

Experimental study and techno-economical evaluation of Khangiran sour natural gas condensate desulfurization process

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ABSTRACT

There is a serious global concern for removal of different sulfur components from a variety of hydrocarbon products leading to strict regulations to keep sulfur content of various fuels as low as possible. Sour gas condensate production in Khangiran refinery has been tripled in the last 6 years due to the approach of reservoir conditions to the natural gas phase envelope boundary. The significant quantity of sour condensate produced creates several operational and environmental difficulties. To propose a proper process for removal of sulfur species from Khangiran sour condensate, several experiments are performed by considering various scenarios. The oxidative desulfurization process provided best results and is able to decrease the total sulfur content from 8500 ppm to less than 700 ppm by eliminating all hydrogen sulfide and mercaptans and severely reducing other heavy sulfur containing compounds. The odor of the treated condensate was completely improved due to removal of all volatile sulfur components. The preliminary techno-economic evaluation of an industrial scale process is performed based on the experimental results. It is clearly shown that the proposed process is beneficial from both financial and environmental standpoints.

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1. Introduction

The Khangiran sour gas processing plant was originally founded in late 1970s, commissioned in early 80s and expanded in 2004. It consists of 5 sour gas treating units (GTUs) with maximum total capacity of 50MMSCMD, 4 sulfur recovery units with a maximum total sulfur production of 2600 tons per day and 2 condensate topping plants with each receiving 183.6 CMD (1155 bbl/day) sweet condensate and producing 770 SCMD Gas, 5 CMD Solvents, 108 CMD Naphtha, 38 CMD Kerosene and 29 CMD Gas oil at design rate.

This refinery has produced over 153,000 metric tons of sour (73,700 tonne) and sweet (80,000 tonne) gas condensate in 2011. Sour gas condensate production has been tripled in the last 6 years, while, sweet condensate production had an annual growth rate of about 10 percent during the same period. The remarkable increase in the production of sour gas condensate has caused several operational and environmental problems.

Addition of sour condensate to the topping plants feed destabilizes the entire distillation operation due to the excess amount of volatile components such as hydrogen sulfide and light mercaptans. The objective of this research was to provide a practical procedure for sour condensate desulfurization. Evidently, the treated condensate can then be added to the topping plants feed.

Table 1 illustrates the properties and analysis of various sour (S500) and sweet (S300 and S400) condensates produced in the Khangiran gas processing plant. As it can be seen, the S500 condensate has a much higher sulfur content, especially hydrogen sulfide (H₂S) and mercaptans (thiols) compared to other streams (S300 and S400). High H₂S content of sour condensates may lead to appreciable discharges of such lethal gas into atmosphere. Exposure to H₂S can cause fatigue, loss of appetite, headache, irritability, poor memory, dizziness and even death depending on the H₂S concentrations in air (Chou, 2003).

Mercaptans are not as lethal as hydrogen sulfide but they also should be removed as much as possible, especially light species including methyl, ethyl and propyl mercaptans which are still toxic as well as noxious. Removal of the latter components alleviates the corrosion problems, offensive odor and high toxicity (Ghanbari et al., 2003). Employment of deep desulfurization processes are usually inevitable to satisfy the existing standards for total sulfur

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Table 1
Properties and analysis of various Khangiran refinery condensates.

Property	Method	Unit	S-300	S-400	S-500
Sp.Gr. 60/60 oF	D-4052	–	0.7360	0.7493	0.7959
Sulphur Content	D-1266	Wt%	0.05	0.06	0.821
H2S	D-1159	ppm	<1	<1	150
RSH	D-3227	ppm	7	37	456
Flash Point	D-93	°C	<–5	<–5	14
Acidity	D-664	mg KOH/g	0.05	0.12	0.19
Doctor Test			–	–	+
Viscosity @ 100 OF	D-445	c.St.	0.8124	1.062	1.143
Water Content	D-4928	Wt%	<0.02	<0.02	<0.02
Molecular Weight	IP-86	g/mol	120	120	140
Aromatics	D-1319	Vol%	14.4	6.2	19.9
Olefins	D-1319	Vol%	<0.3	<0.3	<0.3
Naphthenic	D-1319	Vol%	–	–	64.1
Paraffinic	D-1319	Vol%	85.6	93.8	16
Calorific Value, Gross	D-240	Kcal/Kg	11333.0	11177	11085
pH			≈5	≈5	4.5
Analysis via GC (Area %)					
Components	S-300	S-400	S-500		
C6	25.9	5.8	1.3		
C7	32.8	8.6	2.6		
C8	20.6	12.1	7.5		
C9	11.6	12.9	18		
C10	6.1	13.1	21		
C11	2.2	11.3	17		
C12	0.7	9.6	12.2		
C13	0.1	7.5	7.5		
C14		5.6	4.0		
C15		4.5	3.0		
C16		3.4			
C17		2.4			
C18		1.6			
C19		0.8			
C20		0.5			
C21		0.3			

content of liquid hydrocarbons (e.g. natural gas condensates) (Chandra Srivastava, 2012).

