Assessment of advanced solvent-based post-combustion CO₂ capture processes using a bi-objective optimization technique

Charles A Kang^{a,*}, Adam R Brandt^a, Louis J Durlofsky^a, Indira Jayaweera^b

^a*Department of Energy Resources Engineering, Stanford University, Green Earth Sciences Building, Stanford, California, USA*

^b*SRI International, Menlo Park, California, USA*

Abstract

The optimized performance of two advanced CO₂ capture processes is compared to that of a monoethanolamine (MEA) baseline for a gas-powered CO₂ capture retrofit of an existing coal-fired facility. The advanced temperature-swing processes utilize piperazine and mixed-salt solvents. The mixed-salt treatment involves the use of ammonia for CO₂ absorption and potassium carbonate primarily to control ammonia slip. The processes are represented in terms of energy duty requirements within a modular heat integration code developed for CO₂ capture modeling and optimization. The model includes a baseload coal plant, a gas-fired subsystem containing gas turbines and a heat recovery steam generator (HRSG), and a CO₂ capture facility. A formal bi-objective optimization procedure is applied to determine the design (e.g., detailed HRSG components and pressure levels, gas turbine capacity, CO₂ capture capacity) and time-varying operations of the facility to simultaneously maximize net present value (NPV) and minimize total capital requirement (TCR), while meeting a maximum CO₂ emission intensity constraint. For a realistic scenario constructed using historical data, optimization results indicate that both advanced processes outperform MEA in both objectives, and the mixed-salt process in turn outperforms the piperazine process. Specifically, for the scenario considered, the base case mixed-salt process achieves 16% greater NPV and 14% lower TCR than the MEA process, and 10% greater NPV and 5% lower TCR than the piperazine process. A five-case sensitivity study of the mixed-salt process indicates that it is competitive with the piperazine process and consistently outperforms the MEA process.

Keywords: CO₂ capture, technology assessment, MEA, piperazine, mixed-salt, process optimization, MINLP, bi-objective optimization

1. Introduction

The current baseline technology for post-combustion CO₂ capture from fossil-fuel-fired power plants is a temperature-swing process using an aqueous monoethanolamine (MEA) solvent. Though effective, the MEA process is characterized by high solvent regeneration heat duty requirements. Thus, researchers have recognized the importance of improved CO₂ capture technologies, and substantial research efforts have been undertaken to develop advanced capture processes that use solvents other than MEA [1, 2]. Two such processes that have been developed in recent years are the piperazine process and the mixed-salt process [3, 4]. In this study, we use a consistent optimization framework to optimize the design of a CO₂-capture-enabled facility using the two advanced processes and the MEA baseline. The optimization results are then used to calculate quantitative estimates of the economic value of the advanced technologies.

Previous studies have assessed the performance of advanced CO₂ capture processes. Rochelle et al. [3] considered the piperazine process and found that it significantly outperformed the MEA process, with reduced reboiler heat duty and compressor work duty. Other independent work similarly found outperformance by piperazine over MEA [5, 6]. Sanchez Fernandez et al. [7] considered a process using a

*Corresponding author. Tel: +1 650 479 6189

Email addresses: cak@alumni.stanford.edu, cakang@outlook.com (Charles A Kang), abrandt@stanford.edu (Adam R Brandt), lou@stanford.edu (Louis J Durlofsky), indira.jayaweera@sri.com (Indira Jayaweera)

Preprint submitted to Elsevier

July 14, 2016

blend of piperazine and 2-Amino-2-Methyl-Propanol (AMP), and found that the mixed piperazine-AMP process also outperformed MEA. Hopkinson et al. [8] described an approach to tune the heat of reaction of an idealized ‘conceptual’ solvent to minimize the work requirement for CO₂ capture, as a guide for future solvent development. Using an optimal design and control approach that is somewhat similar to the approach we take in this work, Sharifzadeh et al. [9] found that a new amine-promoted buffer salt also outperformed MEA. Other studies have considered non-solvent-based CO₂ capture approaches, which have their own advantages and disadvantages [10–13].

We assess the economic performance of the advanced processes and the MEA baseline process by modeling the processes within an engineering-economic optimization framework, considering a CO₂-capture-enabled coal-natural gas power station. In this framework, which we first developed using MEA [14], we optimize the detailed heat integration design and operational profile of an auxiliary gas-powered CO₂ capture retrofit of a coal-fired power station. This type of configuration notably differs from retrofit configurations in which the energy used for CO₂ capture is taken parasitically from the coal plant. Key advantages of the auxiliary gas-powered retrofit concept are that this approach reduces the number of modifications required of the existing coal plant, and that the generation capacity of the overall facility is not reduced, which could be of great importance in supply-constrained power grids.

The facility model is designed to be easily usable within a computational optimization framework. The model consists of a set of fit-for-purpose system component modules that exchange mass and energy flows with each other and the model environment. By taking a modular approach, we are able to achieve accurate results with high computational efficiency, which is necessary because computational optimization requires thousands to millions of facility model evaluations. Our representation allows for the consideration of detailed heat integration design, which has been shown to yield significant benefits, alongside consideration of realistic time-varying electricity markets, which can have a large impact on CCS economics [10, 12, 15–18].

The design and time-varying operation of the facility are optimized using computational methods to maximize net present value (NPV) and minimize total capital requirement (TCR). The two objectives of maximum NPV and minimum TCR are key quantities in any investment decision, and can be in conflict with each other. Our procedure identifies facilities that optimize for each of these objectives separately, and also identifies the optimal tradeoff between the two objectives via the construction of the Pareto frontier for these two objectives. Optimization results for the advanced CO₂ capture processes are compared to optimization results for the baseline MEA technology to quantify the value of the advanced processes.

In Kang et al. [19] and Kang [20], we identified two alternative approaches for modeling the CO₂ capture system within the context of the overall facility. In the ‘full-physics’ representation that was the focus in Kang et al. [19], the CO₂ capture system was modeled using a rate-based Aspen Plus process simulation (or a statistical fit of Aspen Plus model output). In the ‘simplified-capture’ representation, which was employed in Kang et al. [14] and Brodrick et al. [21], the CO₂ capture system was represented using a variable for system capacity and constant parameters for a small number of performance specifications. In [19] we showed that the two modeling approaches provided similar high-level designs and objective function values for the cases considered, which suggests that adequate optimization results are obtainable at much less computational effort using the simplified-capture representation. This earlier work, however, addressed only MEA-based CO₂ capture processes. Here, we extend the simplified-capture representation to model the piperazine and novel mixed-salt processes. Moreover, we show that the results from these optimizations can be used to quantify the value of these two advanced CO₂ capture technologies. Finally, we note that the method applied here in the context of a gas-fired CO₂ capture retrofit is quite general, and it could be used for a wide range of system configurations and CO₂ capture technologies. As an example, an earlier form of the model used here was modified to consider integration with solar thermal energy [21].

