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Two-Column Sour Water Stripping without Sending Ammonia to The Sulfur Recovery Unit

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Abstract

In conventional sour water stripping (SWS) in petroleum refineries, H_2S and NH_3 of the sour water are separated simultaneously at the top of a tray column by sending sour gases to the sulfur recovery unit (SRU). At SRU furnaces, ammonia is oxidized to NOx playing the role of a deactivator for downstream catalytic beds. From an industrial engineering viewpoint, it is proposed to strip ammonia and hydrogen sulfide separately using two stripper columns. The striped ammonia may be incinerated if it is allowed or traded as gaseous or aqueous ammonia. In this study, the two-column sour water stripping approach is implemented for the feed stream of an SWS unit currently working in Iran using a simulation model that was verified based on the design data of an industrial unit. It is concluded that the higher the pressure of the H_2S stripper is, the higher the H_2S recovery is, and the lower H_2S impurity with ammonia stream is. Moreover, preheating of the feed stream and sending a lower portion of cold feed stream as a representative of liquid reflux resulted in increased utility consumption and also changed the amount of water sent with sour gas and ammonia stream. Finally, aqueous ammonia (less than 10% purity) or ammonia vapor stream (about 90% purity) can be traded as a value-added product for this unit.

Key words: Sour Water Stripping, Sour Gas, Ammonia, Stripped Water, Process Simulation.

Introduction

In some of the important processes of a petroleum refinery, like crude desalting, crude distillation, catalytic cracking, etc., the presence of components like H_2S and ammonia in wastewater makes it pretreated to remove these pollutants before disposal to the environment [1]. The wastewater containing H_2S and ammonia is called sour water, and sour water stripping is one of the four major pretreatment processes considering effluent within a refinery. The heart of the unit is the stripping column, while the ascending flow of stripping steam or gas removes the pollutants of sour water on each tray of the column [1]. Nowadays, energy and economic analyses of this process are even taken into consideration to improve the performance of the sour water units in the refineries [2].

In sour water stripping units (SWS), the number of columns is based on the objective of the separation [1,3,4]. Conventionally, gaseous ammonia along with hydrogen sulfide is sent to sulfur recovery unit (SRU). In contrast, ammonia is oxidized to NOx at SRU furnaces, either may deactivate downstream catalytic beds or may cause a problem with the size of furnaces to reach the designed outlet target of SO_2 production [3,5]. A maximum allowable of 25 wt.% is considered for the ammonia amount of typical sour gas routed to SRU [6]. Not sending ammonia to SRU, it has been industrialized using two columns in SWS [1,7]. H_2S is striped at the top of the first column, called H_2S stripper, then ammonia is removed from the water in the second column, called NH_3 stripper [3,8]. However, alternative methods are also presented to recover ammonia separately [9].

An SWS unit currently working in Iran sends the ammonia amount of 30 wt% to SRU, higher than the maximum allowable ammonia amount [10]. Gaseous H_2S and NH_3 are stripped from water by using a reboiler or direct steam injection for stripper tray columns. As alternatives for external condenser at the top of the column, a pump-around or cold feed injection on the top tray may be selected [3]. In the industrial single-column case, a top pump-around is used for column condensing systems, and a reboiler is used for supplying the continuous vapor stream of the column. Verification of the sour PR model to accurately predict desorption of H_2S, NH_3 for the industrial case, the feed stream characteristic given in Table 1, is investigated in detail elsewhere [10,11]. It is observed that the simulation results are in good agreement with the design data.

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Table 1: Characteristic of sour water stripper feed.

Source	value
Temperature (°C)	78
Pressure (bar_g)	1.4
Mass Flow (kg/h)	127100
H_2O (wt frac.)	0.964
H_2S (wt frac.)	0.023
NH_3 (wt frac.)	0.012

In this study, the two-column sour water stripping approach by stripping ammonia separately from H_2S to avoid sending ammonia to SRU is implemented for a real feed stream of the SWS unit. The simulation model, verified elsewhere by the design data of the SWS unit currently working in Iran, is used to perform the case studies based on different operational parameters. Then, the optimum operating condition following product streams specifications and utility consumption is investigated.