Various sulfur removal processes have been reported in the literature and are used in numerous oil and gas refineries. A few of these processes which have been used more extensively will be summarized here. "Doctor sweetening" and "Copper sweetening" processes were traditionally used decades ago to transform mercaptans into less harmful disulfides (Kalinowski, 1959; Krause and Tom, 1952). These costly methods were not usually able to reduce the total sulfur content of the sour liquid hydrocarbon fuels. The "Caustic wash" process is a more popular method for sulfur removal from petroleum products, which uses concentrated NaOH solution to remove most of H₂S, light mercaptans and thiophenols from sour liquid fuel (Manihe et al., September (19 81)). This method still attracts much attention due to its ease of operation and its reasonable price. "Caustic wash" is not capable of removing heavy mercaptans and polycyclic sulfur compounds (Liu et al., 2010).

The two well-known catalytic processes of "Merox" and hydrodesulfurization (HDS) are widely used for removal of large quantities of sulfur compounds from liquid hydrocarbon products. The "Merox" process is an efficient and economical catalytic method, which can be divided into two categories (Meyers, 1986).

- Merox extraction: removes mercaptans from petroleum fractions via chemical treatment.
- Merox sweetening: converts mercaptans to less-harmful disulfides but keeps them inside the petroleum fraction.

The Merox process is not an available choice for sour condensate desulfurization in the Khangiran gas processing plant, because the catalyst is not readily available due to the existing sanctions.

More efficient HDS processes, which operate under high partial pressure of hydrogen and use cobalt molybdenum catalysts, also provide deep desulfurization of various petroleum fractions by converting nearly all sulfur compounds into H₂S (Campos-Martin et al., 2010). This process is adequate for reactive compounds having accessible sulfur atoms, such as thiols, disulfides, and some sulfides. On the other hand, larger compounds, particularly those with sterically hindered sulfur atoms especially thiophene derivatives (e.g. benzothiophenes and dibenzothiophenes) cannot be easily hydrogenated via the HDS process and experience significant hydrogen consumption and severe operating conditions to overcome this issue (Shafi and Hutchings, 2000). Employment of the HDS process to remove sulfur compounds is not economical for Khangiran sour gas condensate due to its small production volume compared to petroleum refineries and restricted access to hydrogen from nearby sources (Mohebbali and Ball, 2008).

Finally, the oxidative desulfurization (ODS) process is a relatively new technology for deep desulfurization of petroleum fractions (Campos-Martin et al., 2010). The most important advantages of the ODS process are their mild operating (reaction) condition and economical facility requirements. Interestingly, this process is capable of oxidizing compounds with sterically hindered sulfur atoms which makes it more robust and desirable than the HDS process in some cases (Jiang et al., 2011). The ODS process is based on the ability of bivalent sulfur to incorporate an oxygen atom without needing to break carbon-sulfur bonds by expanding its valence shell and converting them to sulfoxides or sulfones (Nehlsen, 2005).

According to Nehlsen: "most research in the field of desulfurization is focused into three categories: oxidative desulfurization, metabolism of sulfur compounds using microbes, and selective adsorbents. Of these three, only oxidative desulfurization is being widely developed for commercial use" (Nehlsen, 2005).

Various desulfurization scenarios are investigated in this article to entirely remove H₂S and mercaptans while reducing the other sulfur components of Khangiran sour gas condensate as low as possible. Coupled ODS and caustic wash processes are also reviewed in this research. Both processes benefit from simple procedures and mild operating conditions. The effects of various operating variables are examined to achieve optimal conditions for efficient deep desulfurization of sour condensate. The techno-economical aspects of an industrial scale unit are also presented.

2. Experimental scenarios and procedures

A variety of scenarios have been considered to achieve efficient sulfur removal capacities for ODS and caustic wash techniques to remove various sulfur compounds from sour gas condensate. The following scenarios are exploited in this research to investigate the effect of different reactants on the overall performance of the desulfurization process.

- Scenario 1: Caustic wash desulfurization
- Scenario 2: Oxidative desulfurization via H₂SO₄
- Scenario 3: Oxidative desulfurization via H₂SO₄ coupled with caustic wash
- Scenario 4: Caustic wash coupled with oxidative desulfurization
- Scenario 5: Consecutive repetition of scenarios 3 and 4.

Laboratory experiments were carried out to simulate each scenario under a variety of operating conditions such as various acid or caustic concentrations, different volume ratios of acid or caustic to sour condensate, mixing temperatures and mixing times.

For each experiment, 100 ml sour condensate was initially used at ambient pressure. After caustic or acid wash steps, samples were washed using distilled water to minimize the residual salts produced and neutralize excess acid or base remaining in the gas condensates. To enhance mass transfer rate, a magnetic stirrer was used to mix the organic and aqueous phases. A small decanter was employed to separate the light and heavy phases from each other. The sour and treated condensates were sampled and analyzed for H₂S, mercaptans and total sulfur content by ASTM methods of D-1159, D-3227 and D-1266, respectively. All analysis tests were performed by the Khangiran gas processing plant laboratory.

