Optimal plant-wide control of the wet sulfuric acid process in an integrated gasification combined cycle power plant

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1. Introduction

The wet sulfuric acid (WSA) process is a sulfur treatment technology that was patented by Haldor Topsøe in the late 1980s. This process is being increasingly used as emission regulations continue to get tighter, which force industries to properly treat their sulfurous off-gas. The WSA process is not new; in fact, it has been used widely in many industries where sulfur removal is an issue. One of these is in integrated gasification combined cycle (IGCC) power plants.

Each unit of the WSA process requires particular conditions for sound operation. For instance, to safely and effectively conduct the combustion reaction, a sufficient amount of air must be provided. The SO2 conversion reaction produces SO3 at an optimum rate when the temperature of each bed in the reactor is at a particular value. In addition, optimum condensation of H2SO4 gas also occurs at a certain temperature. Therefore, an advanced control structure of the WSA process is essential to fulfill the requirements of each unit.

Despite the maturity of the WSA process, there are no reports that discuss the plant-wide control structure synthesis of this process. Almirall [1] investigated the optimization of wet sulfuric acid plants using exergoeconomics and Robinson [2] discussed the plant-wide control of hybrid IGCC/methanol plants but did not include WSA process in his discussion.

This work discusses plant-wide control system design of the WSA process for optimal operation. The first step was to list the control objective of every key unit in the WSA process. Generally, the control objective of each unit is to attain optimal operation while preserving the safety and stability of the process. The second step was to address the requirements and constraints demanded by each unit in order to ensure safe and stable operation. These requirements and constraints aided in the selection of proper controlled variables for achieving the predefined controlled objectives. Degrees of freedom analysis was subsequently performed to locate the available adjustable variables to maintain the set point of the selected controlled variables.

A process achieves its optimal operation when each of its control loops has an optimum set point [3]. In the studied WSA process, the energy consumption of the hot air blower and the production rate of H2SO4 were the two variables that reflected the operational optimality of the WSA process. Within the process, there are several controlled variables whose set point values directly affect the blower duty and H2SO4 production rate i.e. the temperature of the combustion reaction, and the temperature of the inlet to each bed

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1 Husnil and Andika were contributing equally in this work and are considered as co-first authors.
in the SO₂ converter and to the WSA condenser. The set points of these variables were step tested at three different feed flow rate scenarios to observe their influence on the blower duty and H₂SO₄ production rate. The optimum set point of these variables was not similar for each flow rate case. Therefore, they were selected by considering the feasible and safe operation of the WSA process at the predefined range of the feed flow rate while at the same time keeping the loss at an acceptable level. To the best of the authors' knowledge, this paper is the first published work that examines and discusses WSA process control, even considering its application in industry for the last 30 years.

2. Wet sulfuric acid process

The WSA process is one of the downstream units in the gasification block of an IGCC power plant where sulfurous gases from the upstream are processed to produce more valuable products. The rigorous dynamic model of the WSA process in this work was developed in the Aspen Hysys Dynamic environment and was used for conducting all supporting analyses as well as verifying the performance of the designed control system.

2.1. Process description

The feed to the WSA process in this work came from three different units in the IGCC power plant i.e. acid gas removal, sour water stripper (two feed streams), and sour slurry handling unit (Fig. 1). These four streams carry various components as side products from the coal processing in the upstream units. The sulfur component in these feeds is in a form of H₂S gas where the highest composition of this component is carried by the stream from the acid gas removal unit. The operating conditions of the feed streams to the WSA unit are shown in Fig. 1 while the composition of the feed streams is listed in Table 1.

The main objective of the WSA process is to handle the sulfur content in the flue gas from the IGCC power plant so that it does not exceed the permitted level. This objective is then translated into a more economically beneficial target i.e. converting the sulfur con-

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Fig. 1. Conceptual diagram of the WSA process as part of an IGCC power plant.
Since only tent streams, Table streams minimum SO2 converter, is in condenser, coordination conceptually To the temperature increase reactions the gas stream, are feed the unit. temperature of the stream, mixed of the unit. temperature of the reaction, temperature of the conversion, is to the atmosphere. Table 1 Composition of the feed streams to the WSA unit.