This paper proceeds as follows. In Section 2, we describe the facility model, with which we compute system behavior given a candidate design and operating plan. The optimization framework, which systematically explores the design space, is described in Section 3. Section 4 contains results and discussion of our optimizations for the three CO₂ capture processes. Concluding remarks are provided in Section 5.

2. Facility model

Figure 1 shows a schematic of the facility considered in this work, which is the same as that treated in our earlier studies [14, 19, 20]. The facility consists of a coal-fired power station (CP), a gas-fired combined-cycle gas turbine (CCGT), and a post-combustion CO₂ capture system. The CP has a fixed size of 440 MW, while the capacities of the CCGT and CO₂ capture system are controlled by design decision variables that are determined during the course of optimization. The capacity of an individual gas turbine is constrained to be at most 300 MW, and the total gas turbine capacity (the facility may use more than one gas turbine) is constrained to be between 1 MW and 400 MW. Flue gas from the CP (a baseload plant operated at a constant rate) is scrubbed in the CO₂ capture system, which draws steam for solvent regeneration from the CCGT system. Each of the modules shown in Figure 1 is modeled by a system of nonlinear algebraic equations that derive from component-scale energy and mass balances.

Figure 1: Schematic of CO₂-capture-enabled facility. Adapted from Kang et al. [19].

The process model is developed for the purpose of use in computational optimization. Because optimization requires large numbers ($\sim 10^5 - 10^6$) of function evaluations, computational efficiency is of great importance. The modular design of the process model allows different system components to be modeled at different levels of detail, enabling us to employ customized models for each component that calculate required quantities accurately and with great speed. Consequently, the overall facility model is able to represent realistic system behavior with low computational cost.

In the present work we compare different alternatives for the CO₂ capture process, so our discussion here focuses on the CO₂ capture system model. We also describe certain aspects of the steam cycle model, as the CO₂ capture process is closely linked to the steam cycle (since steam is used to supply heat for solvent regeneration). Our representation of capital costs and electricity markets is also discussed here, as a key feature of our model is its ability to resolve interactions between facility design and time-varying operations. This capability is needed to quantify the economic value of operational and design choices consistent with the advanced capture technologies. Note that some of the discussion presented here, while necessary to understand this work, follows closely after Kang et al. [14]. Additional details on the CO₂ capture process and steam cycle models, along with descriptions of other system component models, can be found in [14, 22]. We reiterate that only the MEA-based capture process was considered in [14, 22].

2.1. CO₂ capture

We consider three different solvent-based CO₂ capture processes. We use the aqueous MEA process as our baseline for comparison with the more advanced mixed-salt and piperazine processes [1, 3, 4]. We first summarize the MEA process, and then describe how the advanced processes differ from the MEA process.

Figure 2: Conceptual diagram of the MEA and piperazine processes, which differ largely in the solvent used (CO_2 compression train not shown).

In the MEA process, shown schematically in Figure 2, flue gas CO_2 is absorbed by an aqueous MEA solvent in an absorption column. This produces a CO_2 -rich solvent stream. The solvent is regenerated in the regeneration column by the application of heat in the reboiler. The resulting CO_2 -lean solvent stream is then recycled into the absorption column, while the CO_2 that is desorbed in the regeneration column is dewatered in the condenser and then compressed to 14 MPa for pipeline transport. In addition, heat recovery between the CO_2 -lean and CO_2 -rich solvent streams is performed to improve thermal efficiency.

In this work we do not consider the detailed design of the capture processes. Instead, we represent the capture processes in terms of their CO_2 capture capacity and energy duty requirements. Using this approach, we represent the design of the CO_2 capture system using a single continuous decision variable (capture capacity) alongside several constant parameters that specify the energy duty requirements. We apply constant specific duty requirements for solvent regeneration heat, compressor work, and pump work (for pumps and blowers). For the MEA process, we employ the following specific duty requirements: solvent regeneration heat, $3.68 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ provided at 407 K; compressor work, $375 \text{ kJ}_e/\text{kg CO}_2$; and pump work, $40 \text{ kJ}_e/\text{kg CO}_2$. The same duty requirements were used in our previous work [14, 22], and are based upon values provided by the Integrated Environmental Control Model (IECM), version 8.0.2, and other published literature [23–25]. Our representations of all of the CO_2 capture processes, including the MEA process, are consistent with an assumed reboiler approach temperature of 10 K.

The piperazine process is very similar to the MEA process, and uses similar equipment but with a different solvent. Some piperazine process designs use a two-stage flash instead of a regeneration column (as is used in the MEA process) for solvent regeneration [3], but this level of detail is not resolved in our representation of the process. The piperazine process has specific regeneration heat duty of $2.6 \text{ MJ}_{\text{th}}/\text{kg CO}_2$, provided at 433 K (618 kPa steam) [1, 3]. Because the piperazine process operates at higher pressure than the MEA process, the compression work duty of $220 \text{ kJ}_e/\text{kg CO}_2$ is less than that of the MEA process. We assume that pump work duty for solvent pumps and blowers is equal to that of the MEA process, $40 \text{ kJ}_e/\text{kg CO}_2$.

The mixed-salt process is also generally similar to the MEA process, but is more complex and involves two-stage absorption and regeneration columns. The key idea of the mixed-salt process is to use ammonia to absorb CO_2 from the flue gas, and potassium carbonate to control ammonia slip [4]. Figure 3 shows a conceptual design of the mixed-salt process. The lower bed in the absorption column has ammonia-rich solvent at the bottom (with ammonia concentration decreasing upwards), and the upper bed has ammonia-lean solvent. This enables the ammonia to be captured in the upper bed, which decreases ammonia loss. In order to maintain this differential concentration in the absorption column, two solvent streams (one rich in ammonia, and the other rich in potassium) are taken from the regeneration

Figure 3: Conceptual diagram of the mixed-salt process with compression train not shown. Adapted from Jayaweera et al. [4].

Table 1: Key operating parameters for the MEA, piperazine, and mixed-salt CO_2 capture processes.

Quantity	Units	MEA	Piperazine	mixed-salt
Reboiler steam temp.	K	407	433	433–443 and 403–413
Reboiler steam pressure	kPa	300	618	618–790 and 270–360
Specific reboiler heat	$\text{MJ}_{\text{th}}/\text{kg CO}_2$	3.68	2.6	0.44–0.66 and 1.76–2.65
Specific compress. work	$\text{kJ}_e/\text{kg CO}_2$	375	220	176–264
Specific pump work	$\text{kJ}_e/\text{kg CO}_2$	40	40	40

column. This entails the application of heat in the regeneration column at two different temperatures (shown in Figure 3 with two reboilers).

