Benchmarking of a Novel Solid Sorbent CO₂ Capture Process for NGCC Power Generation

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Abstract:

The successful deployment of any technology depends on many important factors. One of the critical ones that affect the success of post combustion CO₂ capture technologies is the ability to reduce the increase in cost of electricity associated with retrofitting the capture unit to electricity generating power plants. High capture costs associated with conventional MEA solvent processes has opened an opportunity for the emergence of many competing technologies aimed at reducing the cost of capture. This paper presents the technical and economic benchmarking of a polyethyleneimine (PEI)-based solid sorbent CO₂ capture process as compared to the conventional MEA solvent process, with both technologies applied to a natural gas combined cycle (NGCC) power plant.

The results of this evaluation indicate that the PEI-based sorbent process requires approximately 25% less regeneration steam than the conventional MEA process primarily due to avoiding the energy demand required for heat of vaporization of water in the MEA process. The PEI solid sorbent based process results in approximately 6.3 power plant efficiency point reduction as opposed to an 8 efficiency points for the MEA based process. The design of the fluidized-bed absorption and regeneration reactors in the novel solid sorbent process result in significant

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reductions in capital costs (21% reduction compared to MEA). The economic evaluation shows that the PEI-based solid sorbent process offers attractive reductions in the cost of electricity and the CO₂ capture cost. The estimated cost of electricity, after adding the CO₂ capture process, is found to be 85 and 81.7 \$/kWh for the MEA-based and PEI based process, respectively. Moreover, the cost of CO₂ avoided for both process is estimated to be 69.8 and 62 \$/ton CO₂ for the MEA and PEI process, respectively.

KEYWORDS: Post-combustion capture, solid sorbent, economic evaluation

1. Introduction

Significant efforts are underway worldwide to mitigate the climate change and its impacts. With the world's growing dependence on fossil fuel energy sources for power generation, CO₂ emissions are expected to increase and, along with other greenhouse gases, continue to contribute heavily to worldwide climate change [1] .

The threat of global climate change has resulted in an increased interest in and exploration of gas treatment technologies, such as CO₂ capture technologies. These technologies represent a vital component in a comprehensive strategy to mitigate greenhouse gas emissions and the potential impacts of global climate change [2, 3]. More specifically, carbon capture and storage (CCS) can be an effective method to reduce CO₂ emissions from stationary, point sources such as power generation plants. Of three major categories for CO₂ capture technologies (post-combustion, pre-combustion, and oxy-combustion), post-combustion CO₂ capture (PCC) is the most widespread and adopted approach due to its roots in conventional gas purification and desulfurization processes [4].

Additionally, post-combustion CO₂ capture is an attractive approach as it can be retrofitted to existing power plants with little to no modifications required in the existing plant [5].

Comparative assessment studies have shown that chemical absorption process is currently the preferred option for post-combustion CO₂ capture and the monoethanolamine (MEA) solvent-based process is considered the state of the art post combustion capture technology [19]. Chemical absorption offers high capture efficiency and high selectivity at low partial pressures. However, the major drawback to conventional MEA solvent technologies are their high CO₂ capture costs resulting from the significant solvent regeneration energy demand, high capital costs of the absorption and regeneration equipment, and high operating costs. To overcome the challenges and drawbacks of the MEA-based capture process; researchers and technology developers have been working on the development of novel chemical solvent systems [[20]-[24]]. These development activities focus on the reduction of the energy requirement and the total CO₂ capture cost. Many of these developed technologies have demonstrated improved performance over the MEA-based process. For these advance solvent systems, an improved performance and capture cost is claimed. However, there is still a major need for breakthrough technologies those could result in major CO₂ capture cost reduction.

As with many countries worldwide, the United Arab Emirates (UAE) has a strong dependence on fossil fuels to meet the country's energy demands. In particular, the UAE is dependent on natural gas for power generation and this demand is increasing due to an increasing population and energy use [6]. Given this situation and the UAE's commitment to supporting clean and sustainable energy use, the exploration of post-combustion CO₂ capture processes is of significant interest to the UAE and will help curb the emissions generated from the country's power plants. A major focus of UAE's exploration into carbon capture options is the improvement or elimination of the major drawbacks to conventional MEA-based carbon capture.

The following options, or combinations of these options, are being explored to mitigate the drawbacks of PCC capture:

- Alternative solvents (or mixtures of solvents) which lower their regeneration energy requirement or improve properties such as increasing their CO₂ absorption capacities [7-9]
- Process design and process flow sheet modifications which result in energy penalty reductions for conventional solvent-based systems [10-13]
- Alternative separation processes (solid sorbents, membranes, etc.) with greater potential for reduced energy requirement while maintaining required CO₂ capture capacities.

Related to the third point above, in order to assess the technical and economic advantages of novel processes, a benchmarking study must evaluate the new technology option against the conventional technology on the same basis. It is the aim of the presented work to assess the technical and economic advantages (and disadvantages) of a novel solid sorbent CO₂ capture process, compared to conventional MEA scrubbing, when both technologies are applied to a commercial-scale NGCC power plant.

The solid sorbent of interest is a “molecular basket sorbent” (polyethylenimine-based (PEI) solid sorbent) developed by Dr. Chunshan Song’s research team at Pennsylvania State University (PSU) [14-17]. The PEI-based sorbent is essentially a silica support material impregnated with PEI and is utilized in a continuous, fluidized-bed process developed by the Research Triangle Institute International (RTI).

