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Refining plant energy optimization

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Abstract A refining plant in the Middle East began official production in 2020 to produce strategic products. After production stabilization, the plant operators decided to implement energy optimization. While the refining plant utilizes a large number of units, only four units were selected for optimization. In addition, two scenarios were studied for optimization. Scenario 1 involves stopping the use of the sour water stripper units (SWS) to save 18.85 ton/h of steam used for stripping. The SWSs strip hydrogen sulfide (H_2S) and ammonia from the sour water of the refinery units. Meanwhile, scenario 2 considered optimizing the steam used in the SWSs and amine regeneration units (ARUs). The ARUs regenerate rich amine using H_2S from all the refining units to provide the lean amine required for gas sweetening. The overhead acid gas from the SWSs and ARUs feed the sulfur recovery unit (SRU) to prevent emissions from exceeding environmental regulations. The feed to steam ratio of the SWS reboilers was optimized from 0.17 to 0.16 kg steam/kg feed) and from 0.14 to 0.10 kg steam/kg feed for the ARU reboilers. Overall, scenario 2 saved 23.09 ton/h of stripped steam. After a feasibility analysis, scenario 2 was selected for implementation. Considering the average steam cost of 7.6$/ton, scenario 2 saved 1,537,206.38 $/year. In total, five different simulations were conducted in this study, four of which were performed using Aspen HYSYS V.11. Conversely, the other simulation, which simulated the entire SRU plant, used a special sulfur package. In addition, a combined heat and power model was used to evaluate the thermodynamic properties, material, and energy balance equations for the live steam balances.

1. Introduction

As the world's population rapidly expands, industrialization is increasing, leading to an increase in energy consumption [1]. Achieving optimal energy consumption is considered to be one of the key parameters for community development. Therefore, it is essential to optimize energy consumption and prevent losses by different industries. High energy consumption in chemical processes increases operation and production costs and reduces system efficiency [2]. The petroleum refinery industry is an essential contributor to the global economy, producing different types of products, such as chemicals and fuels, that participate in the global market [3]. However, the sour water produced from the refinery industry contains many

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Abbreviations

AAG	Amine Acid Gas	MPS	Medium Pressure Steam
AG	Acid Gas	NHT	Naphtha Hydrotreating
ARU	Amine Regeneration Unit	R	Regenerator
CCR	Continuous Regeneration Reforming	RA	Rich Amine
CHP	Combined Heat and Power	S	Stripper
CCI	Coal Chemical Industry	SRE	Sulphur Recovery Efficiency
DCU	Delayed Coking Unit	SRU	Sulphur Recovery Unit
DEA	diethanolamine	S/F	Steam to Feed ratio
DEU	De-ethanizer Unit	SWS	Sour Water Stripping
DHT	Diesel Hydrotreating Unit	SWSAG	Sour Water Stripping Acid Gas
HCU	Hydrocracking Unit	TG	Tail Gas
HPS	High Pressure Steam	TGT	Tail Gas Treatment
HPU	Hydrogen Production Unit	TGTU	Tail Gas Treatment Unit
LA	Lean Amine	VDU	Vacuum Distillation Unit
LP	Low Pressure	WHB	Waste Heat Boiler
MDEA	methyl diethanolamine	Wt.	Weight

hazardous pollutants, in which hydrogen sulfide (H_2S) and ammonia (NH_3) are the primary pollutants [4–6]. These two pollutants can be removed by specially designed strippers [7–9]. Meanwhile, the sour gas produced by refineries contains H_2S. Thus, gas sweetening is performed in an amine scrubber unit, wherein sour gas is placed in contact with a lean amine (LA) solution in an absorber to absorb H_2S. The rich amine (RA) with H_2S is then stripped in an amine regenerator column [10,11]. In particular, diethanolamine (DEA) and methyl DEA (MDEA) amines are widely used in gas sweetening [2,12–14]. MDEA is used when acidic gas contains both CO_2 and H_2S, and the process must absorb H_2S and desorb CO_2. DEA has a higher selectivity for H_2S than CO_2 [15–17]. H_2S is a hazardous acidic gas owing to its toxicity and corrosivity. It can produce acid rain, which can damage both equipment and human health [1,2]. Further, it is used as a feed to sulfur recovery unit (SRU) plants to produce elemental sulfur [18,19]. The main purpose of SRU plants is to prevent H_2S emissions from exceeding environmental regulations. The most used process in these plants is the modified Claus process [20–25]. A refinery column in the Middle East began official production in 2020. This refinery plant is designed to convert heavy intermediate products of low economic value into middle distillates of high economic value. The refinery includes a vacuum distillation unit (VDU), delayed coking unit (DCU), hydrocracking unit (HCU), diesel hydrotreating unit (DHT), naphtha hydrotreating unit (NHT), continuous regeneration reforming unit (CCR), de-ethanizer unit (DEU), hydrogen production unit (HPU), and a sulfur recovery unit (SRU). The refinery plant also includes two sour-water stripper units (SWS1 and SWS2) to strip any H_2S and NH_3 from the sour water of the refinery units and two amine regenerator units (ARU1 and ARU2) to regenerate RA to LA. This LA is then recycled for gas sweetening. After achieving production stabilization, the goal of this plant was to optimize plant energy use. The main units of interest for optimization were the SWSs, ARUs, and SRUs. Two optimization scenarios were studied, one of which was selected. In a literature survey, the term optimization is present in thousands of studies concerning various industries. Moreover, there are various studies regarding optimization in petrochemical and refinery industries. For example, Gong et al. studied the energy efficiency enhancement of energy and materials for ethylene production [26], achieving an energy consumption reduction of 4.89% for cracking production. Deymi-Dashtebayaz et al. studied the use of low-pressure steam from a refinery in a desalination process [27]. Javadpour et al. performed a study to increase the performance of cooling towers using nanofluids owing to the use of cooling towers in refining industries. Their results proved that using nanofluids enhanced the performance of the cooling towers [28]. Li et al. studied the energy use of the coal chemical industry (CCI) rather than petroleum fuels as petroleum sources are becoming exhausted. They proved that using this technology leads to a significant decrease in cost [29]. Chehade et al. performed a simulation and optimization analysis of a steam reformer process using HYSYS and MATLAB, achieving an energy consumption reduction of 77.5% [30]. Liesche et al. presented an optimization study called the FluxMax approach for the production of hydrogen cyanide, reducing the total variable cost by 68% [31]. Karimi et al. studied the application of a heat recovery steam generator to optimize plant efficiency in gasoline-kerosene units in a petroleum refinery in Iran. They found that the refinery could generate steam with a maximum profit of $6,650,000 during a 10-year operation period, and save 1500 kg/h of natural gas from burning, preventing a considerable amount of CO_2 release to the atmosphere [32].

However, the literature survey did not reveal any articles linking the sum of the refinery units for energy optimization. Rather, it revealed optimization studies considering only parts of the refinery. For example, large SRUs have been studied, but they are only one individual, small part of the entire refinery. When the steam optimization decision first began, the units considered for steam optimization were the low-pressure steam end users, which is the steam used in stripping strippers 1 2, and regenerators 1 and 2), because any decrease in the LP steam indicates a direct decrease in boiler steam production. Therefore, this study aims to provide an opportunity for researchers and readers to observe the actual situation in large refinery plants with a large number of units. This research study connects the actual experience of a number of engineers in different refining units with the scientific meaning of optimization.