3. Results and discussion

The earlier detailed analysis of S500 sour condensate indicated the existence of numerous sulfur containing derivatives (e.g.: elemental sulfur, H₂S, mercaptans, sulfides, disulfides, cyclic sulfides, thiophenes, benzothiophenes, dibenzothiophenes, methyl dibenzothiophenes, dimethyl dibenzothiophenes). Removal of H₂S and light mercaptans were the main objectives of this project because of their offensive odor and toxicity. Other sulfur compounds are addressed in total sulfur content. Several analyses of fresh condensate and other samples are repeated to compute the error levels (43 ppm and 351 ppm for mercaptans and total sulfur content, respectively) of experimental measured analysis data. The following sections provide the collected experimental results and brief discussions for the five scenarios mentioned earlier. A more detailed discussion on collected results might be more speculative than the brief experimental review presented, due to the diversity of sulfur compounds and the complexity of chemical reactions involved in all scenarios.

3.1. Caustic desulfurization

Experiments using the first scenario were carried out with caustic to sour condensate volume ratios of 0.2, 0.5, 1, 1.33 and 2, at NaOH concentrations of 1, 2 and 3 M, at temperatures of 10, 35 and 60 °C and with mixing times of 10, 20 and 30 min Table 2 shows the amount of residual sulfur compounds measured in treated condensates for the above conditions. As it can be seen, H₂S is completely removed from all samples, while different degrees of reduction were achieved for mercaptans and total sulfur content.

In order to find the optimal operating conditions, the collected results for residuals of mercaptans and total sulfur content are plotted for each parameter in Figs. 1 and 2. The sulfur removal efficiency initially increases with caustic to condensate volume ratio

Table 2

Measured analysis data for various sulfur compounds remaining in treated condensates after caustic wash process.

Caustic wash							
Run no.	Operating parameter				Sulfur content (ppm)		
	Time (min)	Concentration (M)	Temperature (°C)	Vol. ratio	H ₂ S	RSH	Total sulfur
(1)	20	1	35	0.2	NIL	392.4	7660
(2)	20	1	35	0.5	NIL	383	7820
(3)	20	1	35	1	NIL	341	7620
(4)	20	1	35	1.333	NIL	301.2	7430
(5)	20	1	35	2	NIL	363	8390
(6)	20	1	10	1	NIL	380	8330
(7)	20	1	60	1	NIL	334	7520
(8)	20	2	35	1	NIL	352	7840
(9)	20	3	35	1	NIL	334	7570
(10)	10	1	35	1	NIL	420	8070
(11)	30	1	35	1	NIL	333	7550

and reaches a maximum around 1.3 for mercaptans and 1 for total sulfur. The temperature dependency of sulfur removal efficiencies has similar trends for both mercaptans and total sulfurs. The nearly ambient temperature (35 °C) seems more appropriate from economical and technical viewpoints. Furthermore, since the flash point of S500 is about 14 °C, 35 °C is quite acceptable in terms of safety. Figs. 1c and 2c clearly show that the mercaptan and total sulfur removal capabilities of caustic wash process do not change significantly with increasing caustic concentration from 1 M to 3 M. The variation observed for both cases are well inside the accepted error levels. From practical considerations, 1 M caustic solution is more appropriate than higher concentrations.

A closer examination of Figs. 1d and 2d reveals that mercaptan and total sulfur removal rates are relatively slow at low contact times and both of them gain momentum from 10 to 20 min then they slow down again at high residence times above 20 min. The later value seems adequate for the caustic desulfurization process.

The experimental results indicate that the volume ratio of caustic to sour condensate and mixing time are the key variables in the caustic wash process. On the other hand, caustic concentration and mixing temperature have negligible effects on the overall desulfurization performance.

Even the best caustic desulfurization process cannot provide an acceptable decrease in the amount of sulfur containing compounds. It was only able to produce 34 and 7 percent reductions in mercaptans and total sulfur content, respectively. Evidently, caustic wash alone would not be a suitable choice and efforts should be continued to find better scenarios.

3.2. Oxidative desulfurization via H₂SO₄

Sulfuric acid can be considered as an oxidizing agent in the presence of sulfur atoms with a lower oxidation state (such as sulfides and sulfates) (Nehlsen, 2005). This acid oxidizes sulfur compounds and extracts them into the aqueous phase. The experiments for the second scenario were carried out with volume ratios of acid to sour natural gas condensate of 0.5, 1.1, 1.33 and 2, at acid concentrations of 49 and 98 wt %, at temperatures of 10, 35 and 60 °C and mixing times of 5, 15 and 25 min Table 3 provides the amount of residual sulfur compounds in treated condensate after desulfurization via H₂SO₄ for various operating conditions. Interestingly, H₂S and mercaptans are completely removed from all samples except the one with the lowest acid concentration (49 wt %). As shown in Fig. 3a, the volume ratio of acid to condensate of around 2 is enough to oxidize nearly all sulfur compounds. As before, Fig. 3b illustrates that mixing temperature has negligible effect on the overall separation of sulfur compounds and near ambient temperature (35 °C) can be more promising in terms of safety and ease of operation. The results shown in Fig. 3c indicate that the total sulfur removal rates are extremely slow at low acid concentrations and accelerate as acid concentration approaches 98 wt%, which is the actual industrial grade readily available in market. Use of higher acid concentrations is not feasible from economical standpoints.

As shown in Fig. 3d, the sulfur removal increases for the initial 15 min and then remains nearly constant at higher contact times. This issue may be attributed to reaching equilibrium state for some sulfur compounds. Contrary to the previous caustic wash process, acid concentration has a significant effect on the process efficiency while, as before, other parameters have minor effects on the overall desulfurization rate of sour condensate. Desulfurization via H₂SO₄ at optimal conditions not only entirely removes H₂S and mercaptans but also can successfully remove more than 91 percent of total sulfur content from sour condensate which makes this process more superior to previous one.