To date, the RTI process has been developed for coal-fired power plant applications. For the work presented herein, RTI’s process has been adapted to the flue gas emitted from a NGCC plant. A technical and economic assessment for both RTI’s process and a conventional MEA

CO₂ capture process has been performed and the results are compared in order to assess the commercial attractiveness and potential for deployment of RTI's novel CO₂ capture technology.

2. Assessment Methodology

The methodology for developing the economic benchmarking is simplified in Figure 1. The first step in the benchmarking study is to establishing a power plant reference case upon which the work is based. The reference NGCC power plant selected for this work is based on the NGCC case specified in the U.S. Department of Energy's (DOE's) "cost and performance baseline for fossil energy plants" report first developed in 2007 [18]. A schematic of this 555MW reference power plant is shown in Figure 2 with the operational performance showing in Table 1.

Figure 1

Table 1

Figure 2

The NGCC reference plant operates at a LHV efficiency of 55.7% as follows: air and natural gas enter a dry low NO_x burner (LNB) where the natural gas burns producing a combustion gas at 629°C. The gas is then sent to a heat recovery steam generator (HRSG), generating steam which is subsequently used in a steam turbine for power generation. The flue gas leaving the HRSG (shown in Figure 2) is then sent to a hypothetical CO₂ capture unit. More details and specifications of the NGCC power plant and its specifications can be found in the U.S. DOE report [18].

In the work presented, the flue gas specifications from the DOE report were used in Aspen Plus to model two CO₂ capture cases: an MEA process case and a PEI-based solid sorbent process case. Both CO₂ capture systems are designed to remove 90% of the CO₂ present in the NGCC

1 plant flue gas. The process flow diagram and basic assumptions for the MEA process are shown
2 in Figure 3.

3 **Figure 3**

4 The Aspen Plus simulations were built in order to output useful process information for both
5 CO₂ capture cases: 1) energy balances, 2) mass balances, and 3) stream tables for each case, all
6 of which allow for a technical comparison of the power, steam, and cooling water demands for
7 both technologies. There are various methodologies and approaches utilized by organizations to
8 report the costs associated with carbon capture and storage systems [29]. This project is funded
9 by the US DOE NETL and as such, it is a project requirement to utilize the proposed
10 methodology of the US DOE to conduct the economic evaluation. This section aims to provide
11 an overview of the economic assessment methodology utilized in the work. The following is a
12 step-by-step methodology for performing the economic evaluation and benchmarking study:

- 13 1. The simulation of each CO₂ capture process is carried out in Aspen Plus to obtain the
14 necessary information to design and size equipment and other system components
15 (including absorber and regenerator vessels).
- 16 2. Detailed sizing of the absorption and regeneration vessels is carried out using the Aspen
17 Plus simulation data in reactor sizing models described in section 3.3. The remaining
18 process equipment (for each process technology) is sized based on literature correlations.
- 19 3. The equipment costs (and costs of both capture system components) are determined using
20 Aspen Process Economic Analyzer which leads to the determination of the bare-erected
21 cost (BEC) of the equipment. The capital cost nomenclature and calculation method is
22 summarized in Table 2.

- 1 4. The detailed techno-economic evaluation boundary conditions were limited to the
2 compression and dehydration system. However, the costs associated with CO₂
3 transportation, storage and monitoring were included as a fixed value in the final cost of
4 electricity [18].
- 5 5. The MEA-based process is considered mature and commercial technology; therefore, no
6 process contingency was assumed. However, 20% of the total equipment bare-erected
7 cost is added as process contingency costs for the solid sorbent-based technology. In
8 addition, 25% of the BEC is assumed as project contingency for both capture processes.
9 Owner costs represent preproduction costs, working capital costs and financial costs
10 required to fund this capital. Preproduction or startup costs include 6 months operating
11 labor, 1 month maintenance material at full capacity, 1 month non-fuel consumables at
12 full capacity, 25% of one month's fuel cost at full capacity, and 2% of Total Plant Cost.
13 Working capital costs include inventory capital and the cost of land. For the inventory
14 capital, the charges are 0.5% of Total Plant Cost for spare parts, a 60 day supply of fuel
15 and non-fuel consumables at full capacity, and land for the NGCC case is 1000 acres
16 charged at \$3000/acre. Totaling these charges gives us the Total Owner's cost which is
17 used in the calculation of cost of electricity (COE) [18] .
- 18 6. Other cost components such as owner costs are added to the Total Capture Plant Cost. In
19 addition, The fixed operating costs for both processes are determined using the guidance
20 described in Case 14 of the DOE report [18].
- 21 7. The variable operating costs for both the MEA and solid sorbent processes are calculated
22 separately using the simulation mass and energy balances. These costs are estimated
23 using the detailed guideline of the DOE report. Performance and process requirement

parameters for the solid sorbent process were assumed and a summary of these assumptions is included in the process description and in Table 3.

8. CAPEX and OPEX information, developed in the steps above, are used to calculate specific economic metrics (e.g. cost of CO₂ captured) which are used in benchmarking the novel solid sorbent CO₂ capture process.