<table>
<thead>
<tr>
<th>Component</th>
<th>Acid Gas</th>
<th>SSS Gas</th>
<th>SS Gas</th>
<th>SWS Off Gas</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
<td>0.0170</td>
<td>0.0450</td>
<td>0.4664</td>
<td>0.0018</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.6250</td>
<td>0.3000</td>
<td>0.2074</td>
<td>0.3069</td>
</tr>
<tr>
<td>H₂</td>
<td>0.0066</td>
<td>0.0190</td>
<td>0.1888</td>
<td>0.0007</td>
</tr>
<tr>
<td>H₂S</td>
<td>0.2860</td>
<td>0.0720</td>
<td>0.0429</td>
<td>0.0710</td>
</tr>
<tr>
<td>HCN</td>
<td>–</td>
<td>0.0230</td>
<td>–</td>
<td>0.0002</td>
</tr>
<tr>
<td>N₂</td>
<td>0.0024</td>
<td>0.0520</td>
<td>0.0545</td>
<td>0.0001</td>
</tr>
<tr>
<td>NH₃</td>
<td>–</td>
<td>0.0610</td>
<td>–</td>
<td>0.3422</td>
</tr>
<tr>
<td>H₂O</td>
<td>0.0630</td>
<td>0.3880</td>
<td>0.0400</td>
<td>0.2771</td>
</tr>
</tbody>
</table>

Fig. 2. Typical adiabatic conversion process (reconstructed from Louie [5]).

Table 2 List of reactions in WSA process.

<table>
<thead>
<tr>
<th>Unit</th>
<th>Reaction</th>
</tr>
</thead>
<tbody>
<tr>
<td>Combustor</td>
<td>CO + 0.5 O₂ → CO₂</td>
</tr>
<tr>
<td></td>
<td>H₂ + 0.5 O₂ → H₂O</td>
</tr>
<tr>
<td></td>
<td>H₂S + 1.5 O₂ → SO₂ + H₂O</td>
</tr>
<tr>
<td></td>
<td>2 HCN + 2.5 O₂ → 2 CO₂ + H₂O + N₂</td>
</tr>
<tr>
<td></td>
<td>2 NH₃ + 1.5 O₂ → 3H₂O + N₂</td>
</tr>
<tr>
<td>SO₂ Converter</td>
<td>SO₂ + 0.5 O₂ → SO₃ + 99 kJ/mole</td>
</tr>
<tr>
<td>Condenser</td>
<td>SO₃ + H₂O → H₂SO₄(g) + 101 kJ/mole</td>
</tr>
<tr>
<td></td>
<td>H₂SO₄(g) + 0.17H₂O(g) → H₃SO₄(l) + 69 kJ/mole</td>
</tr>
</tbody>
</table>

tent into H₂SO₄. This H₂SO₄ production is accomplished through coordination of four key process units i.e. combustor, SO₂ converter, condenser, and steam drum. The flow diagram of the WSA process is conceptually illustrated in Fig. 1.

To release the sulfur from the H₂S compound, all the feed streams are mixed and then combusted at a certain temperature. Since the feed streams contain many components, there are various reactions occurring during the combustion and furthermore, in another unit of the process. However, to simplify the simulation, only the main reaction in the combustion process is considered. All the main reactions in the WSA process are listed in Table 2 [4].

The gas stream from the combustor subsequently enters the SO₂ converter, which consists of a three-layer catalyst bed. The minimum temperature to activate a typical commercial catalyst is between 410 and 430 °C [1]. SO₂ conversion to SO₃ is a highly exothermic reaction. This reaction is commonly conducted under adiabatic conditions so that the energy released from the reaction can be used to increase the temperature of the catalyst. However, as the temperature increases, the atmosphere is no longer favorable for SO₃ formation. To reach high conversion of SO₂ in this work, this reaction is applied in a stage-wise manner where the total catalyst mass is divided into several catalyst beds (Fig. 2). The hot gas leaving each bed is cooled by the steam to the minimum working temperature of the catalyst before it entered the next bed [1].