For the base case, the mixed-salt process specific regeneration heat duty is specified as follows: $0.44 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 433 K (618 kPa steam), and $1.76 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 403 K (270 kPa steam), for a total of $2.20 \text{ MJ}_{\text{th}}/\text{kg CO}_2$. As in the piperazine process, solvent regeneration in the mixed-salt process occurs at high pressure, so compressor work duty is less than in the MEA process. Specific compressor work duty is $176 \text{ kJ}_e/\text{kg CO}_2$. These specifications are based on early experimental results and modeling performed at SRI International [4]. We also assume that work duty for blowers and pumps is the same as in the MEA process, $40 \text{ kJ}_e/\text{kg CO}_2$.

Because the mixed-salt process is at an earlier stage of development, and thus more uncertain, than the piperazine and MEA processes, we perform a sensitivity study on the specific reboiler heat duty, reboiler approach temperature, and specific compressor work duty. In the first two sensitivity cases, we vary specific reboiler heat duty by increments of +25% ($0.55 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 433 K, and $2.20 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 403 K) and +50% ($0.66 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 433 K, and $2.65 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 403 K). In the third sensitivity case, reboiler approach temperature is increased from 10 K to 20 K, which results in steam requirements of $0.44 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 443 K (790 kPa steam), and $1.76 \text{ MJ}_{\text{th}}/\text{kg CO}_2$ at 413 K (360 kPa steam). In the fourth and fifth sensitivity cases, specific compressor work duty is increased by increments of +25% and +50% to $220 \text{ MJ}_e/\text{kg}$ and $264 \text{ MJ}_e/\text{kg}$.

Table 1 displays the key operating parameters (or their ranges in the case of the mixed-salt process) for the three processes considered. While it should be clear from Table 1 that the advanced processes outperform the MEA process energetically, it is not possible to quantify the economic value of this energetic outperformance simply by inspecting Table 1. Quantification of the value of the advanced

Figure 4: Comparison of West Texas 2011 base scenario Pareto frontiers in full-physics and simplified-capture representations. Adapted and reproduced from Kang et al. [19].

technologies requires that the processes be modeled within a larger engineering-economic framework, which is accomplished in this work.

2.2. Reliability of simplified-capture representation

Our work in [19, 20] showed that the full-physics and simplified-capture representations resulted in similar performance predictions and similar optimal facility designs for facilities employing MEA-based CO₂ capture. This is evident in Figure 4, which compares optimization results using the two representations. These results are for the West Texas 2011 scenario considered in [19], with minimum TCR and maximum NPV as the two objectives (1p, 2p and 3p refer to the number of pressure levels in the HRSG). Each point in Figure 4 represents a unique facility design and operating profile. The lighter points depict all feasible systems evaluated during the optimizations, and the bold points along the upper edge define the so-called ‘Pareto frontier.’ These ‘Pareto-efficient’ points are the key optimization results, as they define the optimal tradeoff between the two objectives (the Pareto frontier is discussed in more detail in Section 4.1).

The important observation here is the generally close agreement between the Pareto frontiers in Figures 4(a) and (b). Both approaches, for example, identify one-pressure HRSG designs for TCR minimization, and three-pressure HRSG designs for NPV maximization. The full-physics representation, however, provides more detailed information on the design of the CO₂ capture system, and in particular can account for the fact that the number of CO₂ capture trains must be a whole number. This effect can be seen in the discontinuity indicated in Figure 4(a): the points to the left of the discontinuity have two CO₂ capture trains, while the points to the right have three CO₂ capture trains.

Nevertheless, the overall close match between the two treatments gives us confidence that the representation of the advanced CO₂ capture processes that we employ here, which is a modified form of the simplified-capture representation, yields results that are consistent with a more physically detailed technique. Based on the consistency observed in [19, 20], we expect that the results presented here provide a reliable assessment of the relative economic performance of the three processes considered. Moreover, applying the simplified-capture representation for all three of the processes (MEA, piperazine, and mixed-salt) ensures that the different technologies are compared at similar levels of detail.

2.3. Steam cycle

In the CCGT system, steam produced in the HRSG is expanded in steam turbines, which drive electric generators. We model the steam turbines as multi-stage expansions in which each stage is modeled using an isentropic efficiency as described in Kim and Ro [26]. The flow rates and thermodynamic

conditions of the steam provided to the steam cycle are controlled by the HRSG design decision variables, which are determined during the course of optimization. After being expanded in the steam turbines, the steam is condensed and the condensate is used as feed to the HRSG. In our system, the steam can be used in three ways: 1) steam is expanded fully to the condenser temperature (330 K, 17.2 kPa), which maximizes electrical output from the steam cycle but does not provide steam for use in solvent regeneration; 2) steam is extracted from the low pressure steam turbine at a partially expanded state (the condition of which depends upon the design of the facility) to provide heat for solvent regeneration; or 3) steam from the HRSG bypasses the steam turbines entirely and is used directly to provide heat for solvent regeneration. The capacity of the steam turbines is determined by the amount of steam produced in the HRSG, which is calculated from the design decision variables for the HRSG and gas turbine. These different uses of steam correspond to the different modes of system operation, as discussed in Section 2.5. The operating modes (including use of steam) considered in this work are the same as in our earlier work [14, 19].

A key modification employed here, beyond what is described in our earlier work, is that the steam cycle model has been extended to accommodate steam extraction from the low pressure turbine at more than one pressure. This is needed to model the mixed-salt process accurately because this process requires heat supply at two different temperatures. We model this dual steam extraction as follows: a fraction of the steam is first extracted from the low pressure turbine at the higher pressure (higher temperature), and the remainder is extracted at the lower pressure (lower temperature). The steam extraction pressure is a design decision variable. At the end of an optimization run, this value typically converges to the steam saturation pressure at the temperature required by the CO₂ capture process. Note that all steam extractions are taken from the low pressure steam turbine, so the low pressure outlet of the HRSG is constrained in the optimization to be at least the highest steam extraction pressure required by the capture process.

The condensate return temperature of steam used in the reboiler depends upon the CO₂ capture system treated, and is equal to the temperature of the heat required for the solvent regeneration process. For the MEA process, the temperature is 407 K; for the piperazine process, the temperature is 433 K; and for the mixed-salt process, the temperature is 427 K in the base case (this is the enthalpy-weighted average of the two reboiler heat requirements). Additional details of our model of the steam cycle are available in Kang et al. [14].

2.4. Capital cost

Capital cost for the facility is calculated here using using a bottom-up procedure based upon the Guthrie method [27–29], which was employed in our earlier work [14, 19]. In this procedure, the capital cost of the entire facility is calculated by summing the purchased equipment costs of the individual components within the facility, after a sequence of multipliers is applied to each component to represent various indirect costs such as labor, piping, escalation, and taxes. The purchased equipment cost of each component is calculated based on the size of the component using the following exponential scaling rule [29]:

$$\frac{C_{PE}}{C_{PE,ref}} = \left(\frac{S}{S_{ref}} \right)^\alpha, \quad (1)$$

where C_{PE} [\$] is the purchased equipment cost of an individual component, $C_{PE,ref}$ [\$] is the reference purchased equipment cost of that component, S_{ref} is the reference size of the item, and α is the scaling exponent. The values of $C_{PE,ref}$, S_{ref} , and α , and of the indirect cost multipliers, are the same as in Kang et al. [14] (see Tables 4 and 5 in the Supplementary Material of Kang et al. [14]). This procedure results in a Class 4 capital cost estimate, which corresponds to approximately $\pm 30\%$ accuracy.

