Techno-economic study of enhanced absorber–regenerator configurations for improving an industrial Sulfinol-M-based acid gas removal processes

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A B S T R A C T

In this study, the rigorous simulation and design of an industrial Sulfinol-M-based acid gas removal (AGR) process was developed. All simulations were performed using Aspen HYSYS with extended NRTL activity coefficient model for the liquid phase and the Peng–Robinson state equation for the vapor fugacity coefficients. The simulation results from VMGSim and Aspen HYSYS were compared with design data from an integrated gasification combined cycle power plant. Several enhanced absorber–regenerator configurations were presented to improve AGR process efficiency. Sustainability analysis was conducted to assess these process alternatives in terms of energy requirements, costs, and environmental impacts. The results showed that the proposed self-heat recuperation configuration could reduce the energy requirements, total annual costs, carbon footprint by 74.7%, 40.1%, and 48.9%, respectively, compared with the conventional configuration.

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Introduction

Over the past several decades, coal has dominated worldwide electricity generation; it is used to produce approximately one-third of the world’s electricity [1]. Coal’s dominance is a result of its low cost, high energy density, and abundance relative to other energy resources. Coal-powered electricity will continue to play an important role in the next decades [1]. However, coal combustion is the source of over 70% of carbon dioxide (CO₂) emissions that arise from coal power plants, and more than 40% of global anthropomorphic CO₂ emissions [2]. To help coal become a cleaner energy source, the coal industry must use better technology to reduce coal’s carbon footprint and environmental impacts.

An integrated gasification combined cycle (IGCC) plant, which has the potential to produce low emissions, might be the answer to a sustainable energy future [2]. An IGCC power plant emits approximately 20% less CO₂ per unit of electricity generated compared to conventional pulverized coal combustion power plants [3]. By using excess heat in its steam cycle, the plant significantly increases its overall thermal efficiency. In an IGCC plant, a regenerative acid gas removal (AGR) process could be combined with a carbon dioxide capture process to remove both CO₂ and hydrogen sulfide (H₂S). Failing to adequately design either an AGR process or its operation will lead to a decrease in downstream performance, and will cause the plant emissions exceed regulated limits. Moreover, poor design will affect electricity generation in the power block because of deposition on and corrosion of turbine blades. Hence, the safe and sustained operation of an IGCC power plant requires a reliable AGR process.

AGR processes normally use three types of solvents including physical solvents, chemical solvents, and mixed solvents to remove acid gas from syngas. Of these, a mixed solvent, which is a mixture of chemical and physical solvents, can retain the benefits of both solvent types, and handle a wide range of acid gas partial pressures. For removing acid gas in an IGCC power plant, the mixed solvent Sulfinol, which comprises an aqueous amine (chemical solvent) and a sulfonate (physical solvent) is one of the most reliable and well-proven solvents [4]. In particular, Sulfinol-M, which is a mixture of sulfonate, MDEA, and water, is widely used for natural gas or coal-based syngas because it can selectively absorb H₂S, COS, and mercaptans, while at the same time partially absorbing CO₂ [4].

Several reported studies related to AGR processes have covered thermodynamic models, process design, and simulation. Deshmukh and Mather presented a thermodynamic model for acid gas removal...
solubility in aqueous alkanolamine solutions [5]. The Deshmukh and Mather model was further developed by Macgregor and Mather to predict acid gas solubility in mixed solvents [6] while Najibi and Maleki developed a model to predict CO₂ solubility in MDEA and piperazine solutions [7]. Other researchers investigated acid gas solubility in a mixed solvent system using an electrolyte non-random two-liquid (NRTL) model [8–10].

The models in the studies described above have only been compared with experimental data at the lab scale. Validation with plant data would be required to achieve more reliable results for industrial-scale operation. An absorber column from an industrial Rectisol-based AGR process was simulated and validated with plant data from Shanghai Coking & Chemical Corporation by Liu et al. [11], while Nayak simulated and evaluated a Selexol-based process [12]. In addition, Liu et al. proposed a configuration that can save around 0.5GJ/tCO₂ compared to a conventional configuration [11]. Dyment and Watanasiri presented an AGR process using an MDEA solvent simulated using an acid gas property package based on the electrolyte non-random two-liquid (NRTL) model in Aspen HYSYS [10]. The simulation results were then compared with experimental and plant data. To the best of our knowledge, there is no study demonstrating the detailed simulation and design of an AGR process using a Sulfinol-M solvent at an industrial scale, in which the design results were validated with plant data.