On this catalyst bed, SO₂ in the gas stream reacted with O₂ by following the stoichiometry of reaction (6). The feed gas to the converter always contains a fraction of water as one of the combustion products. The presence of water in the feed gas promoted the formation of H₂SO₄ on the catalyst bed through reaction (7). Since this reaction consumes some SO₃, reaction (6) is shifted to the right hand side, thus leading to an increase in SO₂ oxidation [6]. Prior to entering the SO₂ converter, steam is added to the feed gas to maintain an excess amount of water to maximize complete hydration of SO₃ to H₂SO₄.

The gas stream moved from the 3rd bed towards the WSA condenser. The gas stream is cooled to 270–290 °C by the steam stream midway between the 3rd bed and WSA condenser. This cooling is necessary because the condensation of H₂SO₄ gas takes place at temperature of 300 °C. Two reactions occur in the condenser unit involving the unreacted SO₃ and H₂SO₄ gas are reactions (7) and (8).

The condenser liquid product, which mostly contains H₂SO₄, then flows to the storage tank. From Fig. 1, it can be seen that some of the H₂SO₄ liquid stream is recycled back to the condenser. The temperature of the recycle stream is lower than the temperature of the feed gas to the condenser. This recycle stream is used to support the condensation process, thus, reducing the amount of necessary external cooling agent.

2.2. Control requirements and process constraints

The safe and effective combustion reaction is defined by two factors: the combustion temperature and the availability of hot air. The feed gas to the combustor contains significant amounts of ammonia, hydrogen cyanide, and other nitrogen compounds that contribute to the formation of NOₓ during combustion [4]. To avoid excessive production of NOₓ, the combustion temperature must not exceed a high level e.g. <1200 °C [1]. In addition, the ability of the combustor material to handle high temperature is limited. The combustion reaction will only take place if there is a sufficient amount of hot air. However, the hot air flow rate must not large
enough to cause a reduction in the amount of waste air released to the atmosphere and energy consumption in the compressor.

The high conversion of SO₂ to SO₃ is dominantly affected by the reaction temperature. The gas temperature to each bed is commonly maintained at 400–500 °C, which is enough to allow the catalyst to function while still providing a favorable environment for SO₃ formation. In this work, the gas temperature was maintained between 350 and 400 °C. It was mentioned in Section 2.1 that to maximize the hydration of SO₃, an additional amount of steam is needed in the feed gas of the SO₂ converter. Therefore, it is necessary to design a set of control structures to calculate the appropriate amount of steam for a given feed gas condition.

The hydration of SO₃ effectively occurred at temperatures of <300 °C. The feed gas from the SO₂ converter exchanged heat with a steam stream to reduce the temperature of the feed gas. An appropriate temperature of the feed stream to the WSA condenser together with a set of control loops to maintain the condensation temperature was necessary to obtain an optimum rate of H₂SO₄ liquid production.

### 3. Control structure design

Plant-wide control is constructed by five layers of hierarchy [7]. The two lowest layers serve as the control layers that work cooperatively to assure the stability of the process while at the same time realizing its economically optimal operation. Stability is controlled by the regulatory layer while the supervisory layer ensures optimal operation. The plant-wide control of the WSA process in this work was developed mainly by placing the proper control structures in the regulatory control layer. The operational optimality of this process was not established through design of the supervisory control layer. Instead, the optimal operation was achieved by selecting the correct set point for each control loop in the regulatory layer. The control structure for the WSA process was designed by carrying out five main steps (Fig. 3). The highlight of each step is listed as follows:

- **The first step was to formulate the correct control objectives of the WSA process that cover operational stability and efficiency.** The proper controlled variables for the defined control objectives are also discussed in this step.
- **The degrees of freedom analysis in the second step aimed to locate the manipulated variables for constructing the regulatory control layer and to identify the free manipulated variables available to achieve the objectives devised for the supervisory layer.**
- **Based on the operational stability objectives derived from the first step, the necessary regulatory control loops were added to the WSA process.**
- **Sensitivity analysis in the fourth step was conducted to find the optimum set point of the variables that were directly affecting the economic efficiency of the WSA process.**
- **In the final step, the resulting control structure was tested to verify its performance of maintaining the operational stability while at the same time ensuring the economic efficiency of the process.**

#### 3.1. Control objectives and process constraints

The main control objective of the WSA process is to maintain the safe and stable production of H₂SO₄ with minimal use of energy despite the variation in the feed gas conditions. All five key units in the WSA process have a significant contribution to the sound operation of this process. The control objectives and constraints for each unit, which are summarized in Table 3, are formulated based on the consideration for safe and effective operation of the process, as described in Section 2.2. After defining the control objectives and constraints in each unit, the proper controlled variables for realizing the objectives were selected as listed in Table 3.