Materials and Methods

Two-Column Simulation Approach

In Figure 1, the scheme of the two-column sour water striping configuration is shown. The feed stream is usually preheated before the entrance to H_2S stripper, first column, by heat exchanging with the two hot streams extracted from the bottom of the two strippers. It will lessen the duty of the H_2S stripper reboiler. Since there is no condensing system at the overhead of H_2S stripper, a portion of the cold feed stream enters into the column on the top tray.

The preheated feed temperature, the molar ratio of cold feed, and the pressure of the column are operational parameters that affect the column performance and utility consumption. To separate ammonia at the bottom of H_2S stripper, column top pressure has to be more than a conventional single column where H_2S and NH_3 are co-stripped at the top. Moreover, it is believed that the higher the overhead pressure of H_2S stripper is, the higher the recovery of H_2S at the overhead stream is, called sour gas will be.

As seen in Figure 2, the bottom stream of H_2S stripper passes through the pre-heat-exchanger-2 to decrease its temperature and then enters to NH_3 stripper. At NH_3 stripper, the objective is to lessen the NH_3 amount of water less than 30 ppmw and separate ammonia as a byproduct from the top. It is considered that the saturated overhead stream passes through a water cooler producing saturated aqueous stream as a rich ammonia product. However, the hydrogen sulfide that leaves the H_2S stripper bottom is stripped at the top of NH_3 stripper as an impurity with ammonia product stream.

Stripped water at the bottom of NH_3 stripper is cooled passing through pre-heat-exchanger-1, as shown in Figure 1, and finally, a water cooler is set the stream temperature for battery limits. A 10-tray column with no condensing system is considered for H_2S stripper. Preheated and cold feed streams are entered into the 9th and 10th trays, respectively. Low-pressure steam (LPS), with a temperature of 210 °C at the pressure which is equal to 4 bar_g, is used as the hot fluid in the H_2S stripper reboiler.

The main specification of the H_2S stripper is to recover more than 0.9995 of ammonia at the bottom, not routing to SRU with sour gas. At the same time, it is preferred to obtain the highest probable H_2S recovery at the top.

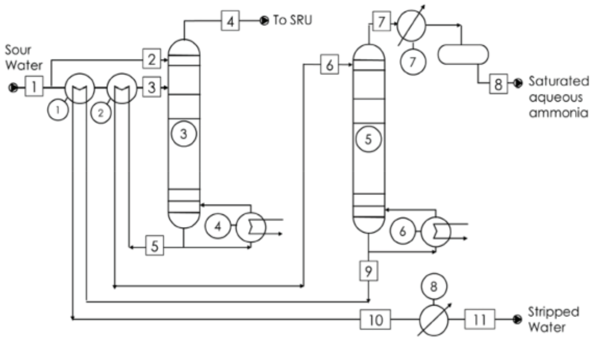

Fig. 1 Scheme of the two-column sour water striping consists of (1) pre-heat-exchanger-1, (2) pre-heat-exchanger-2, (3) H_2S stripper, (4) H_2S stripper reboiler, (5) NH_3 stripper, (6) NH_3 stripper reboiler, (7) am-monia cooler, and (8) stripped water cooler.