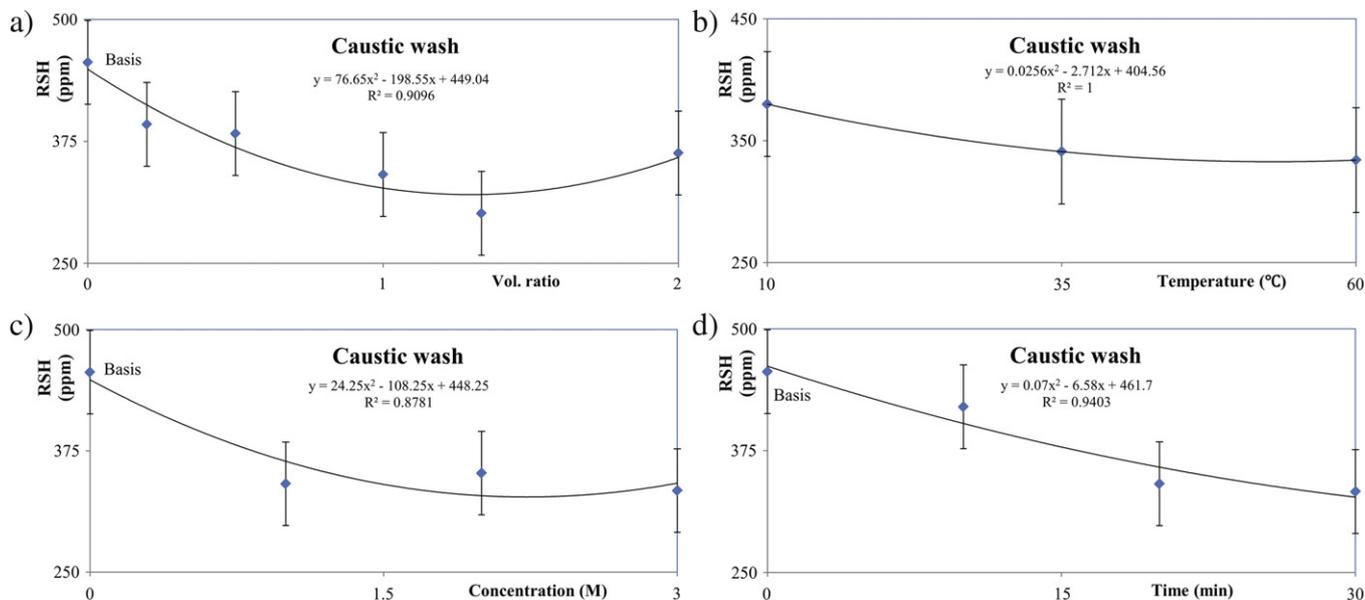


Fig. 1. Variations of mercaptan contents of caustic treated condensates at different a) volume ratios (caustic/condensate) b) Mixing temperatures c) caustic concentrations and d) mixing times.

3.3. Consecutive oxidative desulfurization and caustic wash

Combinations of the two previous processes were used to achieve more sulfur reduction. In the first scheme, a two-step process consisting of ODS via H₂SO₄ (first step) followed by a caustic wash was used. In the next scheme, the processes were used in reverse order. The results shown in part A of Table 4 indicate that the caustic wash process is not able to create more reduction in total sulfur content after H₂SO₄ treatment due to the elimination of all light sulfur compounds in the oxidative step. On the other hand, initial caustic wash followed by oxidative desulfurization can yield greater sulfur reduction. This can be due to elimination of light sulfur compounds in the first step which may provide more

opportunity for the oxidative process to remove heavier sulfur compounds from sour condensate.

Three and four step processes which consisted of consecutive caustic wash and oxidative desulfurization were carried out in order to investigate the possibility of higher sulfur removal efficiencies. The levels of residual sulfur content are presented in part B of Table 4. As shown in Fig. 4, it is superfluous to treat sour condensate via consecutive repetition of oxidative desulfurization and caustic wash processes while the same objective can be achieved by a single ODS process.

As mentioned above, the results presented so far indicate that an economical desulfurization process should consist of a single stage oxidative desulfurization process via H₂SO₄ in a well mixed