#### 3.2. Degrees of freedom analysis

The operational degrees of freedom in the WSA process can be calculated simply by summing up the number of control valves and adjustable electrical or mechanical variables existing in the process [8]. Based on this concept, from Fig. 1, it was calculated that there were 16 control degrees of freedom in the WSA process that
Table 4
Controlled and manipulated variables in the WSA process.

<table>
<thead>
<tr>
<th>Controlled Variable</th>
<th>Manipulated Variable</th>
<th>Controller</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed stream pressures</td>
<td>Feed stream flow rates</td>
<td>P1, P2, P3, P4</td>
</tr>
<tr>
<td>Combustion temperature [Tc]</td>
<td>Hot air flow rate</td>
<td>T1</td>
</tr>
<tr>
<td>Inlet temperature to SO2 converter [Tc1]</td>
<td>Acid gas flow rate (valve)</td>
<td>T2</td>
</tr>
<tr>
<td>Inlet temperature to 2nd bed of SO2 converter [Tc2]</td>
<td>Steam flow rate (valve)</td>
<td>T3</td>
</tr>
<tr>
<td>Inlet temperature to 3rd bed of SO2 converter [Tc3]</td>
<td>Steam flow rate (valve)</td>
<td>T4</td>
</tr>
<tr>
<td>Inlet temperature to condenser [Tc4]</td>
<td>Steam flow rate (valve)</td>
<td>T5</td>
</tr>
<tr>
<td>Steam flow rate (to feed stream to SO2 converter)</td>
<td>Steam valve</td>
<td>F2</td>
</tr>
<tr>
<td>Temperature of cooling air [Tc]</td>
<td>Cool air flow rate (valve)</td>
<td>T6</td>
</tr>
<tr>
<td>Steam tank liquid level</td>
<td>Makeup steam valve</td>
<td>L1</td>
</tr>
<tr>
<td>H2SO4 tank liquid level</td>
<td>H2SO4 product valve</td>
<td>L2</td>
</tr>
<tr>
<td>Hot air flow rate</td>
<td>Hot air blower duty</td>
<td>F1</td>
</tr>
<tr>
<td>Cool air flow rate</td>
<td>Cool air blower duty</td>
<td>F3</td>
</tr>
</tbody>
</table>

Table 5
Procedure for calculating the minimum required mass flow rate of hot air.

| CO | + | 0.5 O2 | → | CO2 |
| 281.56 kg/h | | 160.89 kg/h | | |
| H2 | + | 0.5 O2 | → | H2O |
| 7.78 kg/h | | 62.28 kg/h | | |
| H2S | + | 1.5 O2 | → | SO2 |
| 815.46 kg/h | | 1151.24 kg/h | | |
| 2 HCN | + | 2.5 O2 | → | 2 CO2 |
| 2.11 kg/h | | 3.27 kg/h | | |
| 2 NH3 | + | 1.5 O2 | → | 3H2O |
| 52.34 kg/h | | 73.89 kg/h | | |
| Minimum O2 | 1451.57 kg/h |
| Required excess O2 | 611.60 kg/h (75% of SO2 mass flow) |
| Minimum stoichiometric O2 | 2063.17 kg/h (95% of minimum stoichiometric O2) |
| Safety factor | 103.16 kg/h |
| Total minimum required O2 | 2166.35 kg/h |
| Total minimum required hot air | 10315.95 kg/h |
| Base case syngas mass flow | 3700 kg/h |

3.3. Regulatory control layer

The controlled variables and the associated manipulated variables in the regulatory control layer of the WSA process are listed in Table 4. Due to variations in the upstream process, the feed stream pressures were subject to change. Therefore, pressure controllers were added to each feed stream to reduce pressure variability in the feed streams that flowed into the combustor.

3.3.1. Minimum required air for the combustion reaction

The combustion temperature is maintained by adjusting the set point of the hot air flow rate, which subsequently defines the required blower power. However, the resulting hot air flow rate from this temperature control loop is only from the perspective of the effectiveness and safety of the combustion reaction. It does not include how much hot air is needed based on the stoichiometry of the reaction.