In this paper, an industrial AGR process using a Sulfinol-M solvent was designed and simulated rigorously by using Aspen HYSYS simulator. To test the reliability of model, the simulation results were compared with the results from several commercial simulators, as well as data from an IGCC power plant in Korea. Subsequently, several absorber–regenerator configuration alternatives for enhancing either absorber or regenerator performance were proposed to improve overall process efficiency. For a fair comparison, techno-economic analyses were conducted to assess these configurations in terms of their energy requirements, economic performance and environmental impacts.

Simulation methods

In industry, energy efficiency improvements lead to greater profit, because they reduce operating costs. In an AGR process, energy efficiency improvements are achieved with energy savings in the reboiler. This can be achieved by several methods: (1) using solvents with lower reaction heat or (2) improving operating conditions by using new process configurations. The second method is more applicable than the first method because it involves only minor changes in plant layout. This means that the second one can be easily implemented in an existing plant.

In this study, several process configurations were evaluated using Aspen HYSYS V8.6. A Sulfinol-M based AGR process is a highly non-ideal system that requires simultaneous solution of vapor–liquid equilibrium (VLE) and vapor–liquid–liquid equilibrium (VLLE) equations. The system also has a wide boiling point range concentration range between components. Accordingly, the extended NRTL package of Aspen HYSYS which allows to input values for the Aij, Bij, Cij, Alp1ij, and Alp2ij energy parameters was used for simulation. Aspen HYSYS uses the following equation form for τij and αij

\[ \tau_{ij} = (A_{ij} + B_{ij}t + C_{ij}t^2) / T; \alpha_{ij} = Alp_{1ij} + Alp_{2ij} \]  

where T is in K, and t is in °C. The energy parameters are listed in detail in the Supplementary material.

The COSTALD and Rackett methods are recommended for petroleum and hydrocarbon liquid mixtures to overcome the drawback of poor liquid density calculated by the cubic equation of state [13]. To achieve accurate simulation, the AGR process was divided into two parts: absorption and regeneration. The Rackett density method (based on API – American Petroleum Institute), which incorporates temperature dependence and is more suitable at high pressures, was used for absorption. Meanwhile, the COSTALD density method, which includes both temperature and pressure dependences and gives slightly better results at low pressures, was used for regeneration [13]. In addition, the Peng–Robinson method was used as a vapor model with Poynting correction, which uses each component’s molar volume (liquid phase) to calculate the overall compressibility factor.

Enhanced process configurations

Conventional Sulfinol-M (CS) configuration

Generally, IGCC technology uses a high-pressure gasifier to turn coal and other carbon-based fuels into pressurized gas (syngas). Fig. 1 presents a schematic representation of the IGCC power plant in Korea. In this study, we focused on the AGR unit, which consumes the greatest amount of energy in the syngas treatment process. The conventional process that was modeled was an industrial Sulfinol-M-based process for a 345,000 metric ton/year gasified coal plant. First, the CS was modeled and simulated to verify the thermodynamic model setup. A simplified flowsheet of the CS process with key design parameters and operating conditions is shown in Fig. 2. Essentially, the AGR process has an absorber, which requires high pressure and low temperature, and a regenerator, which requires high temperature and low pressure. The syngas feed considered in the model was obtained from HCN-COS hydrolysis unit design data for the IGCC plant and contained a small fraction of acid gases (0.26 wt% H₂S and 1.34 wt% CO₂) that needed to be treated. The gasifier considered in this study was the Shell® entrained flow type operating at temperature of 1500 °C, pressure of 42 barg, steam/coal mass ratio of 0.05, oxygen/coal mass ratio of 0.75, and nitrogen/coal mass ratio of 0.112 [14,15]. Table 1 lists details of the syngas feed components and conditions. First, the syngas feed and Sulfinol-M are introduced into the absorber to generate both a rich solvent containing a high acid-gas content and a treated gas, which is delivered to the knockout (KO) stages.