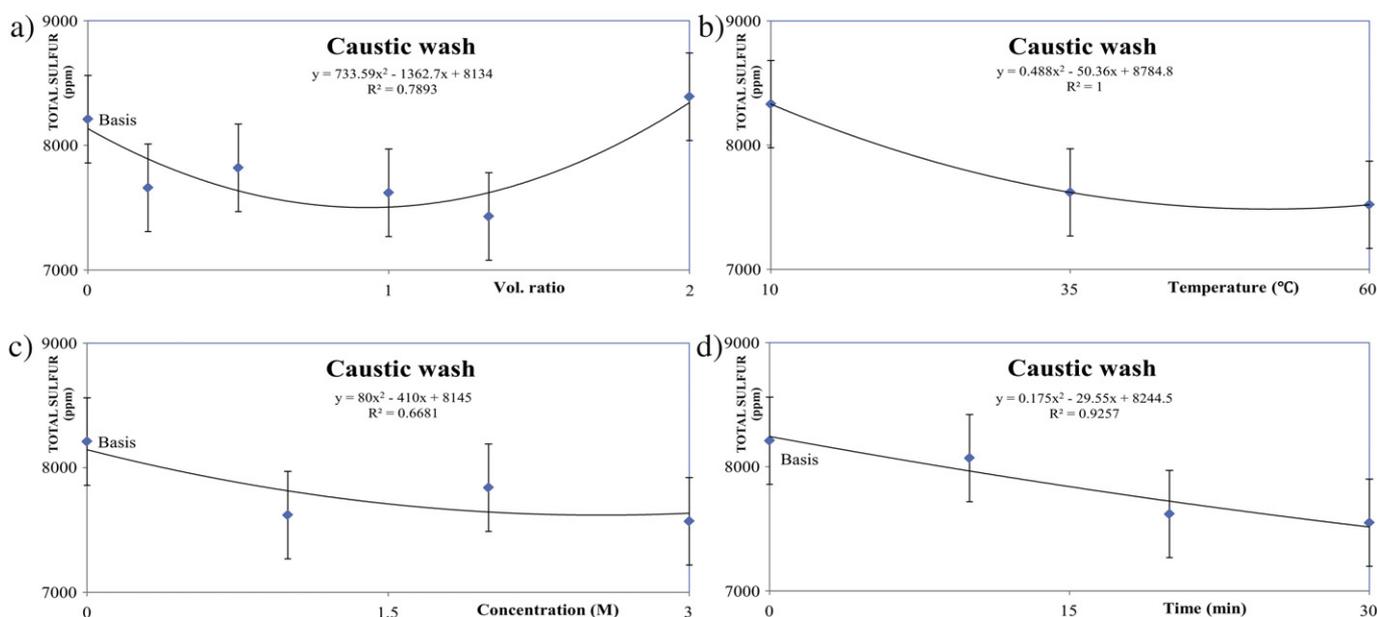


Fig. 2. Variations of total sulfur contents of caustic treated condensates at different a) volume ratios (caustic/condensate) b) Mixing temperatures c) caustic concentrations and d) mixing times.

Table 3

Measured analysis data for various sulfur compounds remained in treated condensates after desulfurization via H₂SO₄.

Oxidative agent							
Run no.	Operating parameter				Sulfur content (ppm)		
	Time (min)	Concentration (wt %)	Temperature (°C)	Vol. ratio	H ₂ S	RSH	Total sulfur
(1)	15	98	35	0.5	NIL	NIL	1200
(2)	15	98	35	1	NIL	NIL	1060
(3)	15	98	35	1.333	NIL	NIL	800
(4)	15	98	35	2	NIL	NIL	700
(5)	15	98	10	1	NIL	NIL	1100
(6)	15	98	60	1	NIL	NIL	800
(7)	15	49	35	1	NIL	352	8080
(8)	5	98	35	1	NIL	NIL	2340
(9)	25	98	35	1	NIL	NIL	1010

agitator at the optimal operating conditions, which are: ambient temperature, acid concentration of 98 wt %, volume ratio of acid to sour condensate around 2 and about 15 min contact time.

After removing the offensive odor, the next issue was to improve the color of sour condensate as much as possible. Fig. 5 shows the colors of various solutions encountered in different stages during the sample preparation processes. As shown in Fig. 5a, the initial sour condensate was yellow. Caustic treated condensates still had a lighter yellowish color as depicted in Fig. 5b. On the other hand, the oxidative desulfurization process totally removes the yellow color

and produces a clear and transparent phase as shown in Fig. 5c. Evidently, the improved color is more attractive from marketing viewpoint.

Fig. 5d illustrates that during mixing of sour condensate with caustic solution no appreciable color change occurs while in the process of mixing the sour condensate with H₂SO₄ the color of the aqueous phase turns to heavy red as shown in Fig. 5e. This sharp change in color is due to the formation of colloidal elemental sulfur in the aqueous phase, which is extracted from organic phase (Zhang et al., 2000).

And last but not least, it should be emphasized that the use of extremely high intensity stirring of condensate and the aqueous phase may cause formation of a stable emulsion as shown in Fig. 5f. Evidently, such stable emulsions cannot be easily broken without the use of certain chemical de-emulsifiers or application of some external electrostatic fields (Anonymous, 2004; Thiam et al., 2009). Both of these treatment methods are not desirable from economical views and may produce some severe environmental risks.