9. The cost of electricity is calculated using the following equation:

$$\text{Cost of Electricity} = \frac{(CCF)(TOC) + OC_{FIX} + (CF)(OC_{VAR})}{(CF)(MWH)} \quad (1)$$

10. The cost of CO₂ capture is calculated using the following equation:

$$\begin{aligned} \text{Cost of CO}_2 \text{ Capture} = & \text{[Cost of all materials and utilities consumed in the CO}_2 \\ & \text{system]} + \text{[cost of treating or disposing of any effluent} \quad (2) \\ & \text{streams from the system, including CO}_2 \text{ TS\&M]} + \\ & \text{[cost of maintenance and materials]} + \text{[fixed O\&M} \\ & \text{costs]} + \text{[capital charge for CO}_2 \text{ capture system]} \end{aligned}$$

Table 2

Table 3

3. Process Simulations and Equipment Sizing

3.1 MEA Process Simulation and Assumptions

The MEA capture process evaluated in the presented work is based on the conventional PCC process described in Case 14 of the DOE report [18]. In this configuration flue gas from the NGCC power plant enters a direct contact cooling water tower where the temperature is reduced to a favored absorption reaction temperature (40°C). The flue gas stream pressure is then boosted prior to entering the absorber in order to overcome the pressure drop in the absorber column. For this simulation, the pressure drop was assumed to be 4.8kPa per stage based on the work of Abu-

Zahra et.al.[2] for a three-stage equilibrium-based absorption column. The CO₂-lean flue gas exits the top of the absorber and is scrubbed with water to remove any volatilized solvent before being vented through the existing stack at the NGCC power plant.

The CO₂-rich solvent exits the bottom of the absorber and is pumped through a cross-heat exchanger before entering the top of the regenerator column. The heat exchanger and regenerator act to increase the solvent temperature to a point where the reverse reaction (i.e. regeneration) is favored (120°C). The CO₂ that evolves from the rich solvent exits the top of the regenerator and is sent through a dehydration and compression train at pressure of 15.27 MPa. To maintain the CO₂ stream water content below the pipeline specifications of <150 ppmv; a triethylene glycol (TEG) dehydration unit was included in this evaluation for both the MEA and the solid sorbent processes [18].

The process described above was simulated using a complete closed flowsheet with design specifications that maintain the system water balance. The following are some assumptions made in completing the MEA process simulation:

- Solvent concentration of 30 wt% MEA.
- 90% CO₂ capture ratio.
- Lean solvent loading of 0.25 mol CO₂/mol MEA. This lean loading value is an optimization variable that is controlled to minimize the regeneration energy requirement based on the degree of regeneration of the solvent. The value assumed in this work was found by Abu-Zahra et.al. [2] to lead to the lowest regeneration energy requirement.
- Cooling water supply conditions are: 15.6°C upon entering and 26.7°C upon exiting the heat exchanger units in the process.

- Solvent losses are avoided completely (i.e. assumed to be zero) due to the inclusion of the scrubbing/reclaimer unit.
- Low pressure steam is extracted from the low pressure steam turbine and supplied to the reboiler at 250 kPa (2.5 bar) and 268.73°C.

3.2 PEI-based Solid Sorbent Process Simulation and Assumptions

The solid sorbent CO₂ capture process evaluated in this work is based on PEI amine, which was developed initially by PSU. PEI contains primary, secondary, and tertiary amine sites which capture the CO₂ primarily through carbamate and bicarbonate reaction pathways. The PEI CO₂ capture process was tested and evaluated experimentally at RTI facilities. The aim of this experimental evaluation was to define the process operational window and requirement. The following major operational conditions and specifications were obtained experimentally by RTI and used as the basis for this solid sorbent evaluation [[25]-[28]]:

- Absorption operating temperatures are between 50 to 90°C,
- Absorption reaction enthalpy is 66 kJ/mol CO₂ (1.5 MJ/kg CO₂).
- Sorbent regeneration temperature is above 110°C
- Desorption reaction enthalpy 78 kJ/mol CO₂ (1.77 MJ/kg CO₂). This desorption enthalpy forms the minimum theoretical regeneration energy. However, the actual (total) regeneration energy would include, in addition to the reaction enthalpy, the sensible heat.
- The solid sorbent specific heat is 1.5 kJ/kg.K, which is used to calculate the sensible heat needed in the regeneration column.
- CO₂ absorption capacity is 8wt% (0.08 kg CO₂/kg sorbent) based on preliminary experimental results with a simulated NGCC flue gas conducted at RTI [20].

- Sorbent degree of regeneration is 95%, which will result in sorbent lean loading capacity of 0.42 wt% (0.0042 kg CO₂/kg sorbent)
- The sorbent absorbs 10% of the water in the flue gas.
- Sorbent particle density is 1200 kg/m³.
- The rate of attrition in both the absorber and regenerator is 25x10⁻⁶ kg of sorbent/kg of sorbent circulated. Fresh sorbent is continuously introduced to the process to overcome the attrition loss.

A schematic of RTI's process (installed within a NGCC power plant) is shown in Figure 4. The flue gas from the plant is sent to a direct contact cooler through a gas blower. The gas blower boosts the gas pressure by 0.1724 bar to overcome pressure drop in the following system components: 1) direct contact cooling tower, 2) absorber (gas distributor and absorber column) and 3) baghouse filter. The pressure drop across each of these units is based on some values pulled from the DOE report [18] and from process engineering calculations provided by RTI. Upon leaving the top of the cooling tower the flue gas is cooled to a temperature of 40°C. Circulating water in the cooling tower cools the flue gas, causing condensation of water to mix with the circulating water. This water is sent to a knock out drum to remove any excess water (keeping a constant circulation flow rate) and the water is recycled back to the cooling tower. The flue gas coming from the direct contact cooler enters the bottom of the absorber via a gas distributor at 40°C and a pressure of 0.1724bar. The gas flows upwards and is mixed with the sorbent which selectively removes CO₂ and releases heat due to the exothermic reaction. The flue gas and sorbent act as minor heat sinks and absorb some of the exothermic heat. In the simulation the flue gas and sorbent are assumed to reach a temperature limit of 70°C. Tempered cooling water (TCW) is used to remove the remaining exothermic heat and maintain the absorber

1 at 70°C. TCW flows through internal heat exchanger tubes installed within the absorber and
2 removes the sorbent bed heat without direct contact with the sorbent. Control of the cooling
3 water temperature is critical as it must be high enough to avoid any condensation of moisture in
4 the flue gas (thus avoiding fouling of heat transfer internals and solids agglomeration) and must
5 be cold enough for sufficient heat removal. As a result, the TCW enters the absorber at 46.1°C
6 and the flow rate is controlled such that a 5.6°C temperature rise is allowed. The TWC leaves the
7 absorber and is cooled back down to the desired supply temperature by mixing it with a stream
8 of circulating cooling water.