![Fig. 1. Schematic diagram of the IGCC power plant.](image-url)
vessel before leaving the AGR process. Because it is a mixed solvent, Sulfinol-M can absorb acid gasses both physically and chemically. The rich solvent absorbed 99.7 wt% of H2S and 44.9 wt% of CO2 in the syngas. Subsequently, the rich solvent was input into the regenerator to release the acid gases. In contrast to the absorber, the regenerator operates at high temperature (~120 °C) and low pressure (~1.8 bar) to improve acid gas desorption. The acid gases then flow to a reflux accumulator before being delivered to the sulfur recovery unit. The lean solvent, which has a low content of acid gases, is collected at the bottom of the sulfur recovery unit and is sent to a heat exchanger, where it exchanges heat with the regenerator feed in order to reduce energy requirements in reboiler. Afterward, the lean solvent is cooled to 45 °C by a cooler and is introduced back into the absorber. Note that a small amount of solvent must be added at this point to account for the small losses of solvent in the treated gas and acid gas streams. Key design parameters for simulating the CS absorber and regenerator are listed in Table 2. To verify the Aspen HYSYS simulation, the calculated results for the treated gas and acid gas streams were compared with both IGCC power plant design data and VMGSim simulation results. As shown in Table 3, the VMGSim produced a good prediction for the absorber, but its results did not fit the design data for the regenerator. The Aspen HYSYS results, however, were in good agreement with the plant design data for both the absorber and regenerator; therefore, the model setup was reliable. In the following sections, several alternative configurations are presented for improving the energy and cost efficiency of the conventional process. Note that the heat exchanger in the CS process was designed by specifying an overall heat transfer coefficient UA of 106 kJ/C·h, which corresponds to the plant design data. New heat exchangers used in the following enhanced configurations are designed to achieve a minimum approach temperature of 10°C.

### Absorber intercooling (AI) configuration

Acid gas loading is the key factor in improving absorber performance. Several studies have proposed an absorber intercooling configuration as an efficient way of increasing acid gas loading [16,17]. In particular, the absorption of acid gasses is favorable at a low temperature. Based on the temperature profile along the absorber, a side stream can be withdrawn, cooled, and returned to the absorber. Cooling the internal liquid can remove the heat released from the chemical reaction between acid gasses and MDEA, resulting in a smooth temperature profile in the absorber and improved acid gas absorption. Accordingly, the absorber requires less solvent, and its size can be decreased. Finding an optimal side-stream location is important for maximizing solvent capacity. Fig. 3 shows details of the temperature and acid gas profiles along the absorber for the conventional process. From these profiles, the 8th stage which has the highest

### Table 1

<table>
<thead>
<tr>
<th>Component</th>
<th>Mass fraction (wt%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>H2O</td>
<td>0.27</td>
</tr>
<tr>
<td>HCN</td>
<td>1.24 ppm</td>
</tr>
<tr>
<td>CO2</td>
<td>2.71</td>
</tr>
<tr>
<td>H2S</td>
<td>0.40</td>
</tr>
<tr>
<td>CO</td>
<td>80.64</td>
</tr>
<tr>
<td>N2</td>
<td>12.02</td>
</tr>
<tr>
<td>Ar</td>
<td>1.60</td>
</tr>
<tr>
<td>H2</td>
<td>2.35</td>
</tr>
<tr>
<td>NH3</td>
<td>15.67 ppm</td>
</tr>
<tr>
<td>CH4</td>
<td>0.01</td>
</tr>
<tr>
<td>Temperature (°C)</td>
<td>45</td>
</tr>
<tr>
<td>Pressure (bar)</td>
<td>37.8</td>
</tr>
<tr>
<td>Mass flowrate (kg/h)</td>
<td>198,936</td>
</tr>
</tbody>
</table>
temperature was determined to be the optimal side-stream location to be feed into a cooler, where it is cooled to 42 °C. Subsequently, the side-stream is recycled back into the absorber. Fig. 4 presents a simplified diagram and key design parameters for the AI configuration. For a fair comparison, the amount of absorbed CO$_2$ and H$_2$S was kept at the same level as in the treated gas in the CS process, leading to a 1.5% decrease in the solvent flow rate required for the AI process (117,018 kg/h) compared to the conventional process (118,800 kg/h). Furthermore, the lower amount of required solvent was directly proportional to lower energy requirements in the regenerator, with a 4.1% reduction relative to the conventional process. Note that the term “energy requirements” in the present study means the energy requirements in reboiler duty of the regenerator.
Split solvent flow (SSF) configuration