2. Optimization scenarios

2.1. Scenario 1

The first scenario was to stop using the SWS1 and SWS2 units to save the steam required in the four reboilers providing steam to Strippers 1 and 2. In this case, the acid gas fed to the SRU unit was provided only from ARU1 and ARU2. The stripped water from SWS1 and SWS2 is normally used as washing water in the process units in order to optimize the use of freshwater. However, the chloride content in the stripped water is high, increasing the corrosion risk to the equipment. Thus, the stripped water was discharged to the sewer instead of being recycled for other process demands. At this stage, scenario one was further examined as the discharge of the stripped water did not benefit the process. Turning off the stripped water units will save 18.85 t/h of stripping steam used in Strippers 1 and 2 reboilers, but other effects should be noted. The overhead acid gas from SWS1 and SWS2 is referred to as sour water stripped acid gas (SWSAG), which feeds the SRU unit with the addition of ARU1 and ARU2 acid gas. Acid gas is then oxidized in the reaction furnace of the SRU via a series of exothermic reactions, forming the flame of the reaction furnace. The gas produced from the furnace is further cooled in the waste heat boiler (WHB), producing 35.3 t/h of high-pressure steam (HPS). Cutting the SWSAG reduces the feed to the furnace by 3674 kg/h. The expected decrease in steam production from the WHB should be studied as the decrease will be compensated by increasing the capacity of the boilers. A feed with a rate of 11975 kg/h from ARU1 and ARU2 was maintained in the reaction furnace. This feed is called an amine acid gas (AAG) feed.

2.2. Scenario 2

The second scenario keeps all the SWSs and ARUs running in normal operation while optimizing steam consumption in the SWSs and ARUs. Normally, reboilers use a stable calculated steam-to-feed (S/F) ratio. The S/F ratio for the SWS reboilers must be 0.17 kg steam/kg feed to achieve a stripped water specification of maxima of 10 ppm-Wt. H_2S and 50 ppm-Wt·NH_3. Meanwhile, for the ARU reboilers, the S/F ratio must be 0.14 kg steam/kg feed to achieve a maximum 0.2 H_2S Wt.%. It was further observed that the water concentrations were 1 ppm H_2S and less than 10 ppm NH_3. The perspective of this scenario is that a chance exists to optimize the steam while achieving certain design specifications. The LA produced from ARU1 and ARU2 has only 0.05 H_2S Wt.%. Therefore, it is possible to optimize the steam used in Regenerator1 and Regenerator2. Herein, the aim was to decrease the steam used for reboilers gradually and work on a new optimized S/F ratio for the SWS and ARUs. The new optimized S/F ratios for the SWS and ARUs are 0.16 and 0.10, respectively.

3. Materials and methods

3.1. Simulation steps and process description

The study is composed of five simulations, including simulations of the Stripper 1 of SWS1, Stripper 2 of SWS2, Regenerator 1 of ARU1, and Regenerator 2 of ARU2, and a complete simulation of an SRU plant.

3.1.1. Simulation of Strippers 1 and 2

The sour water stripping unit removes hydrogen sulfide and ammonia from the process wastewater streams using a stripper column. Hydrogen sulfide and ammonia are stripped from the sour water as they flow downward through the stripper. The H_2S and NH_3 contents in sour water from the stripper bottom cannot be more than 10 ppm-Wt. and 50 ppm-Wt., respectively. The sour water stripper reboiler supplies heat to the stripper to strip the H_2S and NH_3 from the sour water. This reboiler is supplied with a low-pressure (LP) saturated steam as a heating medium. The overhead acid gas from the top of the stripper is then sent to the SRU. Fig. 1 shows the Stripper 1 simulation. The package selected for this simulation was the Peng Robinson package. Note that it is mandatory to select a proper package to prevent complete deviations in the results.

The strippers were first selected as normal distillation columns, but later some modifications and adjustments were made. Initially, the tower requires some initial data, including the number of trays, feed tray, reboiler pressure, condenser pressure, and the vent rate). After, some specifications are added to solve the column calculations, such as the bottom rate and reflux rate. Strippers 1 and 2 use a pump around system instead of normal condensers. Fig. 2 shows Stripper 2 simulation.

Herein, the overhead condenser system was removed, and a pump around the system was added. For normal distillation columns, two specifications are sufficient to solve for the column, but with the addition of the pump, approximately three specifications are required. Product specification was added as the third specification. After the column is already solved in the HYSYS, it is easier to change any flow specifications, such as temperature and pressure, to make the column flexible for any changes in feed conditions or flow rates.

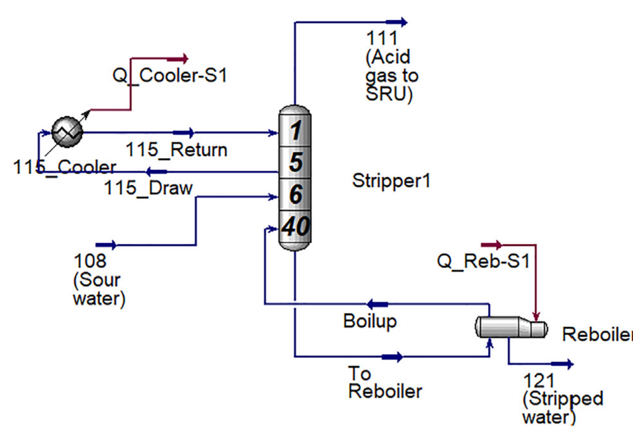

Fig. 1 Stripper 1 simulation.

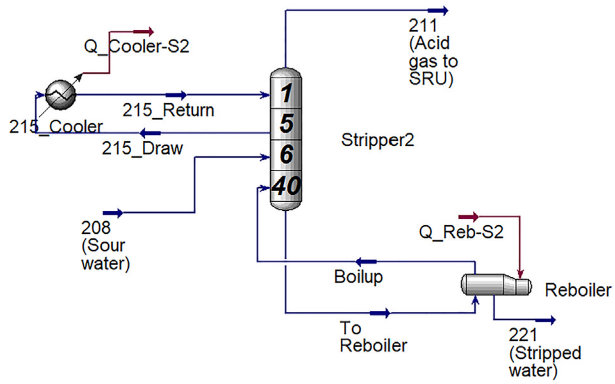

Fig. 2 Stripper 2 simulation.

3.1.2. Simulations of Regenerators 1 and 2

An amine regenerator is designed to remove H_2S from RA. H_2S is stripped from the RA solution as it flows down through the regenerator. The regenerator reboiler supplies heat to the regenerator to strip the H_2S from the RA solution. The maximum H_2S Wt.% in the LA solution is 0.2%. The reboiler is supplied with an LP saturated stream as a heating medium. The acid gas from the regenerator is then routed to the SRU. Fig. 3 shows the simulation of Regenerator 1.

Regenerators 1 and 2 were simulated as normal distillation towers. Fig. 4 shows the simulation of Regenerator 2.

The fluid package selected herein was the "chemical solvent package," which is suitable for the DEA solvent existing in ARU units. In this case, it was mandatory to use this package. Further, the selected package provides highly accurate results. The columns require similar information as the strippers, including the number of trays, feed trays, reboiler pressure, condenser pressure, and vent rate. Only two specifications were added to solve the columns as they are normal distillation columns.