The minimum required mass flow of hot air for complete conversion of sulfur was calculated by considering the following elements: (1) complete combustion of the syngas component (most importantly sulfur), (2) the efficiency of the subsequent SO2 conversion to SO3, and (3) the safety factor. The first element was determined through simple stoichiometric calculation. High efficiency of SO2 oxidation on the other hand was achieved when the O2/SO2 ratio in the gas mixture after combustion was approximately 0.75 [6]. The safety factor was included as a consideration to ensure a smooth combustion reaction despite any unknown disturbance.

The calculation of the minimum total required mass flow of O2 is summarized in Table 5. The minimum stoichiometric O2 needed for the base case syngas mass flow was 1451.57 kg/h. The O2 had to be supplied in excess to the combustor to create the proper mass flow ratio of O2/SO2 at the combustion product stream. In this case, the value of 0.75 for the O2/SO2 ratio was selected and therefore, the required O2 excess was 611.60 kg/h. A safety factor of 5% was chosen and hence, additional O2 was at 103.16 kg/h. The total mass flow of O2 was then multiplied with the conversion factor of 100/21 to obtain the minimum required mass flow of hot air of 10315.95 kg/h. The implementation of the controller above is by measuring the flow of the feed gases (i.e. acid gas, SSS gas, SWS gas, and SWS-off gas) and multiple those with constants used for calculation of required stoichiometrical flow of supply air for complete combustion of feed gases without excess of O2.

The combustion temperature is the dominant variable that defines how much hot air should be supplied to the combustor. The combustion temperature is inversely proportional to the mass flow of hot air. A high selector block was used to ensure that the combustion temperature was always at the desired value without violating the minimum required hot air for the combustion reaction. This block compared two signals: one representing the minimum required hot air (Aspen Hysys spreadsheet) and the other from the combustion temperature controller. Subsequently, the selector block selected the highest signal and sent it to the hot air mass flow controller as the set point. The selected mass flow signal was then translated as the required blower duty to circulate the hot air. Fig. 4 illustrates the control structure surrounding the combustor.
3.3.2. Efficient $SO_2$ conversion

The efficiency of $SO_2$ conversion to $SO_3$ is defined by the feed gas temperature. Therefore, the $SO_2$ converter must be equipped with a scheme to tightly control the feed temperature. There are two possible variables that can be adjusted to keep the temperature at the desired level, the feed gas, and the steam flow rate. Controlling the feed temperature by directly adjusting the feed flow rate is not considered as the best method. The feed gas to the $SO_2$ converter is the product of the combustion reaction. Therefore, if the feed gas flow rate is varied to meet the temperature set point, it will disturb the combustor reactions.

One proven way to control the feed gas temperature without affecting its flow rate is by using the bypass control method (Fig. 5a). The main feed gas stream is split into two streams: one is heading to the heat exchanger and is cooled down by the steam while the other stream is bypassed and then remixed with the outlet stream from the heat exchanger. The bypass valve in this arrangement works in reverse in terms of control action i.e. when the temperature of the mixed stream decreases, the opening of the bypass valve increases to elevate the temperature back to the set point. However, the pressure drop across the heat exchanger limits the flow rate of the stream passing through the bypass valve. Thus, in this type of arrangement, the bypass valve is susceptible to reaching saturation where the valve is either fully opened or closed while the controller fails to meet the temperature set point. Another drawback of this configuration is that the remixing of the main stream and the bypass stream, which is hotter than the main stream, will reduce the temperature difference across the heat exchanger. Therefore, the heat exchanger must have a bigger area to achieve the desired low temperature [9,10].

The limitation in the single valve arrangement can be overcome by employing the bypass valve to maintain the pressure drop across the bypass stream while the main stream valve is adjusted to control the temperature of the process gas (Fig. 5b). The pressure differential controller (PDC) controls the flow rate through the bypass stream by fixing the pressure drop across the bypass valve as being always equal to the total pressure drop across the main stream valve and heat exchanger. In this configuration, when the
flow rate of the main stream gas increases because of the decrease in the outlet temperature, the flow rate through the bypass valve decreases with the same amount as the deviation in the main stream flow rate. Thus, variation in the process gas flow rate can be avoided. The PDC structure, compared to the single valve arrangement, also has faster response both in tracking a new temperature set point (Fig. 6a) and rejecting disturbances (Fig. 6c and d).