9 **Figure 4**

10 During operation of the process, The CO₂-lean flue gas exits the top of the absorber. The flue gas
11 may contain some entrained solids so a baghouse filter is used to remove the particulates present
12 in the gas. The treated flue gas is then vented through the existing plant's stack at a pressure
13 slightly higher than atmospheric.

14 CO₂ rich sorbent stream leaves the bottom of the absorber and is mechanically conveyed to the
15 top of the regenerator. The mechanical mechanism assumed to be used in the simulation is
16 continuous bucket elevators. These elevators are operated in an enclosure to prevent contact of
17 the sorbent with the atmosphere, which could result in irreversible contamination of the sorbent.
18 Upon exiting the mechanical conveyors (and entering the regenerator) the CO₂-rich sorbent is
19 heated to 110°C driving the reverse reaction for CO₂ release from the PEI reagent. Steam from
20 the power plant is used to supply the heat needed for regeneration. Steam is used both in the
21 fluidizing gas to maintain fluidization of the sorbent throughout the regenerator and it is used as
22 a stripping gas, providing heat indirectly to the sorbent bed by condensing in heat transfer
23 internals similar to those used in the absorber. Following regeneration, the CO₂-lean sorbent

exits the bottom of the regenerator and is returned to the absorber, closing the absorption/desorption cycle. The CO₂ gas which has been released from the sorbent exits the top of the regenerator and passes through a baghouse filter for removal of any entrained solids. Finally, this “regeneration off gas” is sent to a dehydration and compression system similar to the one used in the conventional MEA process.

Similar to the method of sorbent transfer from absorber to regenerator, the CO₂-lean sorbent leaving the regenerator is mechanically conveyed to the top of the absorber. Prior to entering the absorber the sorbent is passed through a sorbent cooling system where the fluidized solids are allowed to free fall over heat transfer tubes containing TWC to reduce the sorbent temperature to 70°C.

In addition to the earlier discussed solid sorbent performance specification, the following are some additional assumptions for the solid sorbent process simulation:

- The process is designed to capture 90% of the CO₂ in the flue gas.
- The sorbent absorbs 10% of the water in the flue gas.
- Given limitations to the underlying chemical property data in Aspen, a custom model component for the sorbent’s physical and chemical properties was not created. The sorbent is assumed to be composed of SiO₂ and Na₂CO₃ to mimic the effect of the PEI.
- Gas flows in both the absorber and regenerator have a superficial gas velocity of 0.6 m/s.
- The sorbent regeneration temperature is fixed at 110 °C. This temperature is very close to the regeneration temperature of the MEA conventional process. This would result in similar steam quality and conditions required to be supplied in both cases.

3.3 Absorber and Regenerator Sizing Models

The design model used to size the MEA absorber and desorber is based on the design model detailed in the work by Abu-Zahra et al. [2] and is depicted in Figure 5. The design model takes as inputs data extracted from the Aspen process simulation and outputs design specifications such as column diameter, packing height, and overall height of the absorber.

Figure 5

The solid sorbent process reactor vessels are envisioned to be square cross-section concrete columns with horizontal carbon steel tubes used for heat transfer. The dimensions of these vessels are based on two main factors: 1) the superficial gas velocity (flue gas in the absorber, stripping steam/recycle CO₂ in the regenerator) and 2) the heat transfer area required for cooling/heating the respective vessel. The solid sorbent process absorber / regenerator design model block diagram is shown in Figure 6 and the following steps are followed in designing the solid sorbent absorber and regenerator vessels:

1. The edge length of the square cross-section is determined by using the desired superficial gas velocity (v), the gas volumetric flow rate ($\dot{V}$) and an assumption of the number of vessels (n) in Equation (3).

$$E = \sqrt{\frac{\dot{V}}{v \times n \times 3600}} \quad (3)$$

2. The heights of the absorber and regenerator are determined by adding the heat transfer height (HTH) requirement to the transport disengaging height (TDH) of the vessel (i.e. the height required for particles with terminal velocities less than the superficial gas velocities to fall back into the bed). The HTH requirement is determined based on the required heat transfer area which is calculated using the traditional LMTD method and the selected piping OD/ID of the tubes. The TDH of each vessel is assumed to be a direct

function of the superficial gas velocity. In adding the TDH and HTH together, an additional 25% in height is added to account for other vessel internals such as gas distribution arrays. The total heights of the vessels are then simply the addition of the TDH, HTH, and the extra height added for other internals.

Figure 6

The detailed absorber and regenerator design for both the MEA and solid-based processes were conducted based on the process and sorbent performance and using typical engineering design guidelines and procedures. Table 4 shows an overall summary of the major absorber and desorber design assumptions, performance and material of construction for both MEA-based and solid-based processes.