3.1.3. Simulation and process description of the sulfur recovery plant

The SRU is designed to recover sulfur from AAG and SWSAG with 99.9% sulfur recovery efficiency. Fig. 5 shows the results of the SRU simulations. The SRU plant was simulated with a special sulfur package named SULSIM. This package incorporates empirical correlations fitted to the plant data and provides the ability to use rigorous simulations to accurately model different SRU equipment and other related operations of a sulfur recovery plant. Further, it provides the ability to meet stringent environmental regulations and standards in flare gases, especially H_2S and SO_2 emission limits. The SRU plant is composed of different sections: the Claus section, tail gas treatment unit (TGTU), degassing section, and the incinerator section.

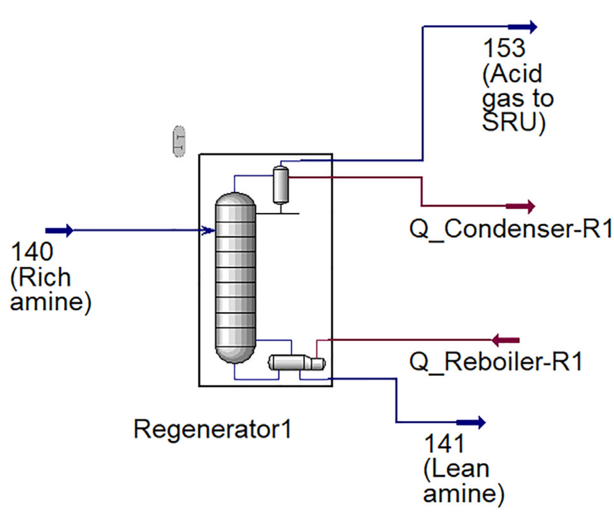

Fig. 3 Regenerator 1 simulation.

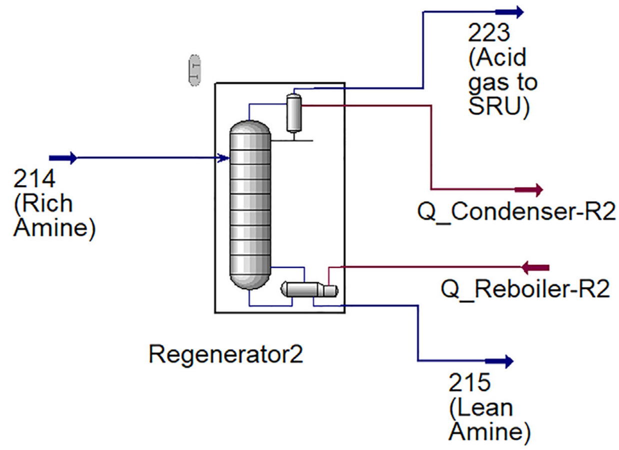

Fig. 4 Regenerator 2 simulation.

3.1.4. Simulation and process description of the Claus section

The Claus section consists of the thermal Claus and catalyst Claus sections. The Claus process is used on acid gas (AG) streams containing H_2S and NH_3. The concept of the process is that one-third of the H_2S contained in the acid gas feed is transformed into SO_2 in the thermal Claus section, which accounts for approximately 70% of the sulfur conversion. Subsequently, the SO_2 reacts with the remaining two-thirds of H_2S to form sulfur in the catalytic Claus section, which performs the remaining 30% of the conversion. The SWSAG and AAG feed the thermal reactor in the Claus section. The air to the main burner of the reactor is sufficient to complete the oxidation of all hydrocarbons and NH_3 present in the total feed gases as well as burn the H_2S. The reactions in the thermal reactor are exothermic. The waste heat contained in the process gas leaving the thermal reactor is recovered by producing high-pressure steam in the WHB. The thermal reactor was described in the SRU package as a "reaction furnace with two chambers". The empirical model used was the NH_3 SWSAG legacy as the feed to the SRU plant contains both gas components. Other models exist but are not suitable for this case. A single-pass WHB was selected from the SRU model pallet. Fig. 6 shows the thermal reactor with the WHB as it is an essential part of scenario 1.

The two reactors used to complete sulfur conversion in the catalytic section were selected to be catalytic converters. The sulfur produced is condensed in sulfur condensers, and thus each condenser is selected as a "sulfur condenser."

3.1.5. Simulation and process description of the TGTU

The TGTU treats the Claus tail gas (TG) from the Claus section to convert SO_2 into H_2S. The converted H_2S is cooled,

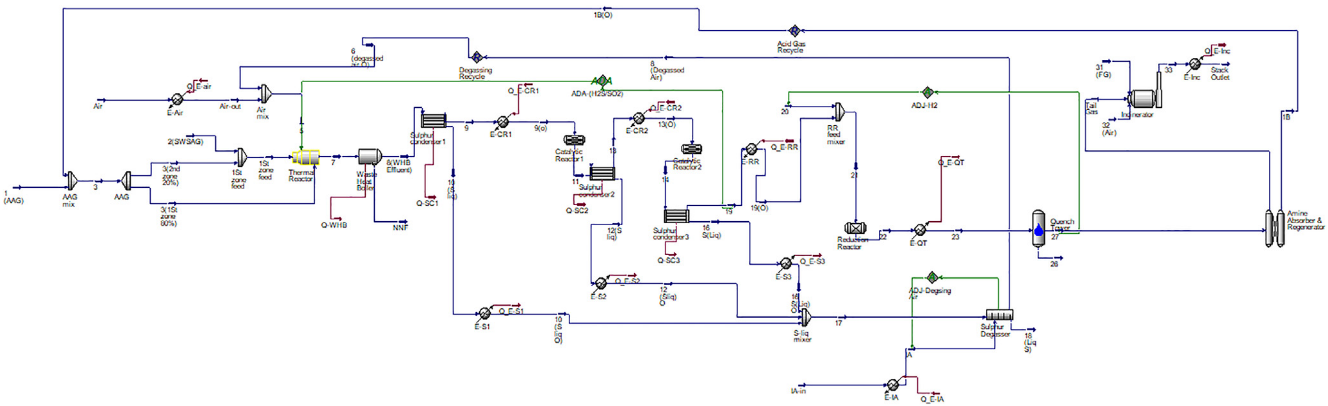

Fig. 5 Sulfide recovery unit (SRU) plant simulation.

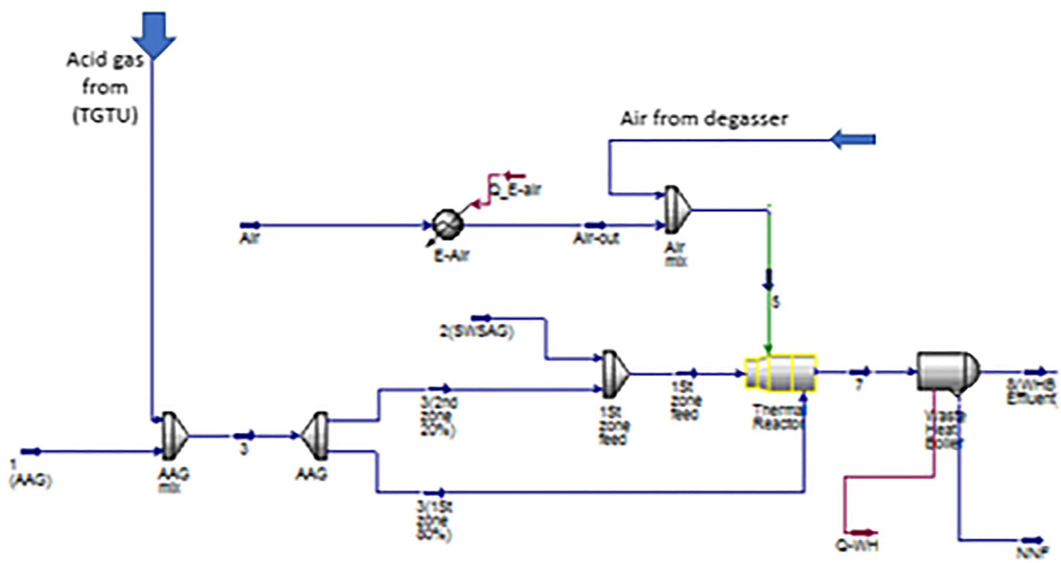

Fig. 6 Thermal reactor and waste heat boiler (WHB).