3.4. Optimal operation of the WSA process

The optimal operation of the WSA process is when the highest production rate of H$_2$SO$_4$ is obtained at a minimum blower duty. This can be realized by selecting an appropriate set point for the control loop that has a direct effect on the blower duty and the H$_2$SO$_4$ production rate. To select the appropriate set point, dynamic simulation was utilized.

3.4.1. Optimum combustion temperature

The minimum flow rate of hot air to fulfill the requirement of the stoichiometric combustion reaction was calculated in the Section 3.3.1. However, this calculation does not include the economic aspect of the reaction. The temperature of the combustion reaction has a direct link to the optimal operation of the WSA process because it drives how much hot air must be supplied and thus defines the blower duty.

From Fig. 7a, it can be seen that the blower duty decreases as the combustion temperature increases. On the other hand, the flow rate of H$_2$SO$_4$ increases proportionally with the combustion temperature until a point where increasing the combustion temperature reduces the H$_2$SO$_4$ flow rate (Fig. 7b). The optimum combustion temperature is shifted to a higher value at a higher feed rate i.e. it is 1020, 1035 and 1065 °C for a −5%, base case and +5% feed rate, respectively. For some of the operational degrees of freedom, although optimal values change according to the operating condition, constant values can be conveniently used with marginal loss of the H$_2$SO$_4$ product. 1040 °C was selected as the set point temperature for combustion considering that at the given range of feed rate variations, in every flow rate case, the H$_2$SO$_4$ production rate was maintained at a reasonably optimum value with relatively low blower duty. This set point was selected also by considering the high temperature limit that the reactor material could handle and
the fact that higher combustion reaction temperature results in higher NOx production.

3.4.2. Optimum SO2 converter temperatures

The search for the optimum set point was extended to other controlled temperatures because these variables also have a direct effect on the hot air blower duty and H2SO4 production rate. From Fig. 8, it can be observed that ±5% variation in the feed flow rate from the base case results in a larger deviation in blower duty but only a small change in the H2SO4 production rate. The optimum set point for every controlled temperature at each flow rate case was not similar, thus, several aspects have to be compromised to find the proper set point. For instance, the optimum value for T2 at the base case is 392 °C while in the +5% case, the optimum set point is 400 °C. Furthermore, at the +5% case, the WSA process has a convergent solution when T2 is >398 °C. Therefore, the set point of T2 should not be lower than 398 °C to assure the sound operation of the WSA process when the feed flow rate is increasing up to 5%. Regardless, choosing the set point for T2 at any value >398 °C i.e. 400 °C will still be advantageous for each flow rate case because the blower duty is almost equal to the optimum value.

This also applies to T3. At higher feed flow rate, the minimum value of T3 that the WSA process can still handle is higher than the minimum value in the system with lower feed flow rate. Therefore, to maintain feasible operation at a wider range of feed flow rate, the minimum set point of T3 should be higher than the lowest acceptable value in the +5% flow rate case i.e. >394 °C. At the right hand side of the optimum value of T2 and T3, the profile of the blower duty and H2SO4 production rate is relatively flat. This flat region shows that there are a wide set of options for the set point of T2 and T3. Again, selecting the highest temperature within this range as the operating temperature for T2 and T3 will still be economically beneficial because the blower duty and the H2SO4 production rate is close to the optimum value in this case.