In the conventional configuration, the lean solvent, which is collected at the regenerator bottom, is recycled back into the absorber. To increase regenerator performance, Shoeld originally proposed an SSF [18] in which the semi-lean solvent is withdrawn from an intermediate stage of the regenerator and recycled to a middle tray of the absorber. This configuration changes the profiles of the regenerator’s top section and reduces liquid flow at the regenerator bottom, resulting in a lower regenerator reboiler duty. In the present process, a portion of the internal solvent liquid can be withdrawn from the 9th stage of the regenerator, which subsequently preheats the rich solvent stream for a more efficient heat recovery. The semi-lean solvent is then cooled to 45 °C before being introduced to the absorber in the 8th stage. Inputting the cooled solvent into the middle of the absorber can overcome the absorber’s characteristic temperature profile, as in the AI configuration. As a result, heat released from the reaction between acid gasses and MDEA can be removed, leading to a smooth temperature profile and increased CO₂ absorption. For a fair comparison, the treated gas in the SSF configuration was kept the same as in the CS configuration. The semi-lean solvent must comprise a sufficiently low content of acid gasses (1.89E-11 kg/h CO₂ and 5.68E-8 H₂S) so that the performance of the absorber is not affected. The required solvent flowrate therefore decreased to 114,170 kg/h, resulting in a 3.8 wt% reduction relative to the stream in the CS configuration (118,740 kg/h). Fig. 5 presents a simplified flowsheet for the SSF process, with key design parameters and detailed operating conditions. The results show that the SSF configuration can reduce the system’s required energy by 2.5% relative to the CS configuration, taking into account the lower vapor flow in the regenerator reboiler. However, the total circulation liquid flow was increased, and greater investment cost might be required in order to install more than one cooler and heat exchanger. These drawbacks reduce the benefits of the SSF configuration, which has consequently not found many applications in industry [19].

Rich-solvent splitting and preheating (RSP) configuration

The thermal condition of a feed is one of the key factors affecting that feed’s energy efficiency in a distillation column. Only a fraction of the thermal energy given to the feed stream by heat exchange with other heat resources can reduce the energy requirement inputs from reboiler [20]. In the CS process, although the regenerator feed is preheated from 47 to 104 °C by the lean solvent stream, the temperature of the top vapor is still high (109 °C). To recover these heat, the rich solvent is split into two streams: one is preheated by the lean solvent, and one exchanges heat with the top vapor of the regenerator. Finding the optimal rich-solvent split ratio is important for maximizing heat recovery and feed preheating efficiency. Accordingly, several simulation runs with different split ratios were performed in order to examine the effect of split ratio on the reboiler duty of the regenerator. It is worth noting that the lean solvent (118,770 kg/h) is much higher than that of the top vapor (7338 kg/h). Therefore, the split ratio of 7:3 achieved a maximum decrease in the reboiler duty (4310 kW) as shown in Fig. 6. The regenerator feed can be preheated up to 110 °C by a simple splitting, which corresponds an increase of 6 °C compared to the temperature of that stream in the conventional process. Fig. 7 presents a simplified flowsheet with key design parameters for the RSP process configuration. The results show that the RSP configuration can reduce energy requirements by 15.7% relative to the conventional process.

Mechanical vapor recompression (MVR) configuration

A distillation column with high energy and high temperatures in the reboiler has low thermodynamic efficiency because the similar amount of energy is released at low temperatures in the condenser [21]. As a consequence, a heat pump that can transfer heat from the top vapor to the reboiler can be integrated into a distillation column to improve its energy efficiency [22]. Fig. 8 shows an MVR heat pump integrated into the stripper of a sulfonil-M process. The top vapor of the regenerator is compressed from

![Fig. 5. Schematic diagram of the SSF configuration.](image-url)
1.7 bar to 3.7 bar, and then exchanges heat with the bottom stream. Note that the compressor ratio was designed to keep the temperature of the outlet vapor below 200 °C, in order to avoid compressor overheating. Compressor overheating is the industry’s most serious compressor problem that can currently be avoided with this configuration. In addition, the minimum temperature approach of 10 °C was chosen for the heat exchanger. The compressor outlet provides a heating effect when it is in the process of condensing, and its latent heat is transferred to the bottom stream. As a result, less energy is required in the reboiler. The simulation results show that the MVR configuration can reduce energy requirements by 20.1% relative to the conventional process.