4. Techno-economic evaluation of sweetening process

The preliminary Techno-economic evaluation of a single-stage desulfurization process via H₂SO₄ is presented in the following section. First, a separation process is introduced based on the experimental results. Then the sizing calculations of various process equipments are performed and their preliminary cost

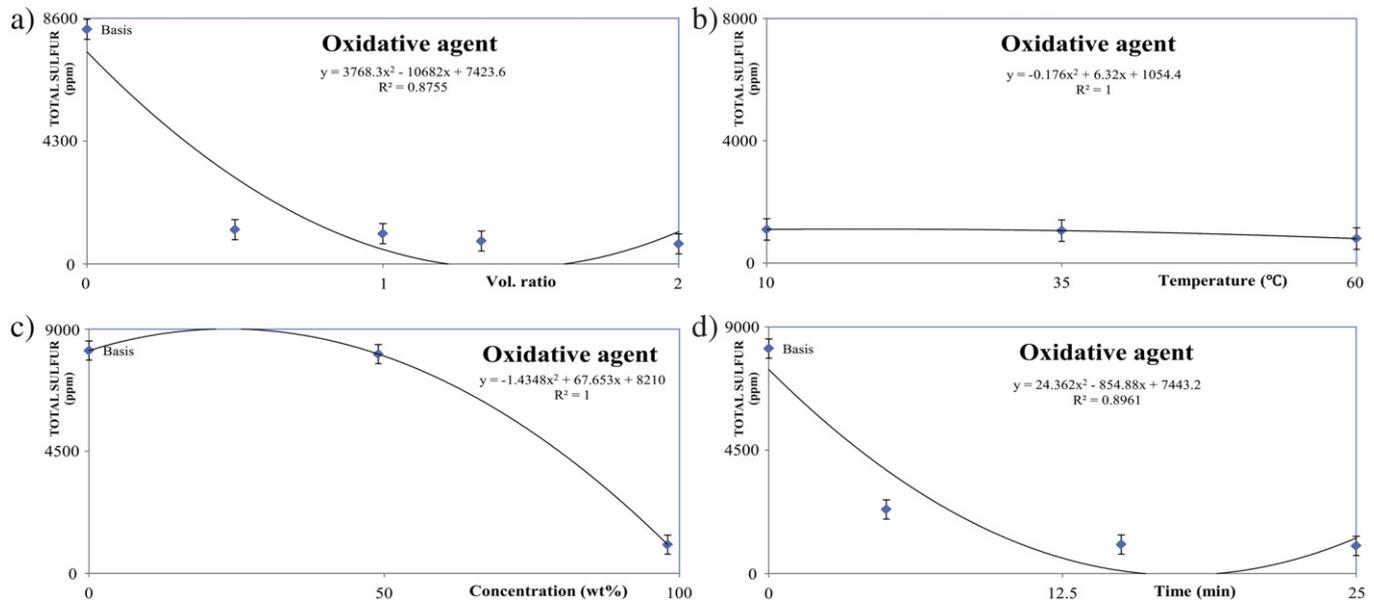


Fig. 3. Variations of total sulfur contents of H₂SO₄ treated condensates at different a) volume ratios (caustic/condensate) b) Mixing temperatures c) caustic concentrations and d) mixing times.

Table 4

Measured analysis data for various sulfur compounds remained in treated condensates after coupled desulfurization processes.

	Run no.	Caustic wash					Oxidative desulfurization					Sulfur content (ppm)			
		Order of step	Vol. Ratio	Temp. (°C)	Concentration (M)	Time (min)	Order of step	Vol. Ratio	Temp. (°C)	Concentration (wt %)	Time (min)	H ₂ S	RSH	Total sulfur	
A	Caustic wash coupled with oxidative desulfurization	(1)	2	1	35	1	20	1	1	35	98	15	NIL	NIL	1060
		(2)	1	1	35	1	20	2	1	35	98	15	NIL	NIL	900
B	Consecutive repetition of oxidative desulfurization and caustic wash	(3)	1 & 3	1 & 1	35	1	20	2	1	35	98	15	NIL	NIL	900
		(4)	2	1	35	1	20	1 & 3	1 & 1	35	98	15	NIL	NIL	1100
		(5)	1 & 3	1 & 1	35	1	20	2 & 4	1 & 1	35	98	15	NIL	NIL	1000

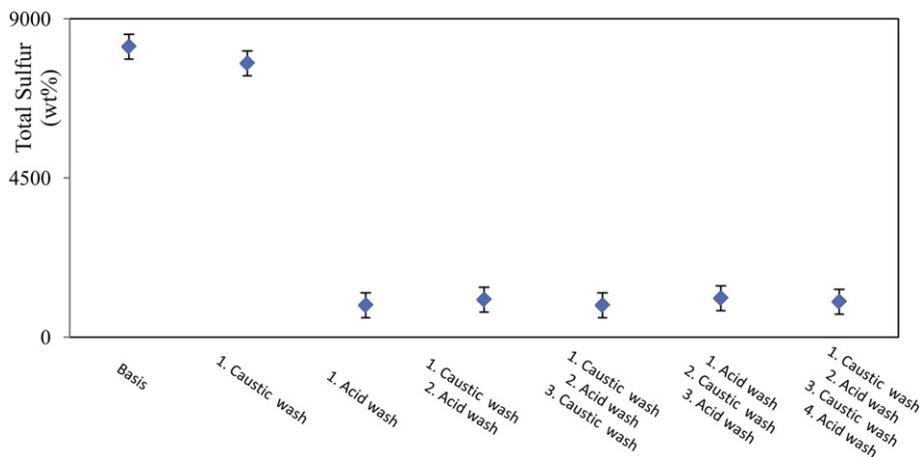


Fig. 4. Comparison of various consecutive process with ODS, caustic wash and base case (sour condensates).

estimations are determined. Finally, a rough profit analysis is presented.