Table 4

4. Results and Discussion

4.1 Technical Performance Comparison of the MEA and Solid Sorbent Processes

In order to benchmark the technical performance of the novel solid sorbent process versus the MEA process, a set of comparison criteria needs to be established. Table 5 lists a set of four critical technical performance criteria on which the two processes are compared: 1) regeneration steam requirement, 2) energy demand, 3) electricity demand and 4) cooling water demand. The calculated values for these variables are exhibited for both CO₂ capture processes. The technical performance results of the MEA process are in excellent agreement with results reported in literature for similar case studies [1-3], thus it is confirmed that this MEA process simulation serves as a viable baseline for comparison to RTI's novel process.

Table 5

The technical performance of the PEI-based solid sorbent process exhibits an improvement over the MEA process in all areas except one (electricity demand). The regeneration steam requirement of the solid sorbent process is lower since the MEA process has a higher sensible heat and has an additional heat requirement to meet the solvent heat of vaporization. This regeneration heat requirement is closely tied with the total energy demand of the process, which is lower in the solid sorbent case. Additionally, the solid sorbent technology has a slightly lower cooling water demand. The electricity requirement of the PEI-based sorbent process, however, is higher than that of the MEA process. Table 6 exhibits the specific electricity requirements for both capture systems. The primary contributor to the higher electricity demand is the blower power requirement for RTI's process. This is a result of the solid sorbent process processing the same flue gas volumetric flow rate, but with a higher pressure drop encountered across the process compared to the MEA technology.

Table 6

In comparing the overall energy requirement (steam, electricity, and cooling water) of both processes, it is evident that the PEI-based solid sorbent process performs significantly better than the MEA process due primarily to a reduced steam requirement.

4.2 Equipment Cost and Capital Expenditure

The following lists the equipment and materials of construction that were determined necessary for both capture processes while building up the Aspen simulations and design models:

1. The MEA process requires 4 stainless steel absorber columns each with a diameter of 15.1m and a height of 27.4m (assuming MellapakTM 250Y packing is used). Additionally, the MEA process requires 1 stainless steel stripper column with a diameter of 11.55m and height of 24m. The work of the Abu-Zahra et. al. [2] has shown that in an MEA process

the capital costs are dominated by the absorber/stripper vessels and blower costs. This is confirmed in the presented work as the costs of the equipment are \$63 million (absorbers), \$34 million (stripper), and \$21 million for the flue gas blower.

2. RTI's solid sorbent process requires 4 square absorber vessels with dimensions of 14.9m x 14.9m and a height of 8.3m and 2 square regenerator vessels of dimensions 6m x 6m x 9.4m. The costs of these vessels are \$23 million and \$18 million respectively. The cost of the flue gas blower (\$21 million) is estimated to be the same for both processes. This similar capital cost is related to the fact that similar flue gas flow rate will be treated in both processes. The flue gas flow rate is the major parameter in defining the total sizing and therefore, the cost of the gas blower.

Table 7 provides a comprehensive equipment cost list for each of the CO₂ capture technologies (including the equipment costs listed above). The equipment costs for both the MEA and the solid sorbent processes were estimated using the 2007 cost as the base year and all costs were corrected as per that set point. In calculating the Total Capture Plant Costs (TCPC) for each process it is clear that RTI's process has a significantly lower TCPC cost than the MEA process and is primarily attributable to the lower absorber/regenerator costs.

Table 7

4.3 Cost of Electricity and Cost of CO₂ Capture

Using the cost of electricity (COE) calculation, shown in Equation (1), as well as the calculated costs for CAPEX and fixed and variable OPEX, the COE for each CO₂ capture process is determined. Table 8 summarizes the COE calculation results for each process showing how CAPEX, OPEX, fuel cost, and the CO₂ transport cost all contribute to the bottom-line COE value. The PEI-based solid sorbent process COE represents a significant savings of 3.3 cents per kWh

1 compared to the MEA process. Table 9 summarizes the technical requirements and annual costs
2 for both processes, which are required in the calculation of CO₂ captured cost (Equation (2)).
3 The CO₂ captured cost of RTI's process is roughly \$7 per ton of CO₂ cheaper than the
4 conventional MEA process which is due to the differences in the capital and operating costs, and
5 reduced energy penalty, as described previously.

6 **Table 8**

7 **Table 9**

8 **5. Conclusions**

9
10 The benchmarking study detailed in this work compares the technical and economic impacts of
11 two CO₂ capture technologies: a novel solid sorbent process and a conventional MEA solvent
12 process. This benchmarking was applied to a typical NGCC power plant. The analysis shows that
13 the PEI-based solid sorbent process offers technical and economic advantages over the
14 conventional process. The regeneration energy requirement for the solid sorbent process is
15 estimated at 2.69 GJ/ton CO₂. This is significantly reduced compared to the MEA process
16 (4.14GJ/ton of CO₂). This lower regeneration energy requirement is primarily due to avoiding
17 significant water evaluation in the solid sorbent process. Significantly less steam is utilized in the
18 solid sorbent process to regenerate the same amount of CO₂, leading to the overall energy
19 penalty savings. The solid sorbent outperforms the MEA process on three of the four primary
20 technical performance factors identified in this paper (steam demand, energy demand, cooling
21 water demand). Electricity demand, due to increased pressure drop across the solid sorbent
22 process (and higher blower electricity requirement), is the only area in which the MEA process
23 exhibits slightly better performance.

The design of the solid sorbent process absorber and regenerator lead to significant capital cost benefits compared to the design requirements of the MEA process. The results exhibited in this paper show CAPEX reductions of \$49 million from the initial amount of \$221 million for the MEA process. Additionally, the COE of installing RTI's process technology in a NGCC power plant is lowered from \$85.0/MWh to \$81.2/MWh compared to the MEA process installed in the same power plant.