We use the same capital cost scaling parameters for the advanced capture processes as for the MEA process (i.e., the capital cost parameters for a CO₂ capture system are not a function of the type of process – MEA, piperazine, or mixed-salt – that is used). We proceed in this way because differences in the capital cost of the different processes are below the level of resolution of our models, given that our method represents the CO₂ capture system using a single capacity parameter. This treatment is also useful because a core purpose of this study is to quantify the economic value of the advanced technologies, and this is aided (as we will see) by holding the capital cost parameters for the three different technologies

constant across optimization runs. The results of the optimizations can then be interpreted as indicating the value provided by the advanced technologies.

Additionally, the piperazine and mixed-salt processes likely both have similar capital costs to the MEA process because they use similar equipment to the MEA process. Moreover, changes in capital costs for individual components in the advanced processes may offset each other to some extent. For example, increases in the cost of the regeneration column with the mixed-salt process (which result because the column operates under higher pressure in this process and thus may require thicker walls) would be offset by decreases in the capital cost of the compressors (which result because the compressors would be required to supply a smaller pressure ratio due to the increased regeneration pressure).

2.5. System operations

As in our previous work [14, 19], we represent time-varying facility operations by defining four discrete operating modes for the facility, dispatched according to the electricity price-duration curve. These four modes correspond to different operating conditions of the CCGT and CO₂ capture system. The coal plant operates at a constant rate in all of these modes. The modes are as follows:

- A. CCGT power only: Steam is expanded to condenser conditions, which maximizes electricity output.
- B. CCGT steam extraction: Steam is extracted at a partially expanded state from the low pressure steam turbine and is used to supply heat for solvent regeneration.
- C. CCGT steam diversion: Steam is diverted around the steam turbines, and is used only to provide heat for solvent regeneration.
- D. Coal only: All components of the CCGT system are deactivated.

The gas turbine is operated at full power in modes A, B, and C. The total net power output from the facility is ordered by mode, A–B–C–D, with mode A having the greatest net power output and mode D having the least.

The operating profit of each mode is a function of the costs of fuel, operations and maintenance (O&M), CO₂ emissions, and price of electricity. We model costs of fuel, O&M, and CO₂ as constant, while we model the wholesale price of electricity as varying on an hourly basis. Note that we employ a CO₂ emissions limit regulation and not a CO₂ tax, though this CO₂ emissions limit induces an implicit price for CO₂ emissions. Assuming that transient effects between modeled hours are of duration of less than one hour, which has been suggested previously [30–32], the preferred operating mode in a given hour is then a function of the hourly price of electricity alone. Consequently, each operating mode is preferred for a specific range of prices. The operating modes are ordered with respect to decreasing price of electricity as A–B–C–D. Because the optimal operating mode depends only on the price of electricity, the modes are separated from each other by threshold prices of electricity, or ‘strike prices.’

We employ the price-duration curve to represent the electricity market. Figure 5 shows the 2011 West Texas hourly electricity prices and the resulting price-duration curve. The price-duration curve consists of hourly wholesale electricity prices sorted from high to low, and as such provides the cumulative duration for which the price of electricity is above a given value. For a given price-duration curve, the operation of the facility is determined by specifying the strike prices. We therefore use the strike prices as decision variables in the optimization to represent the operations of the facility.

3. Optimization problem formulation

We perform bi-objective optimizations for systems that employ the MEA, piperazine, and mixed-salt processes described in Section 2.1. The two objectives considered are maximization of net present value (NPV) and minimization of total capital requirement (TCR). As in Kang et al. [14, 19], we express the optimization problem as follows:

$$\max_{\mathbf{u}_{\text{des}}, \mathbf{u}_{\text{ops}}} [-C_{\text{TCR}}(\mathbf{x}, \mathbf{u}_{\text{des}}), \quad V_{\text{NP}} = -C_{\text{TCR}}(\mathbf{x}, \mathbf{u}_{\text{des}}) + P_{\text{op}}(\mathbf{x}, \mathbf{u}_{\text{des}}, \mathbf{u}_{\text{ops}})] \quad (2)$$

Figure 5: Wholesale electricity hourly price (left) and price-duration curve (right), West Texas Hub Bus, 2011. Figure reproduced from Kang et al. [19].

subject to

$$\mathbf{u}_{\text{des}}, \mathbf{u}_{\text{ops}} \in \Omega \quad (3)$$

$$\mathbf{c}(\mathbf{x}, \mathbf{u}_{\text{des}}, \mathbf{u}_{\text{ops}}) \leq 0. \quad (4)$$

Here $\mathbf{x}$ represents the state variables that are computed in the process model, $\mathbf{u}_{\text{des}}$ represents decision variables for the design of the facility (CO₂ capture system, GT, HRSG, steam cycle), and $\mathbf{u}_{\text{ops}}$ represents the decision variables for the operation of the facility (the strike prices discussed in Section 2.5). The symbol C_{TCR} [\$] represents the TCR objective, V_{NP} [\$] represents the NPV objective, P_{op} [\$] represents capitalized operating profits, and $\mathbf{c}$ represents the optimization constraints.

The calculation of V_{NP} and P_{op} uses the following economic assumptions, which are the same as in [14, 19, 20]: 11% nominal discount rate, 3.3% escalation rate, 30-year time horizon, three-year construction time, 40% corporate income tax rate, and 20-year tax depreciation. The optimization constraints include several physical engineering requirements, along with a regulatory requirement that CO₂ emission intensity of the facility is no greater than 499 kg CO₂/MWh.

Table 2 lists the decision variables used in the optimization problem specified by Equations 2–4. A total of 42 (MEA and piperazine cases) or 43 (mixed-salt) decision variables are present. For the MEA and piperazine cases, there is one decision variable for HRSG steam extraction pressure, while in the mixed-salt case there are two such decision variables (one for each steam extraction).

The optimization problem is solved using the BiPSOMADS algorithm, which is a derivative-free hybrid algorithm that combines Particle Swarm Optimization (PSO) and Mesh Adaptive Direct Search (MADS), and uses a filter method to enforce nonlinear constraints [33, 34]. The BiPSOMADS algorithm displays global search properties (from the PSO component) and can converge to a local optimum (from the MADS component). More details on the optimization procedure can be found in [14, 19, 20].