absorbed by the LA, and then recycled to the thermal Claus section for reprocessing. The tail gases coming from the Claus section are preheated in the TG heater heat exchanger using superheated high-pressure steam as a heating medium. A hydrogen-rich gas stream is mixed with the process gas downstream of the heater to supply the required hydrogen for the hydrogenation of the sulfur species. The reduction reactor is filled with a specific reduction/hydrolysis catalyst to convert any SO_2 compounds into H_2S. Because of the exothermic reactions, the process gas temperature increases. Heat recovery is achieved in the TGT WHB producing LP steam. The final cooling of the process gas occurs in the quench tower. The H_2S absorption step is accomplished using a formulated MDEA-based LA solution with a 45 wt% concentration. The RA solution leaving the bottom of the absorber is pumped using RA pumps to the TGT regeneration section, where it is regenerated. The regenerated LA is cooled in the LA/RA exchanger, pumped through the LA pumps to the LA cooler, and finally routed back to the TGT absorber. The acid gas stream stripped from the RA solution is recycled to the Claus train as additional feedstock. The reduction reactor is selected from the SULSIM as a "hydrogenation bed," and the quench tower has the same name in the package: "quench tower." Further, the amine scrubber unit is a simple amine absorber and regenerator".

3.1.6. Simulation and process description of degassing section

The liquid sulfur produced in the SRU contains soluble H_2S and H_2S_x (hydrogen polysulfides). During sulfur conveyance and handling, the presence of H_2S in the liquid can cause safety and environmental problems because of its toxicity and explosion hazard. Therefore, liquid sulfur is degassed to reduce the H_2S content to a safety value of 10 ppm-Wt. The degasser was selected from the SULSIM package as "sulfur degasser," and the outlet liquid H_2S concentration was set at 10 ppm-Wt.

3.1.7. Simulation and process description of incinerator section

Incinerating the TG produced in the Claus and TGT units is necessary to transform all the sulfur compounds present into SO_2. The flue gas produced during incineration is discharged into the atmosphere via a stack. The TG ignition temperature is much higher than the actual tail gas temperature, as all its fuel components are at very low concentrations. Therefore, TG combustion must be supported by natural gas combustion. The incineration combustion chamber temperature is 650 °C during normal operation. This temperature is necessary to ensure the nearly complete combustion of the H_2S (less than 10 ppm residual H_2S is expected) and other sulfur compounds contained in the TG. The incinerator has the same name in the HYSYS, and thus was selected as "incinerator."

3.1.8. Steam system

Refinery boilers are designed to produce 119.6 ton/h of steam for use in different units. There are three main steam headers in the plant. The HPS header, medium-pressure steam header, and low-pressure steam header. The steam produced from the boilers provides the HPS header. Additional steam production is added to the high-pressure main steam header from the internal equipment in the plant, such as the WHBs. The average steam cost is 7.6 $/ton. In general, this cost is based on the fuel cost and chemicals used for boiler treatment. Fig. 7 shows the simple scheme for the steam system used in our study. The steam flows related to our study are shown in the figure. A combined heat and power (CHP) model was calculated using daily actual steam and power consumption data from the plant. This model was developed based on material balance and energy balance equations and uses steam tables to compute the enthalpy required for the steam calculations.

3.2. Plant calculation concept

The refinery plant requires various equipment, including reactors, heat exchangers, heaters, coolers, mixers, boilers, reboilers, and turbines. Most of this equipment contributes to each equipment category. However, this study is mainly focused on the optimization of steam from the SRU, ARU, and SWS units. Note that any steam reduction has a direct relationship with steam reduction in boilers

3.2.1. Plant steam enthalpy calculations

Heat and power energy are delivered inside the refinery via fired heaters, electricity, and steam. Among these utilities, only steam can deliver heat via steam heaters and power via steam turbines. Moreover, steam can be generated or superheated in the convection section of the fired heaters. Most of the steam-driven compressors and pumps have electric motor-driven equipment. Therefore, heat and power within the refinery are interlinked and interfaced in every process unit. CHP models normally check the heat and mass balance of the entire heat and power network. Specifically, the model identifies imbalances in each main steam header and evaluates the limits of each process unit battery. Further, the model can calculate the net steam demand from the utility boilers to optimize the boiler load. The heat and mass balance of the steam system requires a linked thermodynamic package (steam tables) to calculate each stream enthalpy.

3.2.2. Other plant enthalpy calculations

The process enthalpy calculations for the streams are output from the HYSYS software as it normally requires some dead points for the calculations. For example, using Excel provides values that are far from realistic. Thus, for steam, only the CHP model uses Excel software embedded with energy balance equations and steam tables.

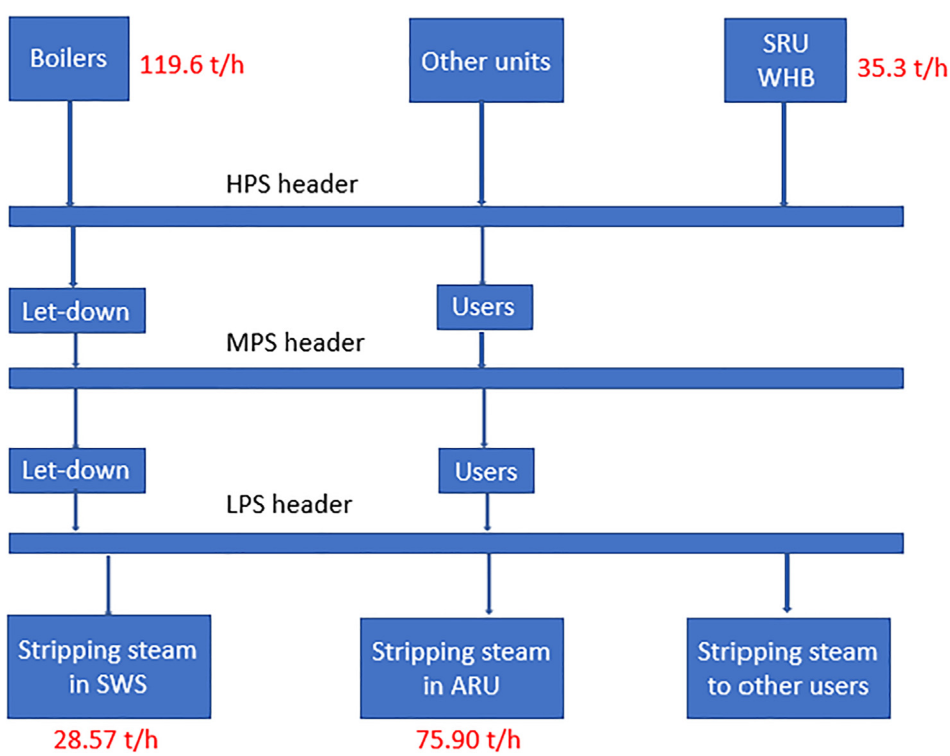

Fig. 7 Steam systems.