**Lean Sulfinol-M flash (LSF) configuration**

In the regenerator, reboiling the bottom stream at a high operating pressure requires intensive energy inputs. However, the AGR unit in this study is one part of whole IGCC power plant, in which the pressure of acid gas outlet from the reflux accumulator must be at least 1.5 bar for further use in the next syngas saturator unit. Therefore, the operating pressure of stripper (1.7 bar) could not be reduced further. By using a valve, the bottom stream can be depressurized down to atmospheric pressure before being input into the reboiler. The boiling temperature of the bottom liquid is reduced accordingly, resulting in lower energy inputs to the reboiler. Fig. 9 illustrates a simplified flowsheet for the LSF configuration, with main design parameters. In particular, the regenerator bottom is depressurized from 1.9 bar to 1.0 bar before being reboiled, resulting in lower energy requirements. The vapor from the reboiler is pressurized to 1.9 bar before being introduced back into the regenerator, while the liquid from the reboiler is recycled back into the absorber. Note that the temperature of those two inputs is 106 °C, a decrease of 19 °C relative to the reboiler outlets in the conventional process. This lower lean solvent temperature results in a lower temperature for the regenerator feed, because the regenerator feed is preheated by the lean solvent. The simulation results show that the LSF configuration can reduce energy requirements by up to 23.7% relative to the CS process.

**Proposed self-heat recuperation (PS) configuration**

To maximize the energy efficiency of the Sulfinol-M-based process, it is possible to apply several techniques simultaneously, as shown in Fig. 10. Integrated AI improved the solvent working capacity, thus reducing the amount of required solvent without reducing gas treatment capacity, leading to a reduction in total energy requirements. In the PS configuration, self-heat recuperation (SHR) technology, which is a combination of MVR and feed preheating, was applied. In the MVR process, the top vapor is compressed. This high-pressure vapor then supplies heat to the bottom stream during condensation, where only the latent heat is utilized [23]. On the other hand, both latent and sensible heat are utilized when applying SHR technology [23,24]. Overhead vapor is compressed, and its partial heat is recovered from a heat exchange with the bottom stream. Next, the compressed vapor supplies heat to the feed preheater, leading to the recovery of its remaining heat. The bottom stream also exchanges heat with the feed stream. Therefore, all the heat in the process can be effectively recirculated. In addition, the bottom stream pressure of the regenerator was expanded to 1.01 bar; thus, the temperature of the lean Sulfinol-M solvent in the proposed process (106 °C) is lower than in the conventional process (125 °C). As a result, compared to the above
RSP configuration, the feed input to the regenerator does not need to be split. The results show that, relative to the conventional process, the proposed process can reduce energy requirements by up to 51.0% and 60.8% in the condensers (including coolers) and reboiler, respectively.

**Economic evaluation**

In addition to the comparison of energy requirements in the regenerator reboiler, the total investment cost (TIC), total annual operating cost (TOC), and total annual cost (TAC) of all configurations were calculated to cope with different energy and capital costs. Equipment cost was estimated using correlations from the Biegler textbook [25], in which the Chemical Engineering Plant Cost Index of 575.5 for 2014 was used to account for inflation. TICs were obtained based on equipment sizes and capacities. The analysis considered major equipment such as compressors, heat exchangers, vessels, and columns (including auxiliary equipment such as reboilers, condensers, and tray stacks). The utility prices including cooling water, steam, and electricity used for TOC estimation are listed in Table 4 [26]. Note that beside the TAC, maintenance cost may increase largely when a design alternative includes complicate devices such as a compressor. However, the TAC, which describes the optimal trade-off between TIC and TOC, constitutes a suitable performance measure to evaluate various design alternatives in a preliminary design stage where an order-of-magnitude estimate of 25–40% error level is normally acceptable [25]. In this study, TAC was calculated as follows [26]:

$$TAC = TIC \times \left[ \frac{i(1+i)^n}{(1+i)^n-1} \right] + TOC$$  \hspace{1cm} (2)
where \( i \) (0.08) is the fractional interest rate per year and \( n \) (10 years) is the plant lifetime. In this study, efforts are focused on reducing energy requirements while achieving specific values of \( \text{H}_2\text{S} \) and \( \text{CO}_2 \) absorption, which affect the composition of the treated gas at the outlet.