4. Results and discussion

We perform optimizations using a scenario constructed from data from West Texas in the year 2011. This scenario is the same as that considered in the WTX-Base scenario in [14] and the West Texas 2011 scenario in [19]. The mean hourly electricity price in this scenario is \$40.73/MWh, with standard deviation \$126.74/MWh. The high degree of variability in electricity prices is due to the fact that West Texas has a high level of wind penetration. We selected West Texas for the scenario in part because the high wind penetration of this region may be representative of future power systems. The price of coal is \$1.41/MMBtu, and the price of natural gas is \$4.73/MMBtu.

As mentioned in Section 2.1, the mixed-salt process is at an early stage of development and its use here constitutes a novel system, so its performance parameters are more uncertain than those for

Table 2: Optimization decision variables

Variable	Type	Symbol	# Vars	Units	Lower bound	Upper bound
Number of GTs	Integer	n_{GT}	1	–	1	3
Total GT capacity	Real	$n_{GT}s_{GT}$	1	MW	1	400
HRSG # pressure levels	Integer	N_{pl}	1	–	1	3
HRSG steam pressures	Real	p_{pl}	3	Pa	1×10^5	19×10^6
HRSG steam extraction pressure(s)	Real	p_{ext}	1-2	Pa	3×10^5	5×10^6
HRSG HP steam reheat pressure	Real	$p_{HP,rh}$	1	Pa	1×10^5	20×10^6
HRSG element type & pl	Categ.	$\mathbf{e}_{pl,type}$	14	–	-1	11
HRSG nondimensional element size	Real	$a_{g,0}$	1	–	0.1	3
HRSG relative element sizes	Real	$\mathbf{a}_{g,rel}$	14	–	0.1	3
CCS relative capacity	Real	c_{cap}	2*	Fraction	0.0	0.9
Strike prices	Real	$\mathbf{d}_{sp}$	3	\$/MWh	(0, -20, -30)	(200, 150, 100)

*One variable for absorption and one variable for regeneration and compression. These two variables converge to the same value during optimization.

the piperazine and MEA processes. Therefore, in our optimizations we performed a five-case sensitivity study on the mixed-salt process in which we varied the specific reboiler heat duty, reboiler approach temperature, and specific compressor work duty.

4.1. Comparison using base case mixed-salt process

The advanced CO₂ capture processes yield different values of the minimum TCR and maximum NPV objective functions as compared to MEA, and also result in different system parameters for optimal facilities. We first discuss our findings in terms of the objective functions and then consider the detailed system parameters for the optimized systems.

4.1.1. Objective functions

Figure 6 shows Pareto frontiers for the mixed-salt base case, piperazine, and MEA processes. A Pareto frontier consists of a set of points (facility design and operation specifications) for which improvement in one objective cannot be attained without worsening the other objective. A point for which improvement in both objectives can be achieved simultaneously is called a ‘dominated’ point – such a point is not part of the Pareto frontier. Points on the Pareto frontier, called ‘Pareto-efficient’ points, define the optimal tradeoff between the two objectives. A decision maker would always select a Pareto-efficient point (facility design and operation), as opposed to a dominated point. The particular Pareto point selected depends upon the decision maker’s preference regarding the relative importance of the two objective functions.

Table 3 summarizes the ranges of Pareto frontier objective functions for facilities using each of the CO₂ capture processes. The bottom five entries are for the sensitivity cases discussed in Section 4.2. From Figure 6 and Table 3, it is evident that the mixed-salt process exhibits superior performance for both objective functions compared to the piperazine and MEA processes, while the piperazine process in turn outperforms the MEA process.

Table 4 summarizes the quantitative improvements in TCR and NPV for optimized advanced process facilities with the optimized baseline MEA facilities. The advanced processes outperform MEA on both metrics, and in turn the mixed-salt base case outperforms piperazine. For the maximum NPV objective, the mixed-salt process achieves a NPV that is 10% higher than that for the piperazine process and 16% higher than that for the MEA process. For the minimum TCR objective, the mixed-salt process achieves TCR 5% lower than the piperazine process and 14% lower than the MEA process.

It is informative to compare the improvement in objective functions to the CO₂ capture capital cost (i.e., the portion of the total capital requirement attributed to the CO₂ capture system) using each of the advanced processes. For reference, the range of CO₂ capture capital costs for NPV-maximizing facilities

Figure 6: Comparison of Pareto frontiers for mixed-salt base case, piperazine, and MEA processes.

Table 3: Ranges of NPV and TCR for different CO₂ capture processes

Process	Pareto NPV range [10 ⁶ \$]	Pareto TCR range [10 ⁶ \$]
MEA	93 – 198	346 – 517
Piperazine	140 – 209	314 – 514
Mixed-salt base case	164 – 230	298 – 509
Mixed-salt +10 K reboiler approach temp.	167 – 225	298 – 518
Mixed-salt +25% specific reboiler duty	158 – 229	315 – 517
Mixed-salt +50% specific reboiler duty	134 – 219	336 – 530
Mixed-salt +25% specific compressor duty	170 – 226	301 – 509
Mixed-salt +50% specific compressor duty	158 – 222	299 – 510

Table 4: Improvement relative to MEA baseline of TCR and NPV objective functions due to advanced CO₂ capture processes

Process	Improvement in maximum NPV			Improvement in minimum TCR		
	Absolute [10 ⁶ \$]	Relative [%]		Absolute [10 ⁶ \$]	Relative [%]	
		NPV MEA*	CCS capital†		TCR MEA*	CCS capital†
Piperazine	11	6	5	32	9	16
Mixed-salt base case	32	16	13	48	14	24
MS +10 K reb. appr. temp.	27	14	11	48	14	24
MS +25% spec. reb. duty	31	16	13	31	9	16
MS +50% spec. reb. duty	21	11	8	10	3	5
MS +25% spec. comp. duty	28	14	11	41	13	22
MS +50% spec. comp. duty	24	12	10	47	14	23

* Improvement as percentage of objective function of optimal facility using MEA CO₂ capture process.

† Improvement as percentage of CO₂ capture system capital cost (inclusive of all direct and indirect costs) of optimal facility using advanced process.

Table 5: Key system parameters for maximum NPV and minimum TCR systems

Process	GT capacity		HRSG gas-side		CCS capacity		CCS util. factor	
	[MW]		surface area		[kg CO ₂ /s]		[%]	
	min	max	min	max	min	max	min	max
	TCR	NPV	TCR	NPV	TCR	NPV	TCR	NPV
MEA	175	300	31.6	137.5	53.5	64.3	82.8	65.3
Piperazine	130	300	20.6	135.7	52.3	63.4	83.1	66.2
Mixed-salt base case	109	300	15.5	118.4	52.2	65.6	83.0	64.0
MS +10 K reb. appr. temp.	109	300	17.6	141.8	51.8	64.2	83.2	65.4
MS +25% spec. reb. duty	139	300	22.3	134.9	50.5	64.8	83.5	64.8
MS +50% spec. reb. duty	163	300	29.3	134.9	51.8	68.5	81.8	61.3
MS +25% spec. comp. duty	112	300	20.0	115.8	52.1	65.8	82.6	63.8
MS +50% spec. comp. duty	107	300	18.8	119.8	52.4	65.8	83.3	63.8

across all cases (including MEA) was \$236-252 million, and for minimum TCR facilities it was \$195-205 million. (Recall that, as stated in Section 2.4, we evaluate capital cost for all capture technologies using the same method and parameters as for MEA.) The CO₂ capture system capital cost for the maximum NPV baseline MEA facility was \$239 million, and for the minimum TCR it was \$205 million.

