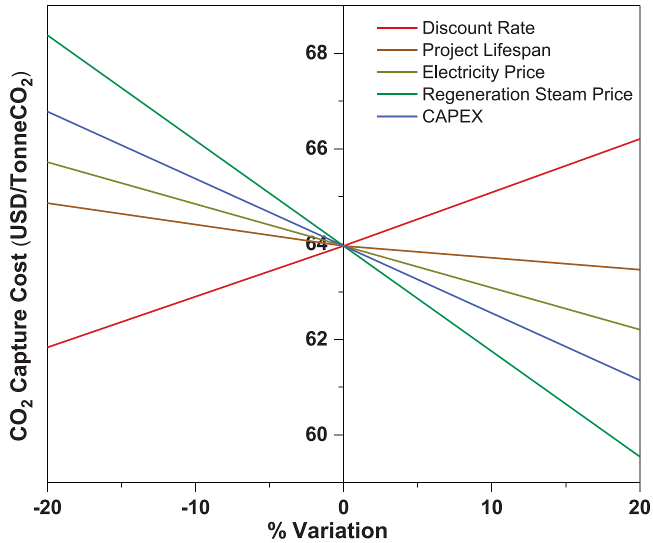

Figure 19. Sensitivity analysis of economic variables on CO_2 capture cost.

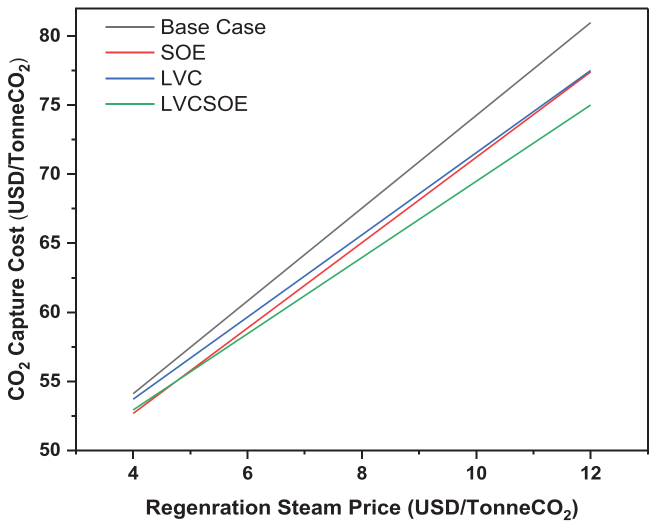

Figure 20. Effect of regeneration steam price on CO_2 capture cost.

cost and remaining VOC. In all the configurations, regeneration energy cost is more than one-third of the total capture cost. In base case regeneration, energy cost contributes 40% of CO_2 capture cost while in LCVSOE configuration it is 35%, thus LVCSOE has the lowest CO_2 capture cost. The lower regeneration energy cost in modified configuration has outweighed the extra CAPEX required for modification. LVCSOE configuration reduces the cost to 64 USD/tonneCO_2 from 67.5 USD/tonneCO_2, a net saving of 5.4 million USD per annum.

Sensitivity analysis

Sensitivity analysis was carried out to understand the effect of different variables on CO_2 capture cost. The economic variables such as regeneration steam price, electricity energy price, project lifespan, capital expenditure, and discount rate were varied by ±20% to study their effect on CO_2 capture cost. Since LVCSOE has shown the least capture cost, so sensitivity analysis was performed on LVCSOE configuration.
Regeneration steam price variation shows the highest impact on capture cost (Fig. 19). A variation of ±20% in regeneration steam price changes capture cost in range of 59.5–68.4 USD/tonneCO_2. After regeneration steam price, variation in CAPEX shows more impact on CO_2 capture cost then other economic variables. Electric energy price and discount rate have minimal effect on capture cost. Variation in CAPEX, electric energy cost, and discount rate by ±20% cause a change in the range of 61.1–66.8, 62.2–65.7, 66.2–61.8 USD/tonneCO_2 respectively, in capture cost. An addition of 6 years in project lifespan reduces cost by 0.5 USD/tonneCO_2, while reducing project life by 6 years increase cost by 0.9 USD/tonneCO_2.

The regeneration steam price is the most sensitive economic variable so its impact was further studied and compared for different configurations. As the steam price decreases, regeneration steam cost contribution decreases to the level that regeneration energy has a very minimal contribution to CO_2 capture cost. Figure 20 shows that at low regeneration steam price, the main contribution to capture cost is from $CAPEX_{Annual}$ and FOC, therefore, modified configurations do not show benefit of reducing regeneration energy and all configurations have almost same CO_2 capture cost. So, at lower regeneration steam price, $CAPEX_{Annual}$ should be kept low as it will have more effect on capture cost than regeneration steam cost. It is more beneficial to implement configuration with minimum regeneration energy at high fuel cost as fuel cost is directly related to the price of regeneration steam.

Conclusion

In this work, technical and economic analysis of CO_2 capture process for coal fired power plant was performed for clean power generation. Conventional process and three stripper modifications, including a new process configuration, were studied. Aspen plus rigorous rate-based model was used for simulation and optimization of the process and Aspen Process Economic Analyzer was used for determining the purchase equipment cost. Economic analysis was performed to calculate the cost of CO_2 capture. Furthermore, the configurations were optimized by studying various design and operational parameters including stripper pressure, LR-HX approach temperature, flash pressure, split fraction, and feed stage location. The results show that

- all modified and optimized configurations are technically and economically feasible,
- SOE modification reduces the capture cost as well as CAPEX requirement,
- LVC modification reduces the capture cost but increases CAPEX requirement,
- hybrid configuration (LVCSOE) reduces the regeneration energy requirement by 18%,
- LVCSOE with comparable CAPEX to LVC shows a reduction of 3.6 USD/tonneCO_2 in CO_2 capture cost as compared base case; a net savings of 5.4 million USD per annum.

Sensitivity analysis was also carried out to understand the impact of variation in economic variable. The CO_2 capture cost is most affected by regeneration steam price, followed by CAPEX, discount rate, electric energy cost, and plant lifespan, respectively. Regeneration steam cost contributes 35–40% in CO_2 capture cost. Since regeneration steam price is the most sensitive variable, sensitivity analysis of regeneration steam price for different configurations was further carried out, which reveals that it is beneficial to implement configuration with lower regeneration energy at higher steam prices while simple configurations with lower CAPEX, which are easy to operate, will be more technically and economical viable at lower steam price.

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