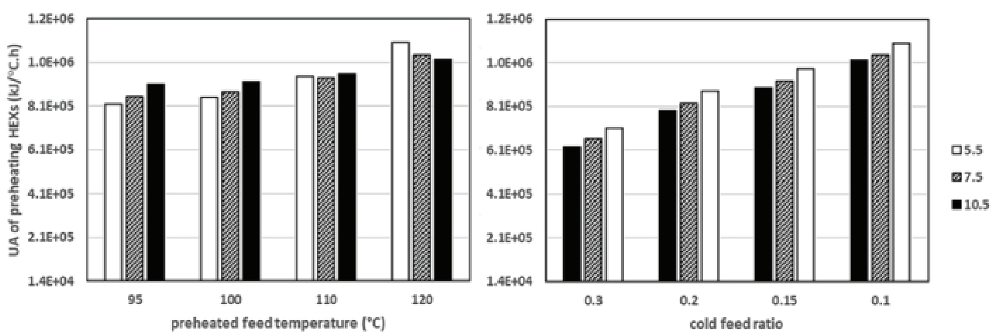

Fig. 2 UA changes vs. preheated feed temperature and cold feed ratio of H_2S stripper.

H_2S recovery and LPS consumption are studied in terms of operational parameters, including the preheated feed temperature, the molar ratio of cold feed, and the pressure of the column. Here, the range of theses parameters are taken 95-120 °C, 0.1-0.3, and 5.5-10.5 bar_g, respectively.

A 15-tray column with no condensing system is considered for NH_3 stripper while the feed stream is entered on the top tray. Also, NH_3 stripper top pressure is kept constant at 1.7 bar_g. The maximum allowable ammonia amount in the bottom stream of NH_3 stripper is usually considered as 30 ppmw according to environmental regulations. NH_3 cooler produced saturated ammonia products and stripped the water cooler to use cooling water (CW) as a coolant at 35 °C.

The economy of this process is based on the quality of the products. For sour gas, it is necessary that NH_3 not be sent to SRU at first, and be decreased in water amount at the possible level. For ammonia products, obtaining high ammonia purity as well as fewer amounts of H_2S at the top of NH_3 stripper is favorable. To avoid sending much more water along with the ammonia product, a condensing system may be installed with a reflux stream based on the controlling column temperature. In addition, NH_3 stripper performance is investigated for a top temperature range of about 55-95 °C.

Results and Discussion

Characteristics of the feed stream used in the process simulation are given in Table 1 (i.e., stream 1 in Figure 1).

Two-Column with no Condensing System at Top

In this simulation approach, no condensing system is considered for both columns, and therefore, case studies are considered based on three operational parameters. The first parameter is top pressure of H_2S stripper (equipment noted as 3 in Figure 1), the second parameter is the temperature of the feed stream inserted to H_2S stripper (stream noted as 3 in Figure 1), and the last parameter is the split ratio of sour water stream divided into cold streams and preheated feed stream (mas flow portion of streams 2 to 3, as seen in Figure 1).

The effect of top pressure adjustment in H_2S stripper, preheating its feed stream and using a portion of sour water stream as liquid reflux in the first stripping column on specifications of outlet stream and utility consumption is investigated. However, this also indirectly changes some specification of NH_3 stripper.

The main objective of the separation process is optimizing utility consumption as well as obtaining desirable recovery of H_2S and ammonia from sour water in a two-column SWS unit producing stripped water.

In Figure 2, the effect of preheated feed temperature and cold feed ratio on the UA of pre-heat exchangers are shown. U is the overall heat exchanger coefficient, and A is the area of heat exchanging. By setting the feed temperature of H_2S stripper and lowering the preheated feed flow as a result of inserting a portion of cold feed separately to the column, the preheat-exchanger design is affected. It is observed that obtaining higher feed temperature and sending a lower portion of the cold feed to the column result in the more required area. However, increasing the flow passes through heat exchangers shows a more significant effect in comparison with the one of outlet temperature.

The effects of preheated feed temperature on column specification of H_2S and NH_3 strippers are given in detail in Tables 2 and 3. Although preheated feed temperature and cold feed ratios affect the utility consumption as given in Table 2, the reboiler temperature is calculated based on the ammonia recovery set for the bottom of the column. As utility consumption, low-pressure steam (LPS) is related to both H_2S and NH_3 stripper reboilers, and cooling water (CW) is related to the ammonia cooler and the stripped water cooler. As seen in Table 2, there is a trade-off between LPS consumption of H_2S and NH_3 stripper separately.