4.1. Process description

With an eye on the results obtained in the previous section, a single-stage oxidative desulfurization process shown in Fig. 6 is recommended for desulfurization of Khangiran sour natural gas condensates. According to Khangiran refinery internal reports, the maximum production rate of sour condensate was around 3000 barrel per day (477 CMD) in winter. This amount is used as the feed rate basis for design of the entire process. The sour condensate

should be desulfurized via the oxidative process using sulfuric acid (H_2SO_4) as the oxidative agent. The acid to sour condensate volume ratio of around 2 is used in the process. Based on the previous experimental results, the operating temperature, mixing time and acid concentration are selected as 35 °C, 15 min and 98 wt %, respectively.

As it can be seen, sour gas condensate (stream 2) and sulfuric acid (stream 4) enter the first agitated stirred tank for mixing. The mixed phases are pumped via stream (6) into a suitable first decanter for separation of the aqueous and organic phases. The heavy acidic phase is collected from the decanter bottom as stream (9) and returns to the acid storage tank for recycling via

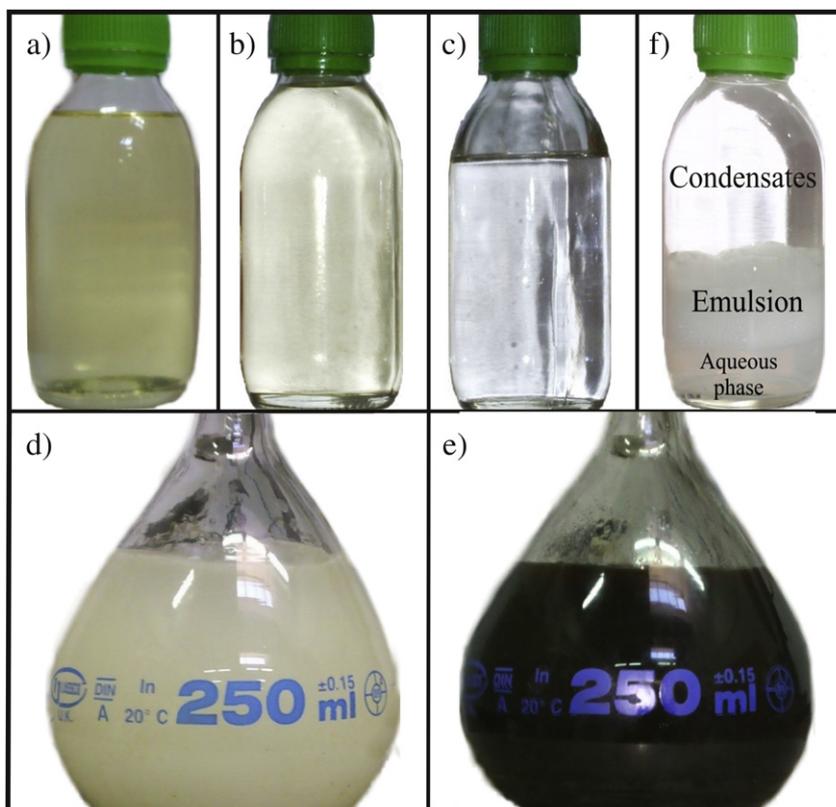


Fig. 5. Colors of various solutions at different stages of experiments: a) Sour condensate, b) Caustic treated condensate, c) Oxidized condensate, d) Mixture of sour condensate with caustic solution, e) Mixture of sour condensate with H_2SO_4 , f) Stable emulsion of two aqueous and organic phases.