These results indicate that the PEI-based solid sorbent technology has the potential to improve the technical and economic constraints of deploying CO₂ capture technologies. However further experimental validation and more detailed process analyses and testing are required to fully realize the potential advantages offered by this novel solid sorbent technology.

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References

- [1] B. Metz, *Carbon Dioxide Capture and Storage: Special Report of the Intergovernmental Panel on Climate Change*: Cambridge University Press, 2005.
- [2] M. R. Abu-Zahra, L. H. Schneiders, J. P. Niederer, P. H. Feron, and G. F. Versteeg, "CO₂ capture from power plants: Part I. A parametric study of the technical performance based on monoethanolamine," *International Journal of Greenhouse Gas Control*, vol. 1, pp. 37-46, 2007.
- [3] C. Dincă, A.-A. Badea, T. Apostol, and G. Lăzăroiu, "GHG emissions evaluation from fossil fuel with CCS," *Environmental Engineering and Management Journal*, vol. 8, pp. 81-89, 2009.
- [4] A. Lawal, M. Wang, P. Stephenson, and H. Yeung, "Dynamic modelling of CO₂ absorption for post combustion capture in coal-fired power plants," *Fuel*, vol. 88, pp. 2455-2462, 2009.
- [5] M. Lucquiaud, O. Erry, H. Chalmers, X. Liang, J. Gibbins and M. Abu Zahra, "Techno-economic assessment of future-proofing coal plants with post combustion capture against technology developments", *Energy procedia*, vol 4, pp 1909-1916, 2011.
- [6] EIA, *Analysis Briefs for the US Energy Information Administration, United Arab Emirates*: 2013
- [7] R. Rivera-Tinoco and C. Bouallou, "Comparison of absorption rates and absorption capacity of ammonia solvents with MEA and MDEA aqueous blends for CO₂ capture," *Journal of Cleaner Production*, vol. 18, pp. 875-880, 2010.

- [8] A. T. Zoghi, F. Feyzi, and S. Zarrinpashneh, "Experimental investigation on the effect of addition of amine activators to aqueous solutions of N-methyldiethanolamine on the rate of carbon dioxide absorption," *International Journal of Greenhouse Gas Control*, vol. 7, pp. 12-19, 2012.
- [9] Y. Artanto, J. Jansen, P. Pearson, T. Do, A. Cottrell, E. Meuleman, and P. Feron, "Performance of MEA and amine-blends in the CSIRO PCC pilot plant at Loy Yang Power in Australia," *Fuel*, vol. 101, pp. 264-275, 2012.
- [10] M. Karimi, M. Hillestad, and H. F. Svendsen, "Capital costs and energy considerations of different alternative stripper configurations for post combustion CO₂ capture," *Chemical Engineering Research and Design*, vol. 89, pp. 1229-1236, 2011.
- [11] A. Cousins, L. Wardhaugh, and P. Feron, "A survey of process flow sheet modifications for energy efficient CO₂ capture from flue gases using chemical absorption," *International Journal of Greenhouse Gas Control*, vol. 5, pp. 605-619, 2011.
- [12] Z. Amrollahi, I. S. Ertesvåg, and O. Bolland, "Optimized process configurations of post-combustion CO₂ capture for natural-gas-fired power plant—Exergy analysis," *International Journal of Greenhouse Gas Control*, vol. 5, pp. 1393-1405, 2011.
- [13] Z. Amrollahi, P. A. M. Ystad, I. S. Ertesvåg, and O. Bolland, "Optimized process configurations of post-combustion CO₂ capture for natural-gas-fired power plant—Power plant efficiency analysis," *International Journal of Greenhouse Gas Control*, vol. 8, pp. 1-11, 2012.
- [14] X. Xu, C. Song, J. M. Andresen, and B. G. Miller, "Absorption separation of CO₂ from simulated flue gas mixtures by novel CO₂ "molecular basket" absorbents," *International journal of environmental technology and management*, vol. 4, pp. 32-52, 2004.
- [15] X. Xu, C. Song, J. M. Andresen, B. G. Miller, and A. W. Scaroni, "Novel polyethylenimine-modified mesoporous molecular sieve of MCM-41 type as high-capacity absorbent for CO₂ capture," *Energy & Fuels*, vol. 16, pp. 1463-1469, 2002.
- [16] X. Xu, C. Song, J. M. Andrésen, B. G. Miller, and A. W. Scaroni, "Preparation and characterization of novel CO₂ "molecular basket" absorbents based on polymer-modified mesoporous molecular sieve MCM-41," *Microporous and mesoporous materials*, vol. 62, pp. 29-45, 2003.
- [17] X. Xu, C. Song, B. G. Miller, and A. W. Scaroni, "Absorption separation of carbon dioxide from flue gas of natural gas-fired boiler by a novel nanoporous "molecular basket" absorbent," *Fuel processing technology*, vol. 86, pp. 1457-1472, 2005.
- [18] M. C. Woods, P. Capicotto, J. L. Haslbeck, N. J. Kuehn, M. Matuszewski, L. L. Pinkerton, M. D. Rutkowski, R. L. Schoff, and V. Vaysman, "Cost and performance baseline for fossil energy plants. Volume 1: Bituminous coal and natural gas to electricity final report," *National Energy Technology Laboratory*, 2007.
- [19] A. Adeosun and M. Abu-Zahra, "Evaluation of amine-blend solvent systems for CO₂ post combustion capture applications", *Energy Procedia*, vol 37, pp 211-218, 2011.
- [20] S. Murai, Y. Kato, Y. Maezawa, T. Muramatsu and S. Saito, "Novel Hindered Amine Absorbent for CO₂ Capture," *Energy procedia*, vol. 37, pp 417-422, 2013.
- [21] W-J. Choi, J-B. Seo, S-Y. Jang, J-H. Jung and K-J. Oh, "Removal characteristics of CO₂ using aqueous MEA/AMP solutions in the absorption and regeneration process". *Journal of Environmental Sciences*, vol 21: pp 907-913, 2009.