Table 2. Simulation results of utility consumption vs. preheated feed temperature and top pressure of H_2S stripper.

Pressure of H_2S stripper (bar_g)	5.5				7.5				10.5			
Preheated feed temperature to H_2S stripper (°C)	120	110	100	95	120	110	100	95	120	110	100	95
Bottom stream temperature from H_2S stripper (°C)	157.9	158.0	158.0	158.0	168.7	168.7	168.7	168.7	181.5	181.5	181.5	181.5
Feed stream temperature to NH_3 stripper (°C)	129.4	130.0	130.4	130.6	130.1	130.5	130.8	131.0	130.7	131.0	131.2	131.3
Top stream temperature from NH_3 stripper (°C)	126.9	127.2	127.5	127.6	127.3	127.6	127.8	127.9	127.7	127.9	128.1	128.1
Bottom stream temperature from NH_3 stripper (°C)	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3
Reboiler duty of H_2S stripper (kJ/h) × 10^(-7)	2.515	3.023	3.525	3.775	3.135	3.643	4.144	4.392	3.886	4.393	4.897	5.139
LPS consumption of H_2S stripper reboiler (kg/h) × 10^(-4)	1.120	1.347	1.570	1.6813	1.396	1.622	1.846	1.956	1.731	1.957	2.181	2.289
Reboiler duty of NH_3 stripper (kJ/h) × 10^(-7)	3.644	3.535	3.435	3.385	3.504	3.404	3.312	3.266	3.352	3.259	3.170	3.127
LPS consumption of NH_3 stripper reboiler (kg/h) × 10^(-4)	1.623	1.575	1.530	1.508	1.561	1.516	1.475	1.455	1.493	1.451	1.412	1.393
CW consumption of ammonia gas cooler (kg/h) × 10^(-6)	0.917	1.013	1.110	1.158	1.033	1.131	1.230	1.279	1.178	1.278	1.377	1.426
CW consumption of stripped water cooler (kg/h) × 10^(-6)	0.383	0.368	0.352	0.345	0.365	0.349	0.334	0.326	0.342	0.327	0.311	0.303
Total LPS consumption (kg/h) × 10^(-4)	2.744	2.921	3.100	3.190	2.957	3.139	3.321	3.411	3.224	3.408	3.593	3.682
Total CW consumption (kg/h) × 10^(-6)	1.300	1.381	1.463	1.503	1.398	1.480	1.563	1.605	1.520	1.604	1.687	1.730

Table 3: Simulation results of mass balance vs. preheated feed temperature and top pressure of H_2S stripper.

Pressure of H_2S stripper (bar_g)	5.5				7.5				10.5			
Preheated feed temperature to H_2S stripper (°C)	120	110	100	95	120	110	100	95	120	110	100	95
Sour gas												
mass flow (kg/h)	2815	2819	2825	2816	2865	2866	2859	2853	2907	2900	2895	2891
H_2O (wt frac.)	0.026	0.024	0.023	0.022	0.023	0.021	0.020	0.019	0.019	0.017	0.016	0.016
H_2S (wt frac.)	0.974	0.975	0.977	0.977	0.977	0.978	0.98 0	0.981	0.981	0.982	0.984	0.984
NH_3 (wt frac.)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2O (recovery)	0.001	0.001	0.001	0.001	0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2S (recovery)	0.926	0.928	0.931	0.929	0.945	0.946	0.945	0.944	0.962	0.961	0.961	0.960
NH_3 (recovery)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Ammonia saturated liquid												
mass flow (kg/h)	17367	19188	21022	21941	19508	21369	23245	24182	22203	24100	25992	26942
H_2O (wt frac.)	0.897	0.907	0.916	0.919	0.911	0.919	0.925	0.928	0.924	0.930	0.935	0.937
H_2S (wt frac.)	0.013	0.011	0.010	0.010	0.008	0.007	0.007	0.007	0.005	0.005	0.004	0.004
NH_3 (wt frac.)	0.090	0.082	0.075	0.072	0.080	0.073	0.068	0.065	0.071	0.065	0.060	0.058
H_2O (recovery)	0.127	0.142	0.157	0.164	0.145	0.160	0.176	0.183	0.167	0.183	0.198	0.206
H_2S (recovery)	0.074	0.072	0.069	0.071	0.055	0.054	0.054	0.056	0.038	0.038	0.039	0.040
NH_3 (recovery)	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998
Stripped water												
mass flow (kg/h)	106917	105092	103252	102342	104727	102864	100995	100064	101990	100099	98212	97266
H_2O (wt frac.)	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000
H_2S (wt frac.)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
NH_3 (wt frac.)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2O (recovery)	0.872	0.857	0.842	0.835	0.854	0.839	0.824	0.816	0.832	0.817	0.801	0.794
H_2S (recovery)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
NH_3 (recovery)	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002