3.3. Validation step

This study was composed of five simulations. Thus, each simulation was separately validated. In general, the validation parameters in the HYSYS simulation depend on the output required from each simulation.

3.3.1. Validation of Strippers 1 and 2

The validation of Stripper 1 mainly depends on the product stream parameters and composition. However, it also depends on the reboiler duty and steam required for the reboiler. Table 1 lists the Stripper 1 validation parameters, revealing that the design and simulation parameters are similar. In particular, the deviation between the design and simulation of the reboiler duty was 1.42% and that of the steam flow was 1.56%. Meanwhile, the temperature, pressure, and mass flow were almost the same. No deviation existed in the composition. Note that this is because a suitable simulation package was selected, which herein was "Peng Robinson."

The validation results for Stripper 2 were also similar. The deviation between the design and simulation of the reboiler duty was 3.42%, and that of the steam flow was 1.56%. Meanwhile, the stripped water temperature, pressure, and mass flow were almost the same, and no deviation existed in the composition of the streams (Table 2).

3.3.2. Validation of Regenerators 1 and 2

The selected parameters to validate Regenerators 1 and 2 were the same as those for Strippers 1 and 2. In general, the design and simulation results were similar. Table 3 summarizes the validation results of Regenerator 1. The reboiler duty deviation was −5.21%. The steam flow deviation was −6.14%, and the 141 stream pressure deviation was −3.85%. Meanwhile, the temperature and flow were almost the same. However, a large deviation appeared for the H_2S in the LA solution. Nevertheless, this is not a problem because less H_2S in the LA solution is better, and the comparison values used are very low. In addition, the most important items in the validation are the reboiler duty and steam flow.

Table 1 Stripper 1 validation.

Reboiler duty (MMkCal/h)	Design	Simulation	Deviation (%)
	11.27	11.11	1.42
Steam flow (kg/h)	Design	Simulation	Deviation (%)
	22,231	21885.18	1.56
Stream	**121 (Stripped water)**		
Property	Design	Simulation	Deviation
Temperature (°C)	125.9	125.8	0.08
Pressure (kg/cm^{2}g)	1.4	1.4	0.00
Mass flow (kg/h)	122802.7	123150.2	−0.28
Component	**Total weight component fraction**		
H_2O	1.0	1.0	0.0
NH_3	0.0	0.0	0.0
H_2S	0.0	0.0	0.0
Phenol	0.0	0.0	0.0
CO_2	0.0	0.0	0.0
Diethanolamine (DEA)	0.0	0.0	0.0

Table 2 Stripper 2 validation.

Reboiler duty (MMkCal/h)	Design	Simulation	Deviation (%)
	3.51	3.39	3.42
Steam flow (kg/h)	Design	Simulation	Deviation (%)
	22,231	21885.18	1.56
Stream	**221 (stripped water)**		
Property	Design	Simulation	Deviation
Temperature (°C)	125.9	125.8464	0.04
Pressure (kg/cm^{2}g)	1.4	1.4	0.00
Mass flow (kg/h)	38506.6	38713.79	−0.54
Component	**Total weight component fraction**		
H_2O	1.0	1.0	0.0
NH_3	0.0	0.0	0.0
H_2S	0.0	0.0	0.0
Phenol	0.0	0.0	0.0
CO_2	0.0	0.0	0.0
Diethanolamine (DEA)	0.0	0.0	0.0

Table 3 Regenerator 1 validation.

Reboiler duty (MMkCal/h)	**Design**	**Simulation**	**Deviation (%)**
	32.86	34.57	−5.21
Steam flow (kg/h)	Design	Simulation	Deviation (%)
	64,627	68594.9	−6.14
Stream	**141 (Lean Amine)**		
Property	Design	Simulation	Deviation
Temperature (°C)	127.00	127.14	−0.11
Pressure (kg/cm^{2}g)	1.30	1.35	−3.85
Mass flow (kg/h)	486,224	485,796	0.09
Component	**Total Weight component fraction**		
Diethanolamine (DEA)	0.250	0.250	−0.06
Water	0.748	0.749	−0.13
H_2	0.000	0.000	0.00
NH_3	0.000	0.000	0.00
H_2S	0.002	0.001	43.89
CO_2	0.000	0.000	0.00

For Regenerator 2, all the selected parameters were similar between the design and simulation values, except for the H_2S concentration in the LA solution. However, for the same reason previously mentioned, this is not a problem and does not affect the reboiler and steam flow validation. The suitable selected package for Regenerators 1 and 2 was "Chemical Solvent." Table 4 lists the Regenerator 2 validation results. Normally, the "Peng Robinson"' package is suitable for most simulations, but it is not suitable in this case because of the existence of a DEA chemical solvent. Note that an incorrect package selection would completely deviate the results.

3.3.3. Validation of SRU

The SRU is a large plant that utilizes various equipment. Thus, it is pertinent to select suitable parameters to represent the validation of the plant. The aim of the SRU is to prevent AG discharge via the stack and produce sulfur from the AG. Therefore, the two streams selected for validation are the

Table 4 Regenerator 2 validation.

Reboiler duty (MMkCal/h)	Design	Simulation	Deviation (%)
	4.44	4.67	−5.14
Steam flow (kg/h)	Design	Simulation	Deviation (%)
	8732	9262.802	−6.08
Stream	**215 (Lean Amine)**		
Property	Design	Simulation	Deviation
Temperature (°C)	127	127.1476	−0.12
Pressure (kg/cm^{2}g)	1.35	1.35	0.00
Mass flow (kg/h)	71126.8	71126.88	0.00
Component	**Total weight component fraction**		
Diethanolamine (DEA)	0.250	0.250	−0.10
Water	0.748	0.749	−0.13
H_2	0.000	0.000	0.00
.NH_3	0.000	0.000	0.00
H_2S	0.002	0.001	51.81
CO_2	0.000	0.000	0.00

liquid sulfur product and the flue gas to stack streams. Table 5 lists the validation results for the SRU plant. All the results were similar, with the exception of some small deviations. Regarding the liquid sulfur product, there was no deviation present in the results, except for a small deviation of 0.1% between the design and simulation for the flow of the liquid sulfur product. Meanwhile, for the flue gas to the stack stream, the results were also similar. However, a minor deviation of −4.16% was found for the flow. The highest deviation in the composition appeared in the CO_2 exit from the stack, reaching a value of 6.91%.

3.4. General refinery unit assumptions

In Section 3.2, the assumptions for each section are described. Meanwhile, in this section, the general assumptions for all plants are listed as follows:

- All the units operate under steady-state conditions. The start-up and shut-down conditions were not considered in the calculations.
- The average steam cost is 7.6 $/ton. This cost is calculated based on the natural gas fuel price of the boilers and the chemical treatment of the boiler. In practice, the steam cost is calculated daily according to changes in the fuel price and chemical treatment. These values can vary and exhibit small fluctuations.
- The annual cost saved is calculated based on the normal full load conditions of the units as decreasing capacities do not occur often and are the result of larger issues.

4. Results and discussion

4.1. Scenario 1 optimization

The scenario1 optimization is to turn off SWS1 and SWS2. This scenario will save 28.57 t/h of stripping steam used in the reboilers of Strippers 1 and 2. Table 6 summarizes the steam production from the WHB at different SWSAG flows. However, this scenario causes a reduction of 3674 kg of SWSAG/h from the feed to the reaction furnace. Consequently, the high pressure saturated steam produced from the WHB following the reaction furnace will decrease from 35.29 ton/h to 25.57 ton/h, which is a 9.72 ton/h loss. Thus, the total saved steam is equal to the difference between the 28.57 ton/h of the steam saved and 9.72 ton/h of lost steam. Therefore, the total saved steam is 18.85 ton/h.