Despite the wide range of set points allowed for T2 and T3, the optimum value for T4 is limited. In contrary to T2 and T3, increasing T4 results in higher blower duty and decreasing H2SO4 production. From the process structure, it can be seen that the outlet steam from the 3rd bed will be returned to the steam tank. When T4 is higher, the outlet temperature of the steam from the 3rd bed will consequently be higher. Subsequently, the returning temperature of the steam after exchanging heat with this stream will also be
increasing. Hence, the temperature of the steam flowing from the tank to cool down the inlet stream to the SO₂ converter is higher. Since a PDC structure is controlling T₂, the temperature controller will respond to this situation by increasing the opening of the main steam valve (V5). As a result, the flow rate of the process gas increases and subsequently, more hot air for combustion is required and blower duty increases. As for T₂ and T₃, the minimum feasible temperature for the 3rd bed at the +5% flow rate is higher than base case and −5% case. However, the profile of the blower duty and H₂SO₄ production rate with varying T₄ is somewhat different compared to the case of T₂ and T₃. There is no flat region in these profiles that can give freedom in choosing the proper set point for T₄. Therefore, the temperature set-point for T₄ that ensures the optimal operation of the WSA process while keeping the operational feasibility is the lowest value in the +5% flow rate case i.e. 359 °C.

3.4.3. Optimum WSA condenser temperature

There is no optimum point of the WSA condenser temperature in its relation to the blower duty and H₂SO₄ production rate (Fig. 9). The lowest blower duty and the highest H₂SO₄ production rate are obtained at the lowest WSA condenser temperature. In this particular feed rate range, the minimum allowable WSA condenser temperature is 297.4 °C. Any value lower than this, with the given equipment design will reduce the feasibility of the process because of saturation in the control element that keeps the WSA condenser temperature at its set point.

3.5. Control structure performance

Six types of disturbance tests were performed on the WSA process through the acid gas feed stream i.e. ±1 kPa in feed pressure, ±3 °C in feed temperature, and ±0.01 in H₂S composition. The test was conducted using the optimum operating conditions as explained in the previous section. The ability of each controller to maintain the predefined set points is not guaranteed; in fact, the magnitude of the disturbances defines the boundary. Most of the temperature controllers in the WSA process, as illustrated in Fig. 10, successfully accommodated the ±1 kPa variation in the acid feed gas stream, except for the controller of inlet temperature to the
condenser, which was unable to damp out this disturbance and failed to keep the set point (Fig. 10).

Similar responses were also found when the acid gas feed stream temperature was step changed by ±3 °C. The temperature of the combustion and each bed in the SO2 converter returned to their set point at relatively short settling time despite this disturbance while the behavior of the WSA condenser temperature (Fig. 11i and j) was somewhat different. After 1000 min, the temperature controller, in spite of having relatively small integral time, was not able to remove the offset in the inlet temperature to the WSA condenser.

The responses of temperatures in the SO2 converter and WSA condenser after variations in H2S composition were quite similar to the responses after feed pressure variation (Fig. 12). The temperature controller in the WSA condenser was not able to maintain...
the temperature set point even though the magnitude of H$_2$S composition variation was considerably small.

These results lead to questions regarding the benefit of controlling the temperature of the inlet stream to the WSA condenser. Fig. 13 compares the profile of the H$_2$SO$_4$ production rate with and without a temperature controller in the inlet stream to the WSA condenser. The comparison is made only for the disturbance cases (+1 kPa of the acid feed gas pressure and +0.01 of the H$_2$S composition) where the WSA condenser temperature controller successfully maintained the temperature set point. The comparison shows that keeping this temperature at a fixed value does result in a constant H$_2$SO$_4$ production rate.

4. Conclusions

This work discussed the control structure design of the WSA process from a plant-wide perspective. The control objectives that must be achieved along with several process constraints were summarized. A regulatory control layer was developed by taking into account control requirements and constraints demanded by each unit. In this work, we proposed a scheme with several controllers, which achieves operational stability of the WSA process. Optimal operation of the WSA process was attained by providing the proper set point for every control loop with a controlled variable that has a direct effect on the blower duty and H$_2$SO$_4$ production rate. The temperature of the inlet stream to the WSA condenser was expected to be the variable that when kept constant would indirectly maintain the H$_2$SO$_4$ production rate at an optimal value. The results however show that the H$_2$SO$_4$ production rate is insensitive to this variable.

Acknowledgement

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Appendix A. WSA process parameter and size of the equipment

See Fig. A1 Table A1.
Table A1
Size of the equipment.