- [22] G. Rochelle, E. Chen, S. Freeman, D. Wagener, Q. Xu and A. Voice, "Aqueous piperazine as the new standard for CO₂ capture technology". *Chemical Engineering*. Vol 171 (3), pp 725-733, 2011.
- [23] F. Chowdhury, H. Okabe, H. Yamada, M. Onoda and Y. Fujioka, "Synthesis and selection of hindered new amine absorbents for CO₂ capture". *Energy Procedia*, vol 4: pp 201-208, 2011.
- [24] Mitsubishi Heavy Industries, "Latest Advancements in Post Combustion CO₂ Capture Technology for Coal Fired Power Plant. The 9th Annual Conference on Carbon Capture and Sequestration, Pittsburgh, Pennsylvania, 2010.
- [25] T. Nelson, L. Coleman, A. Kataria, M. Lail, M. Soukri, D. Quang, and M. Abu-Zahra, "Advanced Solid Sorbent-Based CO₂ Capture Process". *Energy Procedia*, vol 63, pp 2216–2229, 2014.
- [26] T. Nelson, L. Coleman, P. Mobley, A. Kataria, J. Tanthana, M. Lesemann, and L. Bjerger, "Solid Sorbent CO₂ Capture Technology Evaluation and Demonstration at Norcem's Cement Plant in Brevik, Norway". *Energy Procedia*, vol 63, pp 6504–6516.2014.
- [27] T. Nelson, L. Coleman, J. Farmer, A. Kataria, M. Lail, M. Soukri, and J. Tanthana, "Advanced Solid Sorbents and Process Designs for Post-Combustion CO₂ Capture . DOE NETL CO₂ Capture Technology Meeting, Pittsburgh, PA, 2014.
- [28] T. Nelson, "Advanced Solid Sorbents and Process Designs for Post-Combustion CO₂ Capture. NETL CO₂ Capture Technology Meeting, Pittsburgh, PA, 2013.
- [29] E. S. Rubin, C. Short, G. Booras, J. Davison, C. Ekstrom, M. Matuszewski and S. McCoy, " A proposed methodology for CO₂ capture and storage cost estimates," *International Journal of Greenhouse Gas Control*, vol. 17, pp. 488-503, 2013.

Abbreviations

CCS	Carbon Capture and Storage
MEA	Monoethanolamine
DOE	United States Department of Energy
PSU	Pennsylvania State University
RTI	Research Triangle Institute International
NGCC	Natural Gas Combined Cycle
PEI	Polyethyleneimine
MBS	Molecular Basket Sorbent
TDH	Transport Disengaging Height
CCF	capacity charge factor (for the NGCC power plant = 0.105, for NGCC with capture = 0.1111)
TOC	Total Overnight Cost
OC _{FIX}	Total Fixed Operating Cost
CF	Capacity Factor of 0.85
OC _{VAR}	Total Variable Operating Cost at 100% full capacity
MWH	Total Annual Megawatt Hours Generated
HRSG	Heat Recovery Steam Generator
BEC	Bere-Erected Cost
CAPEX	Capital Expenditures
OPEX	Operational Expenditures

1	TS & M	Transporting, Storing and Monitoring
2	O&M	Operating and Maintenance
3	PPMV	Parts Per Million by Volume
4	TCW	Tempered Cooling Water
5	v	Superficial Gas Velocity
6	$\dot{V}$	Gas Volumetric Flow Rate
7	n	Number of vessels
8	E	Edge Length of the square cross section of the column
9	HTH	Heat Transfer Height
10	LMTD	Logarithmic Mean Temperature Difference
11	OD	Outer tube Diameter
12	ID	Internal tube Diameter
13	TCPC	Total Capture Plant Costs
14	COE	Cost of Electricity
15	TEG	Triethylene Glycol
16	ICARUS	ASPEN Process Economic Analyzer
17	LHV	Low Heating Value
18	TPC	Total Project Cost
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Figure 1: Benchmarking Methodology

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Figure 2: NGCC Power Plant Definition (adapted from [18])

Figure 3: MEA Process Flow Diagram and Assumptions

Figure 4: PEI-based Solid Sorbent Process Flow Diagram and Assumptions

Figure 5: MEA Absorber/Stripper Design Model [2]

Figure 6: PEI-based Solid Sorbent Process Absorber / Regenerator Design Model

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Table 1: Reference Power Plant Operational Performance [18]

Variable	Units	Value
Number of Gas Turbines	-	1
Gas Turbine Power	MWe	362.2
Fuel Temperature	°C	38
Steam Turbine Power	MWe	202.5
Total Auxiliaries	MWe	9.62
Net Power Poutput	MWe	555
Thermal Power Input (LHV Basis)	MWth	997
Net Plant Efficiency (LHV Basis)	%	55.7
Flue Gas Specifications		
Temperature	°C	106
Pressure	atm	1
Composition	mol %	
CO ₂		4.04
N ₂		74.32
H ₂ O		8.67
O ₂		12.09
H ₂ S		0
Ar		0.89
Flue Gas Mass Flow Rate	kg/s	897.4