A similar performance is observed for CW consumption. When the temperature of the feed stream decreases, the CW consumption of ammonia cooler is increased while the CW consumption of the stripped water cooler is decreased. However, LPS and CW consumption is increased by decreasing feed temperature to H_2S stripper.

As shown in Table 3, there is no significant effect on H_2S recovery through sour gas by preheating the feed stream. On the other hand, top pressure increase from 5.5 to 10.5 bar_g results in H_2S recovery increase from 0.93 to 0.96, obtaining less H_2S impurity with ammonia product stream left the NH_3 stripper from the top. However, it is given in Table 3 that the amount of stripped water is decreased regardless of column top pressure.

Figure 3 shows the mass flow of water sent to SRU with sour gas at the top of the H_2S stripper (left plot), and the mass flow of water leaves the top of NH_3 stripper with ammonia product stream (right plot) by the effect of preheated feed temperature. As shown in Figure 3, although preheating the feed stream increases the water flow sent to SRU, ammonia product is concentrated with less water. However, the effect of increasing column pressure is more significant for sending less water to SRU and more water with ammonia products. Of course, lowering the water flow amount to SRU changes sour gas purity based on H_2S.

Moreover, higher top pressure and lower feed temperature result in more water that leaves NH_3 stripper at the top. For example, 12% of water is recovered with the ammonia product at 120 °C of the feed stream and pressure of 5.5 bar g. On the contrary, it is increased up to 20% water recovery at 95 °C of the feed stream and pressure of 10.5 bar g. It results in ammonia purity from 9% to 5.8% with the ammonia stream at the top of NH_3 stripper (find details in Table 3). Case study results based on the effect of cold feed ratios are given in detail in Tables 4 and 5. As seen in Table 4, LPS consumption of NH_3 stripper reboiler is constant to some extent. While for H_2S stripper, it is increased by about 20% for pressure of 10.5 bar_g up to 34% for a pressure of 5.5 bar_g. Moreover, the CW consumption of ammonia cooler and stripped water cooler is also increased. By decreasing the cold feed ratio, more stripped water is produced, and less water is sent to ammonia product stream.