Consider the minimum TCR mixed-salt base case facility. This facility showed an improvement in TCR of \$48 million relative to the MEA baseline. Viewing this \$48 million reduction in TCR relative to the \$201 million capital cost of the mixed-salt base case CO₂ capture system, we conclude that mixed-salt is preferred over MEA even if it has a capital cost premium over MEA of up to 24%. Table 4 summarizes the capital cost premiums calculated in this manner for both of the advanced processes. The results indicate that the advanced processes can be preferred to MEA even with capital cost premiums over MEA of as much as 13-24% for the mixed-salt base case and 5-16% for piperazine, depending upon the choice of objective.

Note that the improvements in the TCR and NPV of these systems are less than the 30% capital cost uncertainty associated with our capital cost method. Errors in capital cost estimates are likely correlated, however, because the different processes use very similar equipment. Thus, differences of less than 30% in capital cost may indeed be significant.

4.1.2. System parameters

We now consider the specifications of the Pareto-efficient systems. Table 5 summarizes key system parameters for NPV-maximizing and TCR-minimizing facilities using each of the CO₂ capture processes. These parameters are visualized in Figure 7, which also shows the relative ranges observed in all Pareto-efficient systems (not only the minimum-TCR and maximum-NPV systems). Figure 7 also shows the relative values of key system parameters for the maximum NPV and minimum TCR facilities. Interestingly, the maximum NPV and minimum TCR facilities exhibit values that are close to the maximum or minimum of the range in each of these parameters, but they do not always take on the most extreme values.

Note that the ranges of system parameters such as CO₂ capture capacity of Pareto-efficient systems differ slightly for each of the capture processes, as seen in Figure 7. These differences are not related to capacity limitations specific to the capture processes, but arise naturally from the optimization results. The system parameters can take on any value allowed by the constraints summarized in Section 3 and Table 2, which are the same across all of the capture processes.

The reason the NPV-maximizing facilities using the advanced processes outperform the NPV-maximizing baseline MEA facility is that the advanced processes have lower duty requirements for reboiler heat and compression work. This results in increased net power output from the facility because more electricity can be generated from the steam cycle, and less electricity is used to compress the CO₂. Similarly, the lower energy duty requirements for the mixed-salt process compared to the piperazine process result in the NPV-maximizing mixed-salt facility outperforming the NPV-maximizing piperazine facility. Interestingly, the different energy duty requirements for the three processes do not appear to

Figure 7: Relative ranges of key parameters of Pareto-efficient systems employing the mixed-salt base case, piperazine, and MEA processes. Displayed values are scaled relative to the largest values observed for any Pareto-efficient system (excluding the mixed-salt sensitivity cases, which are not included in this figure), which are as follows: GT capacity, 300 MW; HRSG surface area, 137,500 m²; CCS capacity, 65.8 kg CO₂/s; CCS utilization, 83.6%.

have a large impact on time-varying operations, as can be seen in the similar ranges for CO₂ capture system utilization factor in Figure 7 for all cases considered.

The advanced process TCR-minimizing facilities outperform the TCR-minimizing MEA facility because the advanced processes have lower specific heat duty requirements. The smaller heat duty requirement entails a smaller CCGT system to capture the same amount of CO₂, thus reducing capital cost, even for CO₂ capture systems with similar capacities. Consider the first three rows of Table 5 (the last five rows concern the sensitivity cases, which are discussed in Section 4.2). We observe that the TCR-minimizing systems have CO₂ capture capacity in a narrow range, 52.2–53.5 kg CO₂/s, but the GT capacity of TCR-minimizing systems varies from 109 MW (mixed-salt base case) to 175 MW (MEA). Total HRSG gas-side surface area exhibits an analogous pattern, in which the TCR-minimizing surface areas are smaller for the mixed-salt base case (15,500 m²) and piperazine (20,600 m²) than for MEA (31,600 m²). The reduction in capital investment requirement arising from the use of the advanced processes in turn contributes toward the large differential in NPV observed in the TCR-minimizing designs for the different processes.

4.2. Sensitivity study on mixed-salt process

In this sensitivity assessment we vary the following three parameters for the mixed-salt process: specific reboiler heat duty, reboiler approach temperature, and specific compressor work duty. Figure 8 shows Pareto frontiers for this sensitivity study. Detailed specifications for the maximum NPV and minimum TCR systems are provided in Table 5. The dominant uncertainty in comparing the mixed-salt process with the piperazine and MEA processes is the specific reboiler heat duty. This can be seen in the relatively small effects on the objective functions from increases in reboiler approach temperature and specific compressor duty in Table 3 and Figure 8. Increasing the reboiler approach temperature by 10 K has a marginally negative impact on NPV for TCR-minimizing systems, and greater negative impact for NPV-maximizing systems. Nevertheless, the +10 K sensitivity case outperforms both piperazine and MEA regardless of choice of objective. Similarly, the +25% and +50% specific compressor work duty sensitivity cases outperform both piperazine and MEA along the entirety of the Pareto frontier.

Figure 8: Pareto frontiers for all processes including mixed-salt sensitivity study.

The +25% specific reboiler heat duty mixed-salt process sensitivity case outperforms piperazine along the entirety of the Pareto frontier. This is in part because the mixed salt process has a lower specific compression work duty than the piperazine process. This can be seen by comparing the mixed-salt +25% specific reboiler heat duty sensitivity case with the piperazine process. With reference to Table 1 and Section 2.1, both of these cases have similar total heat duty (2.75 MJ/kg CO₂ for the mixed-salt +25% reboiler duty sensitivity, and 2.6 MJ/kg CO₂ for piperazine), so both processes have similar minimum GT capacity and HRSG surface area (as seen in and Figure 7). The mixed-salt sensitivity case, however, has greater net power sales due to the lower compression work duty.

Further increasing the mixed-salt specific reboiler heat duty to +50% results in a Pareto frontier that still dominates the MEA Pareto frontier, but intersects the piperazine Pareto frontier. The mixed-salt sensitivity case outperforms piperazine for maximum NPV, while piperazine is preferred for minimum TCR. This indicates that, given a 50% increase in specific reboiler heat duty requirement, the choice of objective function between minimizing TCR and maximizing NPV would determine whether piperazine or mixed-salt is preferred. Regardless of the choice of objective, MEA is never the preferred technology in this assessment.