Table 5 presents key results for all configurations, including energy requirements and costs, while Fig. 11 conveniently illustrates comparisons of TIC, TOC, and TAC. Compressors were used in several processes to upgrade low-quality energy in the condenser (sensible heat). It acts as a heat pump that moves heat energy from the top column to the reboiler. Compressor utilization has a clear positive effect reboiler duty for MVR, LSF, and PS processes, leading significant TOC decreases for these processes of 13.1%, 13.2%, and 54.5%, respectively. In addition, the proposed process requires the lowest reboiler and condenser duty, and the highest compressor duty, which takes into account the fact that more sensible heat is pumped and upgraded to the column relative to MVR and LSF processes. Compressors, however, are expensive, and their use raises investment costs considerably. The costs of MVR, LSF, and PS processes increase by 23.5%, 23.7%, and 37.3%, respectively, relative to the CS process. Clearly, compressor use affects investment cost, and has a trade-off effect with operating cost. AI, SSF, RSP, MVR, LSF, and PS configurations can reduce TACs by 3.3%, 1.8%, 12.4%, 5.6%, 5.7%, and 40.1% relative to conventional processes. Obviously, the PS process is superior to the others in terms of its economic benefits. Furthermore, the simple payback period calculated as follows was 18.8 months, making the

Table 5

<table>
<thead>
<tr>
<th>Structural alternative</th>
<th>CS</th>
<th>AI</th>
<th>SSF</th>
<th>RSP</th>
<th>MVR</th>
<th>LSF</th>
<th>PS</th>
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<tr>
<td>Condenser duty (kW)</td>
<td>5032</td>
<td>4683</td>
<td>4920</td>
<td>4235</td>
<td>4350</td>
<td>4126</td>
<td>1793</td>
</tr>
<tr>
<td>Condenser duty savings</td>
<td>–</td>
<td>3.4%</td>
<td>2.2%</td>
<td>15.8%</td>
<td>13.6%</td>
<td>18.0%</td>
<td>64.4%</td>
</tr>
<tr>
<td>Reboiler duty (kW)</td>
<td>5113</td>
<td>4901</td>
<td>4987</td>
<td>4310</td>
<td>4085</td>
<td>3901</td>
<td>1296</td>
</tr>
<tr>
<td>Reboiler duty savings</td>
<td>–</td>
<td>4.1%</td>
<td>2.5%</td>
<td>15.7%</td>
<td>20.1%</td>
<td>23.7%</td>
<td>74.7%</td>
</tr>
<tr>
<td>Total investment costs (US k$)</td>
<td>3616</td>
<td>3625</td>
<td>3638</td>
<td>3643</td>
<td>4465</td>
<td>4475</td>
<td>4967</td>
</tr>
<tr>
<td>Total annual investment costs savings</td>
<td>–</td>
<td>–0.2%</td>
<td>–0.6%</td>
<td>–0.7%</td>
<td>–23.5%</td>
<td>–23.7%</td>
<td>–37.3%</td>
</tr>
<tr>
<td>Total annual operating costs (US k$)</td>
<td>2111</td>
<td>2023</td>
<td>2059</td>
<td>1779</td>
<td>1835</td>
<td>1832</td>
<td>961</td>
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<tr>
<td>Total annual operating cost savings</td>
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<td>4.1%</td>
<td>2.5%</td>
<td>15.7%</td>
<td>13.1%</td>
<td>13.2%</td>
<td>54.5%</td>
</tr>
<tr>
<td>Total annual cost savings</td>
<td>–</td>
<td>3.3%</td>
<td>1.8%</td>
<td>12.4%</td>
<td>5.6%</td>
<td>5.7%</td>
<td>40.1%</td>
</tr>
</tbody>
</table>
proposed configuration an attractive option for industrial implementation.

Payback period = Cost of project/Saving per year (3)

Environmental analysis

Carbon dioxide emissions are clearly linked to energy requirements. However, this pattern does not always hold true when different energy sources are utilized for energy-saving purposes [27]. In this study, the environmental impacts of AGR processes compared. In particular, CO₂ emissions generated from steam reboilers and compressors were predicted using a model developed by Gadalla et al. [28]. In the real world, both fossil fuels and renewable resources are used to produce electricity. To make a fair comparison, however, electricity in the model was assumed to be produced from natural gas (similar to steam generation). The carbon footprint of steam reboilers was calculated as follows [28]:

$$\text{CO}_2\text{emissions} = \left(\frac{Q_{\text{fuel}}}{\text{NHW}}\right) \cdot \left(\frac{C\%}{100}\right)$$