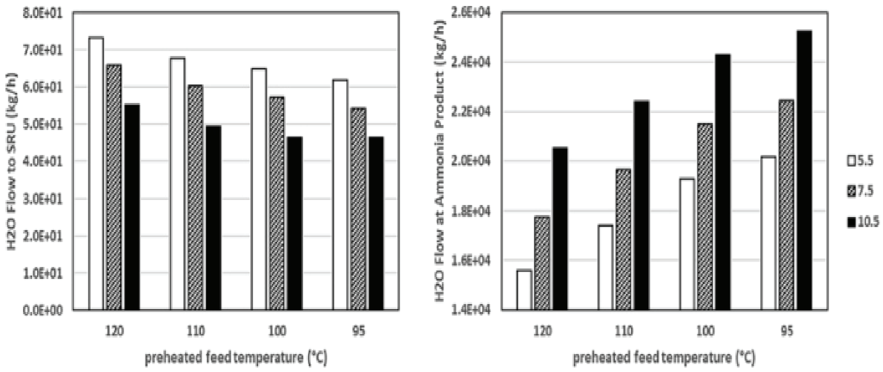

Fig. 3 Mass flow of water at the top of H_2S stripper (left) and at the top of NH_3 stripper (right) vs. preheated feed temperature of H_2S stripper.

Table 4 Simulation results of utility consumption vs. cold feed ratios and top pressure of H_2S stripper.

the pressure of H_2S stripper (bar_g)	5.5				7.5				10.5			
cold feed ratio	0.1	0.15	0.2	0.3	0.1	0.15	0.2	0.3	0.1	0.15	0.2	0.3
bottom stream temperature from H_2S stripper (°C)	157.9	158.0	158.0	158.0	168.7	168.7	168.7	168.7	181.5	181.5	181.5	181.5
feed stream temperature to NH_3 stripper (°C)	129.4	129.5	129.6	129.8	130.1	130.2	130.2	130.4	130.7	130.7	130.8	130.9
Top stream temperature from NH_3 stripper (°C)	126.9	127.0	127.0	127.1	127.3	127.4	127.4	127.5	127.7	127.7	127.8	127.8
bottom stream temperature from NH_3 stripper (°C)	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3	131.3
reboiler duty of H_2S stripper (kJ/h) × 10^(-7)	2.515	2.718	2.921	3.328	3.135	3.337	3.540	3.952	3.886	4.088	4.291	4.697
LPS consumption of H_2S stripper reboiler (kg/h) × 10^(-4)	1.120	1.210	1.301	1.483	1.396	1.486	1.577	1.760	1.731	1.821	1.911	2.092
reboiler duty of NH_3 stripper (kJ/h) × 10^(-7)	3.644	3.622	3.603	3.570	3.504	3.492	3.470	3.438	3.352	3.336	3.318	3.291
LPS consumption of NH_3 stripper reboiler (kg/h) × 10^(-4)	1.623	1.613	1.605	1.590	1.561	1.555	1.546	1.531	1.493	1.486	1.478	1.466
CW consumption of ammonia gas cooler (kg/h) × 10^(-6)	0.917	0.932	0.947	0.980	1.033	1.050	1.065	1.098	1.178	1.194	1.210	1.244
CW consumption of stripped water cooler (kg/h) × 10^(-6)	0.383	0.408	0.434	0.484	0.365	0.390	0.415	0.466	0.342	0.367	0.393	0.443
LPS consumption (kg/h) × 10^(-4)	2.744	2.824	2.906	3.072	2.957	3.042	3.123	3.292	3.224	3.307	3.389	3.558
CW consumption (kg/h) × 10^(-6)	1.300	1.340	1.381	1.464	1.398	1.440	1.481	1.564	1.520	1.562	1.603	1.687

Table 5 Simulation results of mass balance vs. cold feed ratios and top pressure of H_2S stripper.