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Table 2: Components of Capital Cost Estimation

Element	Components	Used in this Study	Source
Bare Erected Cost (BEC) +	Equipment and Setting Piping Civil Structural Steel Instrumentation Electrical Insulation Paint	Assumptions from Process Economic Analyzer	-
Engineering, Procurement and Construction Costs (EPCC) +	Class II (similar to AACE Class 3)	10% of BEC	USDOE [18]
Process Contingency Cost +	Class II (similar to AACE Class 3)	No process contingency for MEA 20% of BEC Process contingency for PEI-based Solid Sorbent Process	USDOE [18]
Project Contingency=	Class II (similar to AACE Class 3)	25% of BEC	USDOE [18]
Total Plant Cost (TPC) +	-	Summed as shown on the left	USDOE [18]
Owners Costs=	Pre-production costs Inventory capital Cost of land	6 months operating labor 1 month maintenance material at full capacity 1 month non-fuel consumables at full capacity 25% of one month's fuel cost at full capacity 2% of Total Plant Cost Inventory capital are 0.5% of Total Plant Cost for spare parts 60 day supply of fuel and non-fuel consumables at full capacity land required is 1000 acres charged at \$3000/acre.	USDOE [18]
Total Overnight Cost (TOC)	-	Summed as shown on the left	USDOE [18]

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Table 3: Economic Evaluation Assumptions

Parameter	Unit	Value	Comments
Plant life time	years	30	-
Construction Period	years	3	-
Economic Analysis Period	years	33	-
Plant Capacity Factor (CF)	%	0.85	
Capacity Charge Factor (without capture) (CCF)	%	0.105	For capital cost escalation and inflation in current dollars
Capacity Charge Factor (with capture) (CCF)	%	0.111	For capital cost escalation and inflation in current dollars
Fuel Cost	\$/GJ	7.85	As per US DOE Report [18]
MEA Cost	\$/ton	2249.89	As per US DOE Report [18]
Operating Labor Rate	\$/h	34.65	As per US DOE Report [18]
Cooling Water	\$/ton	0.2592	As per US DOE Report [18]

Table 4: Absorber and Regenerator Design and Performance Parameters

Parameter	MEA Process	MBS Process
Capture Rate	90%	90%
Regeneration Rate	95%	95%
Superficial Gas Velocity	0.6m/s	0.6m/s
Solvent Losses	0	25kg / 1000 tons of sorbent
Cooling Water In (oC)	15.6	15.6
Cooling Water Out (oC)	26.7	26.7
Absorber Type	Regular absorber with Mellapak packing	Square cross section with horizontal tubes for heat transfer
Regenerator Type	Regular regenerator with Mellapak packing	Square cross section with horizontal tubes for heat transfer
MOC	Stainless Steel	Concrete vessels with carbon steel tubes for heat transfer
Rotating Equipment MOC	Stainless Steel	Stainless Steel
Static Vessel MOC	Stainless Steel	Stainless Steel

1 Table 5: PEI-based Solid Sorbent Process Technical Performance Benchmarking

Variable	MEA Process	PEI-based Solid Sorbent
Regeneration Steam (kg/s)	64.65 [*]	48.11
Energy Requirement (GJ/ton CO ₂)	4.14	2.69
Electricity (kWe)	33,779	39,233
Cooling Water (kg/s)	5654.69	4926.82

^{*}Based on equivalency factor = 0.1716 kWh/kg-steam

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Table 6: Power Requirements of MEA and PEI-based Solid Sorbent Processes

Power Requirements	MEA	PEI-based Solid Sorbent
Flue Gas Blower Requirement (kWe)	9.6	20.0
Pumping Requirement (MWe)	5.677	1.0
Compression Requirement (MWe)	18.5	18.5
Sorbent Transport Requirement (MWe)	0	0.39

1 Table 7: Total Overnight Cost of MEA and PEI-based Solid Sorbent Processes

Capital Cost	MEA	PEI-based Solid Sorbent
Equipment cost, \$	121,004,776	81,181,321
Bare erected cost, \$	164,130,471	111,868,067
Eng CM H.O. & Fee, \$	16,413,047	11,186,807
Process contingency, \$	-	22,373,613
Project contingency, \$	41,032,618	27,967,017
Total Capture Plant Cost, \$	221,576,135	173,395,503
Total Power Plant Cost, \$	340,367,000	340,367,000
Total Project Cost (TPC), \$	561,943,135	513,762,503
Total Overnight Cost, \$	686,492,435	638,066,007

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Table 8: COE Calculation

Cost of Electricity		
	MEA	PEI-based Solid Sorbent
Capital cost, \$/MWh	21.6	19.6
Fixed cost, \$/MWh	5.6	5.5
Variable cost, \$/MWh	2.6	2.5
Fuel cost, \$/MWh	52.1	50.9
CO ₂ TS&M, \$/MWh	3.2	3.2
Cost of electricity, \$/MWh	85.0	81.7

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Table 9: Cost of CO₂ Capture

Variable	Units	MEA	PEI-based Solid Sorbent
Annual Costs			
Treatment and Disposal	\$/yr	5,595,634	5,888,720
Materials and Utilities	\$/yr	43,533,350	39,884,274
O&M Levelized Costs	\$/yr	308,669	308,669
Capital Charge	\$/yr	38,775,824	30,344,213
Economic Parameters			
Total Project Cost	\$	221,576,135	173,395,503
Cost of CO ₂ Capture	\$/ton-CO ₂	69.82	62.13
Percentage Decrease in Plant Efficiency (LHV basis)	%	8%	6.30%

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