Taken together, the sensitivity study for this new process indicates that the mixed-salt process is very promising. Even with large (+25% and +50%) increases in operating duty requirements, the mixed-salt process is competitive with the piperazine process, and exhibits superior performance to the MEA process. This is true for both the minimum TCR and maximum NPV objective functions. Additionally, with reference to Table 4, we find that the mixed-salt process could have a capital cost premium of 5-24% and still be preferred over MEA, depending upon the sensitivity case and choice of objective function.

5. Concluding remarks

We have extended our previously developed heat integration optimization framework [14, 19] to represent two advanced CO₂ capture processes. The advanced processes, which are the mixed-salt process and the piperazine process, were employed within the heat integration framework alongside the baseline MEA process. Within this framework, computational optimization was used to determine the design and operations of a gas-powered CO₂ capture retrofit of a coal-fired power plant.

Results from bi-objective optimization for maximum net present value (NPV) and minimum total capital requirement (TCR) provide insight into the value of each of these technologies. These results indicate that, assuming equal capital cost scaling parameters (which is reasonable given the modeling level of this assessment), the piperazine and mixed-salt processes outperform the MEA process, and that the mixed-salt process in particular is quite promising. Facilities employing the base case mixed-salt process outperformed the MEA process by up to 16% in maximum NPV and 14% in minimum TCR. Using these results, we estimate that the base case mixed-salt process could have a capital cost premium of 13% (for a decision-maker concerned with maximizing NPV) to 24% (minimum TCR) above the capital cost of the MEA process and still be preferred over MEA.

A five-case sensitivity study on the mixed-salt process indicated that the mixed-salt outperformance of MEA was robust. The mixed-salt process also outperforms piperazine in the base case, but in two out of the five sensitivity cases considered (both involving increased specific reboiler heat duty), piperazine was competitive with or superior to the mixed-salt process. Recall that these results assume similar capital-cost scalings for the facilities, and therefore can be seen as benefits associated with the advanced processes if investment costs are consistent between designs.

The detailed quantitative results presented here are specific to the scenario considered in this study. However, the overall trends observed – notably, that the advanced processes both clearly outperform the MEA baseline, and that the mixed-salt process is highly competitive with the piperazine process – are likely to hold true for other scenarios because they arise from the fundamental physical performance of these systems. Additionally, we believe that the general range of the estimates calculated in this study for the value of the advanced processes (on order 5–24% improvement over MEA depending on the technology and choice of metric) is likely to be representative for scenarios and system configurations that are not radically different from those considered here.

Further research in this area should consider the detailed design of the piperazine and mixed-salt CO₂ capture processes within the optimization, along with more accurate representations of the capital costs for these processes. This could be accomplished using an approach employing proxy process models and bottom-up capital costing similar to that developed in Kang et al. [19]. Additionally, the capture processes could be assessed in different regulatory environments and for different plant configurations such as parasitic coal and parasitic gas-fired facilities.

Acknowledgments

Computational resources for this work were provided by the Stanford Center for Computational Earth and Environmental Science. The first author was supported by the Illich-Sadowsky Interdisciplinary Graduate Fellowship. The mixed-salt portion of this material is based upon the work supported by the US Department of Energy under award number DE-FE0012959.

Disclaimer

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

References

- [1] M. E. Boot-Handford, J. C. Abanades, E. J. Anthony, M. J. Blunt, S. Brandani, N. Mac Dowell, J. R. Fernandez, M.-C. Ferrari, R. Gross, J. P. Hallett, R. S. Haszeldine, P. Heptonstall, A. Lyngfelt, Z. Makuch, E. Mangano, R. T. J. Porter, M. Pourkashanian, G. T. Rochelle, N. Shah, J. G. Yao, P. S. Fennell, Carbon capture and storage update, *Energy & Environmental Science* 7 (2014) 130–189, doi:10.1039/C3EE42350F.