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Modeled By</th>
<th>Size</th>
</tr>
</thead>
<tbody>
<tr>
<td>K.O Vessel SSS gas</td>
<td>Separator</td>
<td>Diameter: 0.398 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 1.5 m</td>
</tr>
<tr>
<td>K.O Vessel acid gas</td>
<td>Separator</td>
<td>Diameter: 0.8 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 1.5 m</td>
</tr>
<tr>
<td>K.O Vessel SWS gas</td>
<td>Separator</td>
<td>Diameter: 0.398 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 1.5 m</td>
</tr>
<tr>
<td>K.O Vessel SWS off-gas</td>
<td>Separator</td>
<td>Diameter: 0.398 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 1.5 m</td>
</tr>
<tr>
<td>Combustor</td>
<td>Gibbs Reactor</td>
<td>Diameter: 1.2 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Length: 1.5 m</td>
</tr>
<tr>
<td>Steam Drum</td>
<td>Tank</td>
<td>Diameter: 1.6 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Length: 2.6 m</td>
</tr>
<tr>
<td>SO₂ Converter</td>
<td>3 Conversion Reactor and 3 Heat Exchanger</td>
<td>1st Conversion Reactor (1st Bed)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Diameter: 4.7 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 3.1 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>2nd Conversion Reactor (2nd Bed)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Diameter: 4.7 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 2.81 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>3rd Conversion Reactor (3rd Bed)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Diameter: 4.7 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 2.25 m</td>
</tr>
<tr>
<td>Condenser</td>
<td>Conversion Reactor, Heat Exchanger, and Separator</td>
<td>Conversion Reactor</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Diameter: 1.6 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Height: 1 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Heat Exchanger</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Tube Volume: 1.2 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Shell Volume: 1.8 m</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Separator</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Volume: 20 m³</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Diameter: 1.5 m</td>
</tr>
<tr>
<td>H₂SO₄ Tank</td>
<td>Separator</td>
<td>Volume: 6.805 m³</td>
</tr>
</tbody>
</table>

Fig. A1. WSA process and its parameter.
Table B1
Reactions parameter in WSA process model.

<table>
<thead>
<tr>
<th>Equipment</th>
<th>Modeled By</th>
<th>Reaction Parameter</th>
</tr>
</thead>
<tbody>
<tr>
<td>Combustor SO2 Converter</td>
<td>Gibbs Reactor 3 Conversion Reactor and 3 Heat Exchanger</td>
<td>Gibbs Reactions</td>
</tr>
<tr>
<td></td>
<td></td>
<td>1st Conversion Reactor (1st Bed)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Conversion (%) = 99%</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Temperature (°C) = 357.2</td>
</tr>
<tr>
<td></td>
<td></td>
<td>2nd Conversion Reactor (2nd Bed)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Conversion (%) = 0.4623 ° Temperature (°C) = 96.65</td>
</tr>
<tr>
<td></td>
<td></td>
<td>3rd Conversion Reactor (3rd Bed)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Conversion (%) = 1.656 ° Temperature (°C) = 517.1</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Conversion = 99%</td>
</tr>
<tr>
<td>Condenser</td>
<td>Conversion Reactor, Heat Exchanger, and Separator</td>
<td></td>
</tr>
</tbody>
</table>

Appendix B. Reactions parameter

In combustor, Gibbs reactor is used as a model. The chemical equilibrium state is often used in combustion engineering as a reference point for chemical kinetics. In addition, in a spontaneous change, Gibbs energy always decreases and never increases. Hence, because combustion reaction is spontaneous, it can be considered getting to equilibrium state with minimum Gibbs free energy.

In SO2 converter, conversion of SO2 have strong relation with temperature as depicted in Fig. 2. Therefore, the authors made regression and used the regression model in a spreadsheet to calculate the conversion, then input the conversion in reaction conversion modeled by conversion reactor.

WSA condenser in real plant has three functions which are:

a) Reactor (converting SO3 to H2SO4)
b) Heat exchanger (providing area to condensate SO3 gas and H2SO4 gas to H2SO4 liquid)
c) Separator (separating the clean gas and liquid H2SO4)

In Aspen Hysys, these three functions cannot be simulated in one equipment, so that three equipments are used to simulate one WSA condenser which are conversion reactor, heat exchanger, and separator. In the real plant, the reaction of SO3 to H2SO4 is affected by the cooling air temperature and mass flow (how much heat is absorbed by cooling air). Generally, almost all SO3 is converted to H2SO4. Hence, it is assumed that the conversion is 99%.

See Table B1.

References