4.2. Scenario 2 optimization

The scenario 2 optimization strategy was to optimize the steam used in the reboilers of Strippers 1 and 2, and Regenerators 1 and 2. The design S/F ratio for Strippers 1 and 2 is 0.17 kg steam/kg feed. Meanwhile, the optimized ratio is 0.16 kg

Table 5 Sulphur recovery unit validation.

Stream description		Liquid sulfur product			Flue gas to stack		
property	**Unit**	**industrial**	**Simulation**	**% error**	**industrial**	**Simulation**	**% error**
Temperature	°C	135	135	–	652	652	–
Pressure	kg/cm^2 g	0.01	0.01	–	0.01	0.01	–
Flow	kg/h	12,430	12438.06	0.1	41,283	43002.28	−4.16
Component		**mole fraction**					
H_2		0	0	0	0.012	0.011	4.60
H_2O		0	0	0	0.116	0.120	−3.19
CO		0	0	0	0	0.000	0.00
N_2		0	0	0	0.828	0.819	1.09
O_2		0	0	0	0.02	0.020	0.00
CO_2		0	0	0	0.025	0.023	6.91
H_2S		0	0	0	0	0.000	0.00
SO_2		0	0	0	0	0.000	0.00
COS		0	0	0	0	0.000	0.00
CS_2		0	0	0	0	0.000	0.00
Liquid S		1	1	0	0	0.000	0.00
NH_3		0	0	0	0	0.000	0.00

Table 6 Sulfide recovery unit (SRU) waste heat boiler (WHG) steam production at different sour water stripped acid gas (SWSAG) flows.

SWSAG (kg/h)	High pressure saturated steam (ton/h)	WHB duty (kJ/h)	WHB power (kW)
3674.00	35.29	82307372.21	22863.16
2755.50	32.89	76704169.85	21306.71
1837.00	30.45	71012573.21	19725.71
918.50	28.01	65315510.11	18143.20
0.00	25.57	59620656.43	16561.29

steam/kg feed. The design S/F ratio for Regenerators 1 and 2 is 0.14 kg steam/kg feed, whereas the optimized ratio is 0.10 kg steam/kg feed. The stripped water and LA solution analyses were over-specified. The laboratory results were less than 1 ppm-Wt. for H_2S and less than 10 ppm-Wt·NH_3. Note that the required limits are less than 10 ppm-Wt. H_2S and less than 50 ppm-Wt·NH_3. Further, the LA was always less than 0.05 wt %. The optimized S/F ratio was based on monitoring lab results to maintain the specifications of the stripped water and LA solutions within the required limits. The calculated values ranged from 50% to 100% unit capacities. Below this ratio, the unit is unstable and cannot operate the normal or optimized conditions.

4.2.1. Stripper 1 normal and optimized results

The required steam flow increased in a linear proportional relationship from 50% to 100% capacity. The S/F calculations under normal conditions are listed in Table 7.

At 100% capacity, the required steam is 21885.18 kg/h, while at 50% capacity, it is 10942.59 kg/h. The division between steam and feed remains constant under normal conditions at 0.17 kg steam/kg feed, even at different loads. The optimized steam calculations, including the steam saved at different loads, are listed in Table 8. The saved steam at the full load capacity is 1.71 ton/h. As the new optimized S/F ratio is 0.16, each steam flow value under normal condition was multiplied by 0.16 to compute the optimized rates.

4.2.2. Stripper 2 normal and optimized results

The Stripper 2 normal conditions are listed in Table 9. Note that the flow to Stripper 1 is much higher than that of Stripper 2 under full capacity conditions because it handles the sour water of all the units, except for one unit (DCU), which is handled by Stripper 2.

The division between the steam and feed remains constant under normal conditions, maintaining 0.17 kg steam/kg feed at different loads. The Stripper 2 optimized conditions are listed in Table 10. Overall, the saved steam at full capacity is 0.45 ton/h.

For Stripper 2, the optimized S/F ratio is 0.16. Thus, each steam flow under normal conditions was multiplied by 0.16 to compute the optimized rates.

4.2.3. Regenerator 1 normal and optimized results

The steam used at full capacity for Regenerator 1 is 496025.80 kg/h. In a normal reboiler, the feed flow to the towers exhibits a directly proportional relationship between the reboiler duty and reboiler power. The division between the steam and feed remains constant under normal conditions, maintaining 0.14 kg steam/kg feed at different loads. The Regenerator 1 normal condition results are listed in Table 11. The steam flow used to regenerate each amine solution at full capacity was 68585.38 kg/h. The steam flow decreases with lower loads.

The steam flows of Regenerator 1 under optimized conditions are listed in Table 12. The saved steam at full capacity is 18.98 ton/h, which is significant. Any decrease in steam should be examined using a laboratory analysis of the LA solution to check that the H_2S content does not exceed 2 wt%.

As the optimized S/F ratio is 0.1, multiplying each steam flow under normal conditions by 0.1 was performed to determine the new steam flow optimized rates. Herein, we used Regenerator 1 at a full load as an example to differentiate between normal and optimized conditions. The main difference and key factor was working with lower S/F ratios in the four units. Usually, this ratio is constant at different loads. For example, for Regenerator 1 the division between steam and feed at 100%, 75%, and 50% loads is constant in the conventional process (0.14 kg steam/kg feed). As the feed at a 100% load is 496025.8 kg/h, the required steam for normal conditions is 496025.8 kg/h × 0.14 = 68585.38 kg/h. Similarly, under the new optimized ratio of 0.1 kg steam/kg feed is constant at different loads. Therefore, the newly calculated required steam for the reboiler at a 100% load is 49602.58 kg/h (496025.8 feed kg/h × 0.1 = 49602.58 steam kg/h), and the saved steam is 19980 kg/h (68585.38 kg/h − 49602.58 kg/h). However, in the plant in the distributed control system, a ratio controller computes the steam required for each load, and the new ratio of 0.1 is inserted into the controller to compute the required new steam value.

4.2.4. Regenerator 2 normal and optimized results

The feed of Regenerator 1 is 496025.80 kg/h, while the feed of Regenerator 2 is 73176.40 kg/h at a full load because the LA solution of all the units is regenerated in Regenerator 1, except for one unit (DCU), which is regenerated in Regenerator 2. The results of Regenerator2 normal conditions are listed in Table 13. The required steam at normal operation to regenerate the RA solution is 9262.80 kg/h.

Generally, in the towers, steam optimization should be conducted gradually by monitoring the column temperature and pressure profiles. The results of the Regenerator 2 optimized conditions are listed in Table 14. The steam saved for Regenerator 2 is 1.95 ton/h. This decrease in steam is related to the laboratory analysis of the LA solution.

4.3. Feed and steam relationship in towers

The feed relationships with the reboiler steam, reboiler duty, and reboiler power are linear directly proportional relationships. Fig. 8 shows the relationship between steam and feed for Regenerator 1 as an example. Note only one example was provided to minimize redundancy.

The ratio between steam rate and feed rate remains constant at different loads at a value of 0.1 kg/kg under optimized conditions.