- [2] K. Goto, K. Yogo, T. Higashii, A review of efficiency penalty in a coal-fired power plant with post-combustion CO₂ capture, *Applied Energy* 111 (2013) 710–720, doi:10.1016/j.apenergy.2013.05.020.
- [3] G. Rochelle, E. Chen, S. Freeman, D. V. Wagener, Q. Xu, A. Voice, Aqueous piperazine as the new standard for CO₂ capture technology, *Chemical Engineering Journal* 171 (3) (2011) 725–733, doi:10.1016/j.cej.2011.02.011.
- [4] I. Jayaweera, P. Jayaweera, R. Elmore, J. Bao, S. Bhamidi, Update on mixed-salt technology development for CO₂ capture from post-combustion power stations, *Energy Procedia* 63 (0) (2014) 640–650, doi:10.1016/j.egypro.2014.11.070, 12th International Conference on Greenhouse Gas Control Technologies (GHGT-12), Oct. 6–9, 2014, Austin, TX, USA.
- [5] M. Rabensteiner, G. Kinger, M. Koller, G. Gronald, C. Hochenauer, Investigation of carbon dioxide capture with aqueous piperazine on a post combustion pilot plant – Part I: Energetic review of the process, *International Journal of Greenhouse Gas Control* 39 (2015) 79–90, doi:10.1016/j.ijggc.2015.05.003.
- [6] M. Rabensteiner, G. Kinger, M. Koller, G. Gronald, C. Hochenauer, Investigation of carbon dioxide capture with aqueous piperazine on a post combustion pilot plant – Part II: Parameter study and emission measurement, *International Journal of Greenhouse Gas Control* 37 (2015) 471–480, doi:10.1016/j.ijggc.2015.04.011.
- [7] E. Sanchez Fernandez, E. L. V. Goetheer, G. Manzolini, E. Macchi, S. Rezvani, T. J. H. Vlugt, Thermodynamic assessment of amine based CO₂ capture technologies in power plants based on European Benchmarking Task Force methodology, *Fuel* 129 (2014) 318–329, doi:10.1016/j.fuel.2014.03.042.
- [8] D. Hopkinson, D. Luebke, Z. Li, S. Chen, Solvent optimization of conventional absorption processes for CO₂ capture from postcombustion flue gases, *Industrial & Engineering Chemistry Research* 53 (17) (2014) 7149–7156, doi:10.1021/ie403869y.
- [9] M. Sharifzadeh, P. Bumb, N. Shah, Carbon capture from pulverized coal power plant (PCPP): Solvent performance comparison at an industrial scale, *Applied Energy* 163 (2016) 423–435, doi:10.1016/j.apenergy.2015.11.017.
- [10] S. M. Safdarnejad, J. D. Hedengren, L. L. Baxter, Plant-level dynamic optimization of Cryogenic Carbon Capture with conventional and renewable power sources, *Applied Energy* 149 (2015) 354–366, doi:10.1016/j.apenergy.2015.03.100.
- [11] P. Kumar, K.-H. Kim, Recent progress and innovation in carbon capture and storage using bioinspired materials, *Applied Energy* 172 (2016) 383–397, doi:10.1016/j.apenergy.2016.03.095.
- [12] S. M. Safdarnejad, J. D. Hedengren, L. L. Baxter, Dynamic optimization of a hybrid system of energy-storing cryogenic carbon capture and a baseline power generation unit, *Applied Energy* 172 (2016) 66–79, doi:10.1016/j.apenergy.2016.03.074.
- [13] X. Zhu, Y. Shi, N. Cai, Integrated gasification combined cycle with carbon dioxide capture by elevated temperature pressure swing adsorption, *Applied Energy* 176 (2016) 196–208, doi:10.1016/j.apenergy.2016.05.068.
- [14] C. A. Kang, A. R. Brandt, L. J. Durlofsky, Optimizing heat integration in a flexible coal-natural gas power station with CO₂ capture, *International Journal of Greenhouse Gas Control* 31 (2014) 138–152, doi:10.1016/j.ijggc.2014.09.019.
- [15] E. Martelli, L. O. Nord, O. Bolland, Design criteria and optimization of heat recovery steam cycles for integrated reforming combined cycles with CO₂ capture, *Applied Energy* 92 (2012) 255–268, doi:10.1016/j.apenergy.2011.10.043.
- [16] R. S. Middleton, J. K. Eccles, The complex future of CO₂ capture and storage: Variable electricity generation and fossil fuel power, *Applied Energy* 108 (2013) 66–73, doi:10.1016/j.apenergy.2013.02.065.
- [17] R. Khalilpour, Multi-level investment planning and scheduling under electricity and carbon market dynamics: Retrofit of a power plant with PCC (post-combustion carbon capture) processes, *Energy* 64 (2014) 172–186, doi:10.1016/j.energy.2013.10.086.
- [18] N. Abdul Manaf, A. Qadir, A. Abbas, Temporal multiscalar decision support framework for flexible operation of carbon capture plants targeting low-carbon management of power plant emissions, *Applied Energy* 169 (2016) 912–926, doi:10.1016/j.apenergy.2016.02.052.
- [19] C. A. Kang, A. R. Brandt, L. J. Durlofsky, A new carbon capture proxy model for optimizing the design and time-varying operation of a coal-natural gas power station, *International Journal of Greenhouse Gas Control* (2016) 234–252, doi:10.1016/j.ijggc.2015.11.023.
- [20] C. A. Kang, Computational Optimization of Design and Variable Operation of CO₂-Capture-Enabled Coal-Natural Gas Power Plants, Ph.D. thesis, Stanford University, URL <https://pangea.stanford.edu/ERE/pdf/pereports/PhD/Kang2015.pdf>, 2015.
- [21] P. G. Brodrick, C. A. Kang, A. R. Brandt, L. J. Durlofsky, Optimization of carbon-capture-enabled coal-gas-solar power generation, *Energy* 79 (2015) 149–162, ISSN 0360-5442, doi:10.1016/j.energy.2014.11.003.
- [22] C. A. Kang, A. R. Brandt, L. J. Durlofsky, Optimal operation of an integrated energy system including fossil fuel power generation, CO₂ capture and wind, *Energy* 36 (12) (2011) 6806–6820, doi:10.1016/j.energy.2011.10.015.
- [23] M. B. Berkenpas, K. Kietzke, H. Mantripragada, S. McCoy, E. S. Rubin, P. L. Versteeg, H. Zha, IECM Technical Documentation Updates Final Report, Tech. Rep. DE-AC26-04NT41917, National Energy Technology Laboratory, Carnegie Mellon University, Pittsburgh, PA, USA, URL <http://www.cmu.edu/epp/iecm/>, 2009.
- [24] M. S. Jassim, G. T. Rochelle, Innovative Absorber/Stripper Configurations for CO₂ Capture by Aqueous Monoethanolamine, *Industrial & Engineering Chemistry Research* 45 (8) (2006) 2465–2472, doi:10.1021/ie050547s.
- [25] E. Rubin, M. Berkenpas, C. Zaremsky, User Manual: Integrated Environmental Control Model, Carnegie Mellon University, Pittsburgh, PA, USA, URL <https://www.cmu.edu/epp/iecm/IECM.Publications/2007ra%20Rubin%20et%20al,%20IECM%20User.pdf>, 2007.
- [26] T. S. Kim, S. T. Ro, The effect of gas turbine coolant modulation on the part load performance of combined cycle plants. Part 2: Combined cycle plant, *Proceedings of the Institution of Mechanical Engineers, Part A: Journal of Power and Energy* 211 (6) (1997) 453–459, doi:10.1243/0957650981537348.
- [27] J. R. Couper, D. W. Hertz, F. L. Smith. In: D. W. Green and R. H. Perry, editors, *Perry's Chemical Engineers' Handbook*, chap. 9. Process Economics, McGraw-Hill, New York, 8th edn., 9–1–9–56, 2007.
- [28] K. M. Guthrie, Data and techniques for preliminary capital cost accounting, *Chemical Engineering* 76 (6) (1969) 114–142.
- [29] G. D. Ulrich, P. T. Vasudevan, *Chemical Engineering Process Design and Economics: A Practical Guide*, Process Publishing, Durham, NH, USA, 2nd edn., 2004.
- [30] H. Chalmers, M. Lucquiaud, J. Gibbins, M. Leach, Flexible operation of coal fired power plants with postcombustion capture of carbon dioxide, *Journal of Environmental Engineering* 135 (6) (2009) 449–458, doi:10.1061/(ASCE)EE.1943-7870.0000007.

- [31] R. Brasington, H. Herzog, Dynamic response of monoethanolamine (MEA) CO₂ capture units, in: Carbon Management Technology Conference, Feb. 7-9, 2012, Orlando, FL, USA, 151075-MS, doi:10.7122/151075-MS, 2012.
- [32] S. Ziaii, G. T. Rochelle, T. F. Edgar, Dynamic modeling to minimize energy use for CO₂ capture in power plants by aqueous monoethanolamine, *Industrial & Engineering Chemistry Research* 48 (13) (2009) 6105–6111, doi:10.1021/ie801385q.
- [33] O. J. Isebor, L. J. Durlofsky, D. Echeverria Ciaurri, A derivative-free methodology with local and global search for the constrained joint optimization of well locations and controls, *Computational Geosciences* 18 (3–4) (2014) 463–482, doi:10.1007/s10596-013-9383-x.
- [34] O. J. Isebor, L. J. Durlofsky, Biobjective optimization for general oil field development, *Journal of Petroleum Science and Engineering* 119 (2014) 123–138, doi:10.1016/j.petrol.2014.04.021.