Pressure of H_2S stripper (bar_g)	5.5				7.5				10.5			
Cold feed ratio	0.1	0.15	0.2	0.3	0.1	0.15	0.2	0.3	0.1	0.15	0.2	**0.3**
Sour gas												
mass flow (kg/h)	2815	2817	2813	2809	2865	2860	2855	2850	2907	2900	2896	2891
H_2O (wt frac.)	0.026	0.024	0.023	0.021	0.023	0.021	0.019	0.018	0.019	0.017	0.016	0.014
H_2S (wt frac.)	0.974	0.976	0.977	0.979	0.977	0.979	0.980	0.982	0.981	0.983	0.984	0.985
NH_3 (wt frac.)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2O (recovery)	0.001	0.001	0.001	0.000	0.001	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2S (recovery)	0.926	0.928	0.928	0.928	0.945	0.945	0.945	0.945	0.962	0.962	0.962	0.961
NH_3 (recovery)	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001
Ammonia saturated liquid												
mass flow (kg/h)	17367	17646	17940	18552	19508	19832	20119	20739	22203	22515	22812	23457
H_2O (wt frac.)	0.897	0.899	0.901	0.904	0.911	0.913	0.914	0.916	0.924	0.925	0.926	0.928
H_2S (wt frac.)	0.013	0.012	0.012	0.011	0.008	0.008	0.008	0.008	0.005	0.005	0.005	0.005
NH_3 (wt frac.)	0.090	0.089	0.088	0.085	0.080	0.079	0.078	0.076	0.071	0.070	0.069	0.067
H_2O (recovery)	0.127	0.129	0.132	0.137	0.145	0.148	0.150	0.155	0.167	0.170	0.172	0.178
H_2S (recovery)	0.074	0.072	0.072	0.072	0.055	0.055	0.055	0.055	0.038	0.038	0.038	0.039
NH_3 (recovery)	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998	0.998
Stripped water												
mass flow (kg/h)	106917	106636	106346	105737	104727	104407	104124	103509	101990	101684	101391	100752
H_2O (wt frac.)	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000
H_2S (wt frac.)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
NH_3 (wt frac.)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
H_2O (recovery)	0.872	0.870	0.868	0.863	0.854	0.852	0.850	0.845	0.832	0.830	0.827	0.822
H_2S (recovery)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
NH_3 (recovery)	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002

In contrast, the maximum probable ammonia purity of 9% is obtained at a feed temperature of 120 °C and a cold feed ratio of 0.1 under top pressure of 5.5 bar_g. On the contrary, more H_2S impurity is observed with ammonia streams in comparison with the pressure of 10.5 bar_g with the same feed temperature and cold feed ratio.

Figure 4 shows the effect of the cold feed ratio on the mass flow of water sent to SRU with sour gas at the top of H_2S stripper (left plot), and the mass flow of water leaves the top of NH_3 stripper with ammonia product stream (right plot). Since the cold feed ratio is directly set the flow rate passed through pre-heat exchangers, it changes the inlet temperature of the stream entering NH_3 stripper. In contrast, preheated feed temperature to H_2S stripper is constant at 120 °C. It may affect water amount with ammonia stream at the overhead of NH_3 stripper and, therefore, different ammonia product purity.

As seen in Table 5, the cold feed ratio does not have a significant effect on H_2S recovery and sour gas flow rate. However, the higher the top pressure of H_2S stripper is, the lower the H_2S impurity with ammonia products is. On the other hand, NH_3 recovery is not changed by an increase in the top pressure of H_2S stripper, and it remains constant at 0.998.

Ammonia Stripper Equipped with Condensing System

In the scenario of two columns with no condensing system at the top, ammonia product is going to be condensed using the CW utility of the unit. In contrast, more water with ammonia stream results in less ammonia purity of the product (less than 10%).

If ammonia stripper is equipped with a condensing system at the top (equipment noted as 7 in Figure 1), setting the top temperature (stream noted as 8 in Figure 1) can improve the purity of ammonia products. In this simulation approach, case studies are considered based on the top temperature of ammonia stripper.

In Figure 5, the ammonia purity of ammonia product in terms of the condenser temperature of NH_3 stripper is shown. The condensing system is used at the top of NH_3 stripper to make ammonia purity higher. On the contrary, the ammonia stream cannot be condensed only with CW utilities, and ammonia product is stable in the gas phase while it contains some H_2S as an impurity.

When H_2S and ammonia are stripped simultaneously, it is possible to meet solid NH_4HS deposition due to high partial pressure of them at the stripper reflux drum [3]. For the two-stripper column configuration, the low partial pressure of H_2S at the ammonia stripper reflux drum results in no risk of any precipitation followed by failure of related instruments. However, the lower temperature imposes the greater duty of reboiler, and the higher top temperature makes the more water vapor exit with the ammonia product stream.