The relationship between the feed flow and reboiler duty from a 50% load to a full load is shown in Fig. 9. Understanding

Table 7 Stripper 1 steam flows at different capacities under normal conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Q_Reboiler (kJ/h)	Power Reboiler (kW)	Steam/feed ratio (kg/kg)
126066.30	100.00	21885.18	46505624.23	12918.23	0.17
110308.01	87.50	19149.53	40692421.19	11303.45	0.17
94549.73	75.00	16413.89	34879218.17	9688.67	0.17
78791.44	62.50	13678.24	29066015.14	8073.89	0.17
63033.15	50.00	10942.59	23252812.11	6459.11	0.17

Table 8 Stripper 1 steam flows at different capacities under optimized conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Steam flow Optimized (kg/h)	Saved steam (kg/h)	Saved steam (ton/h)	Steam/feed ratio (kg/kg)
126066.30	100.00	21885.18	20170.61	1714.57	1.71	0.16
110308.01	87.50	19149.53	17649.28	1500.25	1.50	0.16
94549.73	75.00	16413.89	15127.96	1285.93	1.29	0.16
78791.44	62.50	13678.24	12606.63	1071.61	1.07	0.16
63033.15	50.00	10942.59	10085.30	857.29	0.86	0.16

Table 9 Stripper 2 steam flows at different capacities under normal conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Q_Reboiler (kJ/h)	Power Reboiler (kW)	Steam/feed ratio (kg/kg)
39000.00	100	6686.96	14209680.21	3947.13	0.17
34125.00	87.5	5851.09	12433470.19	3453.74	0.17
29250.00	75	5015.22	10657260.16	2960.35	0.17
24375.00	62.5	4179.35	8881050.13	2466.96	0.17
19500.00	50	3343.48	7104840.11	1973.57	0.17

Table 10 Stripper 2 steam flows at different capacities under optimized conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Steam flow Optimized (kg/h)	Saved steam (kg/h)	Saved steam (ton/h)	Steam/feed ratio (kg/kg)
39000.00	100	6686.96	6240.00	446.96	0.45	0.16
34125.00	87.5	5851.09	5460.00	391.09	0.39	0.16
29250.00	75	5015.22	4680.00	335.22	0.34	0.16
24375.00	62.5	4179.35	3900.00	279.35	0.28	0.16
19500.00	50	3343.48	3120.00	223.48	0.22	0.16

this relationship is of great importance for performing optimization scenarios (see Fig. 9).

The relationship between the feed flow and reboiler power from a 50% to 100% load is shown in Fig. 10. This relationship is important when compared with equipment that uses power as the main unit, such as electrical motors, compressors, and turbines, or when making calculations relating to power units. The ratio between the reboiler duty and feed will remain constant at 291.7 kJ/kg for different loads.

The ratio between the reboiler power and feed will remain constant at 0.08 kWh/kg under different loads.

4.4. Selected scenario

Overall, the total saved steam for scenarios 1 and 2 were 18.85 ton/h and 23.09 ton/h, respectively. Any steam saved indicates a reduction in the steam amount produced from the boilers. Moreover, if scenario 1 was implemented, all the H_2S and NH_3 existing in the sour water would exit

Table 11 Regenerator 1 steam flows at different capacities under normal conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Q_Reboiler (kJ/h)	Power Reboiler (kW)	Steam/feed ratio (kg/kg)
496025.80	100.00	68585.38	144692306.69	40192.31	0.14
434022.58	87.50	60012.21	126605768.36	35168.27	0.14
372019.35	75.00	51439.03	108519230.02	30144.23	0.14
310016.13	62.50	42865.86	90432691.68	25120.19	0.14
248012.90	50.00	34292.69	72346153.35	20096.15	0.14

Table 12 Regenerator 1 steam flows at different capacities under optimized conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Steam flow Optimized (kg/h)	Saved steam (kg/h)	Saved steam (ton/h)	Steam/feed ratio (kg/kg)
496025.80	100	68585.38	49602.58	18982.80	18.98	0.10
434022.58	87.5	60012.21	43402.26	16609.95	16.61	0.10
372019.35	75	51439.03	37201.94	14237.10	14.24	0.10
310016.13	62.5	42865.86	31001.61	11864.25	11.86	0.10
248012.90	50	34292.69	24801.29	9491.40	9.49	0.10

Table 13 Regenerator 2 steam flows at different capacities under normal conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Q_Reboiler (kJ/h)	Power Reboiler (kW)	Steam/feed ratio (kg/kg)
73176.40	100.00	9262.80	19541426.51	5428.17	0.13
64029.35	87.50	8104.93	17098712.91	4749.64	0.13
54882.30	75.00	6947.08	14656018.45	4071.12	0.13
45735.25	62.50	5789.24	12213359.48	3392.60	0.13
36588.20	50.00	4631.39	9770688.16	2714.08	0.13

Table 14 Regenerator 2 steam flows at different capacities under optimized conditions.

Feed flow (kg/h)	Capacity %	Steam flow (kg/h)	Steam flow Optimized (kg/h)	Saved steam (kg/h)	Saved steam (ton/h)	Steam/feed ratio (kg/kg)
73176.40	100	9262.80	7317.64	1945.16	1.95	0.10
64029.35	87.5	8104.93	6402.94	1702.00	1.70	0.10
54882.30	75	6947.08	5488.23	1458.85	1.46	0.10
45735.25	62.5	5789.24	4573.53	1215.71	1.22	0.10
36588.20	50	4631.39	3658.82	972.57	0.97	0.10

the bottom of the tower and be directed to the wastewater treatment unit. However, this study proved that this unit is unable to handle H_2S and NH_3 in the wastewater. Further, the pipes designed for stripped water are not capable of handling acidic solutions, whereas the normal pipes at Strippers 1 and 2 are designed to handle acidic gases. Based on these findings, scenario 1 was deemed unsuitable because it would damage the system. Thus, scenario 2 was implemented.

4.5. Cost saving

The cost-saving calculations depend on the cost of one ton of steam ($/ton). The average steam cost for 1 ton is approximately 7.6 $/ton. The main influencers of this cost are the natural gas fuel used in the boilers and the chemical treatment for the boilers. Table 15 lists the costs saved from each unit as well as the total saved cost. Note that these costs may vary according to the daily price of natural gas.

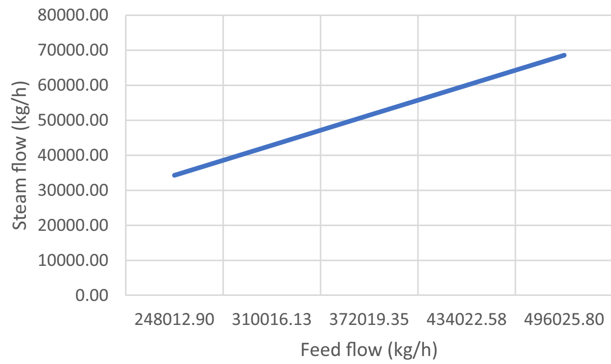

Fig. 8 Regenerator 1 feed to steam relationship.

As shown in Fig. 11, the cost saving contributions of ARU1, ARU2, SWS1, and SWS2 are 82.21%, 8.42%, 7.43%, and 1.94%, respectively.

Table 15 Costs saved using scenario 2.