(4)

where NHV is the net heating value of fuel and C% is the carbon content. For natural gas, NHV is 48,900 kJ/kg and C% is 0.41 kg/kg. The molar mass ratio of CO₂ and C is α = 3.67. Q_{\text{fuel}} represents the amount of fuel used, which was calculated as follows:

$$Q_{\text{fuel}} = \left(\frac{Q_{\text{proc}}}{\lambda_{\text{proc}}}\right) \cdot \left(\frac{h_{\text{proc}} - 419}{T_{\text{FI}B} - T_{\text{FI}B}}\right)$$

(5)

where Q_{\text{proc}} is the required heat duty of the system, and λ_{\text{proc}} (kJ/kg) and h_{\text{proc}} (kJ/kg) are the latent heat and enthalpy of the steam, respectively. The flame temperature, stack temperature, and ambient temperature were T_{\text{FI}B} (1800 °C), T_{\text{FI}B} (160 °C) and T_{\text{FI}} (25 °C), respectively. For a compressor, Q_{\text{fuel}} can be calculated from the efficiency of a gas turbine (η_{\text{GT}}) and the Carnot factor (η_{\text{C}}) [28]:

$$Q_{\text{fuel}} = \left(\frac{Q_{\text{proc}}}{\eta_{\text{GT}}}\right) \cdot \left(\frac{1}{1 - \eta_{\text{C}}}\right)$$

(6)

Table 6 lists the results for the total annual CO₂ emissions (TCE) of all Sulfinol-M-based processes, while Fig. 11 provides a convenient visualization. Clearly, given similar energy requirements, a compressor has a larger carbon footprint than a reboiler. For instance, the CO₂ emissions are quite similar from steam reboilers and compressors for the proposed process. However, the energy required by the reboilers is almost twice that of the compressors. This is because electricity efficiency is normally lower than heat efficiency. Therefore, a heat pump system consisting of compressors can reduce energy requirements but might produce more CO₂ emissions. It can be seen that the MVR and LSF configurations can (respectively) reduce reboiler duty energy by 13.6% and 18%, but can only reduce CO₂ emissions by 5.6% and 7.7%. In addition, the results show that AI, SSF, and RSP processes can reduce CO₂ emissions by 4.1%, 2.5%, and 15.7%, respectively relative to a conventional process. As energy requirements are linked closely to CO₂ emissions, the proposed process is the most environmentally friendly process, which can reduce the carbon footprint by up to 48.9% relative to a conventional process.

Furthermore, the proposed configuration can be applied in both retrofits and new plant configurations. In a retrofit of an existing process, new heat exchangers and compressors can be installed, while the number of trays and column diameters stay the same. Additional pipe work will also be needed to link equipment. Technically, these studies can be easily accomplished, with short modification time.

Conclusions

An industrial AGR process using Sulfinol-M as a solvent in an IGCC power plant was simulated and designed to be more energy-efficient. All simulations were performed using Aspen HYSYS. The Extended NRTL activity coefficient model for the liquid phase and the Peng–Robinson state equation for the vapor fugacity coefficients have proven to be reliable for predicting properties of all components in the AGR process. The simulation results from Aspen HYSYS were more accurate than those from other simulators and agreed well with data from an IGCC power plant design in Korea. To the best of our knowledge, this is the first study in which simulated results were validated with plant data for a simulation of an industrial AGR process using Sulfinol-M as a solvent.

Several enhanced process configurations were studied for their ability to improve the energy efficiency of a conventional AGR process. These configurations included AI, SSF, RSP, MVR, LSF, and PS. The results of a techno-economic analysis confirmed these configurations’ potential benefits for improving energy and economic performance and reducing carbon emissions. Among those configurations, the PS produced the most remarkable improvements: the configuration reduced the system’s energy requirements, TACs, and CO₂ emissions by 74.7%, 40.1%, and 48.9%, respectively, relative to a conventional Sulfinol-M-based AGR process. The proposed configurations are not only promising in retrofit cases with short implementation time but also in grass-root cases for other AGR processes.

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Appendix A. Supplementary data

Supplementary data associated with this article can be found, in the online version, at http://dx.doi.org/10.1016/j.jiec.2017.06.028.

References