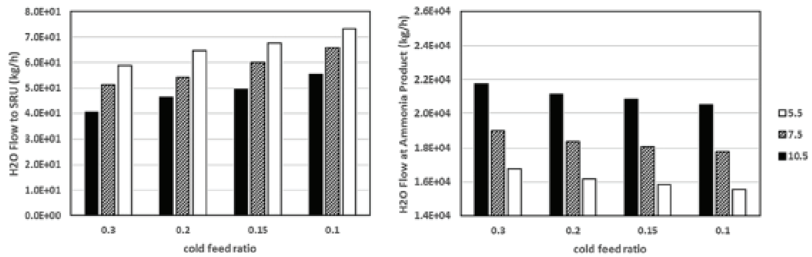

Fig. 4 Mass flow of water at the top of H_2S stripper (left) and at the top of the NH_3 stripper (right) vs. cold feed ratio of H_2S stripper.

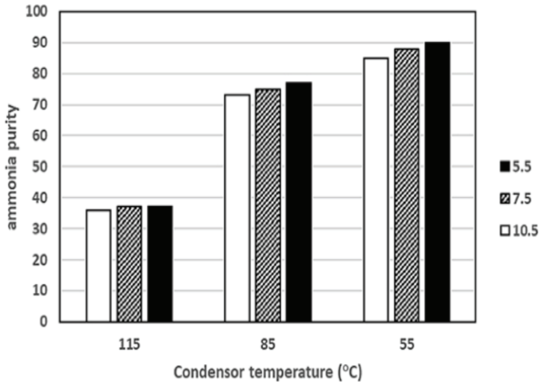

Fig. 5 Ammonia purity of ammonia product vs. condenser temperature of NH_3 stripper and top pressure of H_2S stripper.

If it is necessary, a 10-tray scrubber column can separate H_2S impurity with ammonia streams (overhead of NH_3 stripper). The drain stream which leaves scrubber is sent back to the H_2S stripper column. However, it is then probable to convert it to aqueous ammonia using ammonia scrubbers.

Conclusions

A two-column SWS configuration is simulated for the feed stream of an SWS unit currently working in Iran. The incineration of ammonia in SRU furnaces may be avoided by converting it to gaseous or aqueous ammonia. Of course, it is necessary to decrease the water amount of sour gas at the lower level. Ammonia product is preferred with high ammonia purity and fewer H_2S as an impurity. Two scenarios are studied; the first two columns have no condensing system at the top, and at the latter, a condensing system considered for the NH_3 stripper. H_2S recovery, NH_3 purity, and utility consumption are studied in terms of operational parameters, including the preheated feed temperature, the molar ratio of cold feed, and the pressure of the H_2S stripper. By adding condensers to NH_3 strippers at the top, the temperature of ammonia product leaving NH_3 strippers from the top also affects the product streams specification. It is demonstrated that the higher the pressure of the H_2S stripper is, the higher the H_2S recovery and the lower H_2S impurity with ammonia stream will be. Moreover, preheating feed stream and sending a lower portion of cold feed stream as a representative of liquid reflux increase utility consumption and also change the amount of water sent with sour gas and ammonia streams. The condensing system of NH_3 stripper avoids sending much more water with an ammonia product stream. Ultimately, the implementation of this scenario of two columns of SWS increases investment cost and utility consumption. However, NH_3 removal from SRU feed is an advantage, and aqueous ammonia (less than 10% purity) or ammonia vapor stream (about 90% purity) can be traded as a value-added product for this environmentally benign unit.

Nomenclatures

Guage Pressure: 2-bar_g and bar g
CW: Cooling water
LPS: Low-pressure steam
SRU: Sulfur recovery unit
SWS: Sour water stripping

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