Unit	Saved steam (ton/h)	Avg. steam cost ($/ton)	Saved cost ($/year)
Sour water steam unit 1 (SWS1)	1.71	7.60	114,149.43
SWS2	0.45		29,757.06
Amine regenerator unit 1 (ARU1)	18.98		1,263,798.81
ARU2	1.95		129,501.08
Total	23.09		1,537,206.38

5. Summary and conclusions

In this study, different scenarios for optimizing the energy consumption of a refining plant that began official production in 2020 were considered. The refining plant uses an SRU to recover sulfur from H_2S with 99.9% sulfur recovery efficiency. The AG feed to the SRU contains mostly H_2S and NH_3 and is

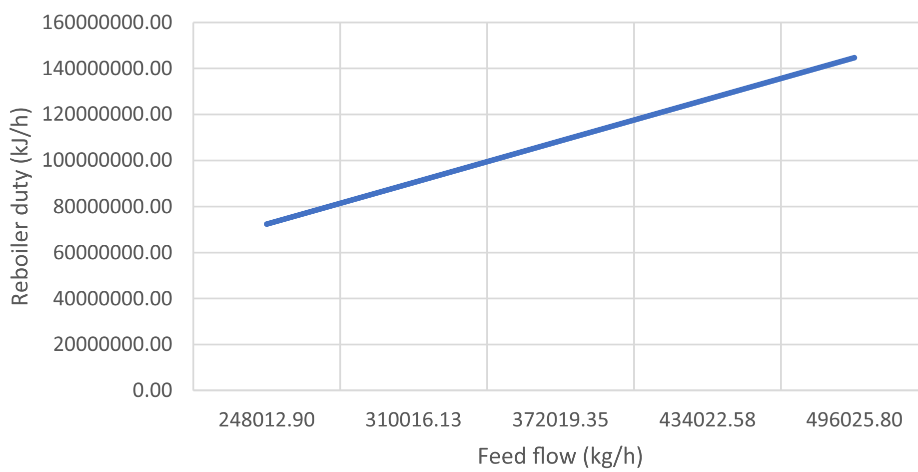

Fig. 9 Feed flow and reboiler duty relationship.

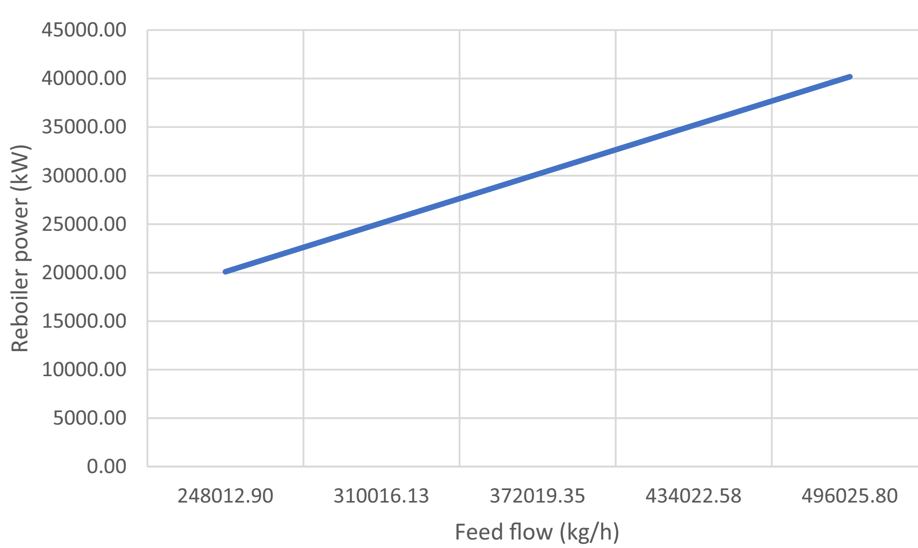

Fig. 10 Feed flow and reboiler power relationship.

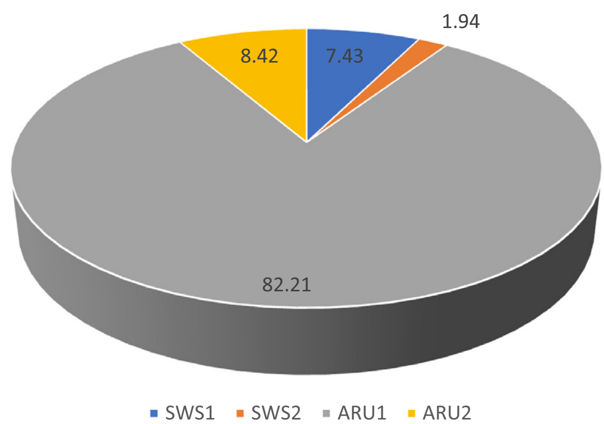

Fig. 11 Contributions of each unit to total cost saved.

provided by SWS1, SWS2, ARU1, and ARU2. Thus, the selected units for optimization were SWS and ARU. Four columns were simulated with using Aspen HYSYS V.11 (Stripper 1 of SWS1, Stripper 2 of SWS2, Regenerator 1 of ARU1, and Regenerator 2 of ARU2), and correspondingly four simulations were conducted to determine the amount of steam required for each column from 50% to 100% loads. A turndown ratio below 50% was not included in our study as steady-state operations were assumed. The SRU has a direct relationship with these units because the overhead acidic gas from the four columns feeds the SRU. Therefore, an additional simulation was conducted for the SRU using a special package in HYSYS named SULSIM. Overall, two scenarios were considered for energy optimization. The first scenario was to stop use of the SWS1 and SWS2 units to save 28.57 ton/h of the steam required for stripping in Strippers 1 and 2. In this scenario, the AG feed to the SRU unit is provided only from the ARU1 and ARU2 units. By eliminating AG production from SWS1 and SWS2, the feed to the furnace was reduced by 3674 kg/h, which equates to a 23.48% decrease, as compared with normal operating conditions. In addition, the high pressure steam produced from the WHB following the reaction furnace will decrease from 35.29 ton/h to 25.57 ton/h. Further, the SRU simulation computed the steam produced from the WHB without a SWSAG feed. The total decrease in steam production was 27.54%. Thus, the boiler must increase steam production by 9.72 ton/h to compensate for this decrease. A CHP model based on daily actual steam and power consumption in the plant was also used. The total steam saved from scenario 1 was 18.85 ton/h. However, if scenario 1 is implemented, all the H_2S and NH_3 in the sour water will exit the bottom of the tower and be directed to the wastewater treatment unit, which cannot treat large amounts of H_2S and NH_3 and exposes the pipelines at the bottoms of Strippers 1 and 2 to as only the pipes at the top of the towers can handle acidic conditions. Consequently, scenario 1 was deemed unsuitable for implementation Meanwhile, scenario 2 considered optimizing the steam used in the four columns. Normally, column reboilers use a stable calculated S/F ratio. For this scenario, the S/F ratio in the SWS reboilers was optimized from 0.17 to 0.16 kg of steam/kg feed and from 0.14 to 0.10 kg steam/kg feed for the ARU reboilers. The saved steam flows from Strippers 1 and 2, and Regenerators 1 and 2 were 1.71 ton/h, 0.45 ton/h, 18.98 ton/h, and 1.95 ton/h, respectively. Thus, the total saved steam flow was 23.09 ton/h. As the average steam cost is 7.6 $/ton, the saved cost from ARU1 was 1,263,798.81 $/year, contributing 82.21% of total cost saved, that from ARU1 was 129,501.08 4/year, which is 8.42% of the total saved cost, that from SWS1 was 114,149.43 $/year, which is 7.43% of the saved cost, and that from SWS2 was 29,757.06 4/year, which contributes only 1.94% to the total cost saving. Overall, the total saved cost from all the units was 1,537,206.38 $/year.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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