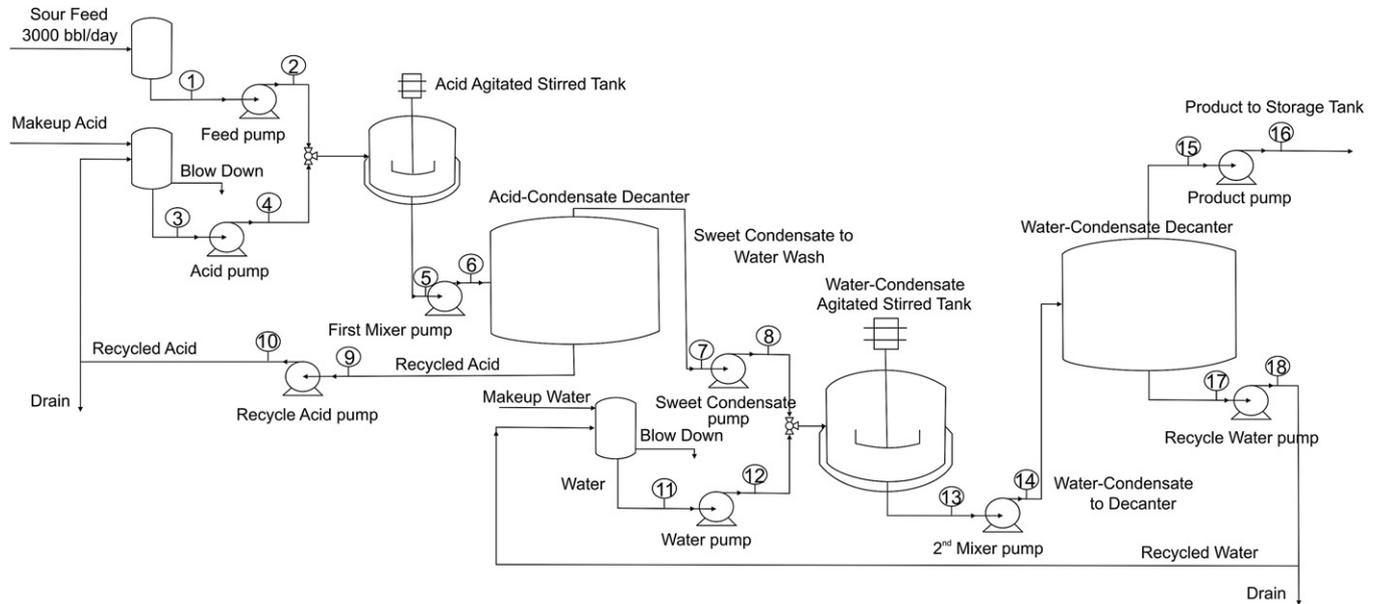


Fig. 6. Process flow diagram for sour condensate desulfurization unit.

stream (10). An appropriate bleed ratio of around 3% per day (and the same amount of fresh make-up acid) is considered to ensure the recycled acid retains its proper strength. The 3% bleed ratio is selected from the experimental results obtained via titration of the spent acid with standard caustic solution. Evidently, improper disposal of such large amounts of acid (around 180 bbl/day) may have severe environmental adverse effects (if not catastrophe). Anionic membranes, lime neutralization and microwave energy cracking can be used to tackle this issue (Faust and Orford, 1958; Masse and June 9, 1987; Simpson and Buckley, 1988).

The sweet condensate (stream 7) is pumped to a second agitated stirred tank for water wash. The volume ratio of water to condensate is chosen as 5 to 1 to ensure complete removal of residual acid and salts formed during the previous stage. The water-wash is used to minimize the residual salts produced and remove all the excess acid contaminating the hydrocarbon phase. Afterward, stream 14 enters the second decanter for proper separation of sweet condensate and the aqueous phase. Final product (stream 15) with less than 700 ppm sulfur content will be pumped to a sweet storage tank and a portion of recycled water¹ (stream 17) can be pumped back into the water storage tank. Surge drums are used prior to each pump to damp the process flow oscillations and provide an additional facility for separation of accumulated contaminants by periodically blowing down the vessel. As mentioned earlier, make-up streams are also included to compensate for acid and water consumption due to blow downs and condensate contaminations. Various pumps are used to ensure proper circulations of all streams.

4.2. Process design

All equipment available in the recommended process flow diagram should be designed appropriately. For this preliminary techno-economic evaluation the design is not developed to any significant extent. The following sections tackle this issue.

4.2.1. Agitated stirred tanks and decanters

Table 5 shows the computed dimensions of vertical and cylindrical agitated stirred tanks and corresponding decanters for both single and two stage (tanks in parallel) processes. The design strategy of reference (Coulson and Richardson, 1955) is used. For the sake of operational flexibility, the two stage scheme is selected. The agitated tanks should be equipped with one-tenth baffles to prevent vortex formation. Flat blade turbine blades with impeller diameter equal to one third of tank diameter is used. Each tank contains single stirrer due to the relatively high viscosity of the mixture. All stirrers are assumed to be driven by a geared electric motor and revolve at 80 rpms. According to the appendix "A" of Iranian petroleum standards (IPS-E-PI-221) carbon steel should be used for sulfuric acid environments of 65 wt% or more in the temperature range of 0–60 °C (Committee, January 2000). In dilute form (below 50 wt %) sulfuric acid is corrosive to carbon steel. Therefore, other alloys (such as Hasteloy or Teflon lining) should be used. According to Handbook of Corrosion Data, written by Bruce D. Craig, 1995, pp. 848 (Craig et al., 1995), "aluminum is used to process a thin (dilute) aqueous slurry containing around 1% sulfuric acid at room temperature". In the ODS process, the sulfuric acid is either very strong (around 98 wt% in the acid agitation tank and its decanter) or it is extremely diluted (in water wash agitation tank and its decanter), therefore carbon steel and aluminum tanks are proved to be the most economical choices for acid tank and water wash vessels, respectively.

4.2.2. Process piping and pumps

Table 6 presents the compositions, volumetric flow rates and computed sizes of pipes connecting various facilities. The stream numbers are previously shown in Fig. 6. The following equation is

Table 5
Dimensions of designed agitated stirred tanks and corresponding decanters.

Name	One vessel			Two parallel vessels			
	D(m)	H(m)	V(m ³)	D(m)	H(m)	V(m ³)	V _r (m ³)
Acid-condensate mixer	2.44	2.44	11.40	2.00	2.00	6.26	12.52
Acid-condensate decanter	4.00	8.00	100.48	2.90	6.00	39.48	78.97
Water-condensate mixer	2.44	2.44	11.40	2.00	2.00	6.26	12.52
Water-condensate decanter	4.00	8.00	100.48	2.80	6.00	36.81	73.62

¹ Specification of this issue requires more experiments.

Table 6
Process designed pipe diameters.

Parameter	Notation	Unit	Stream numbers								
			1	3	5	7	9	11	13	15	17
Composition		–	Sour cond	H ₂ SO ₄ (98%)	H ₂ SO ₄ + Sweet cond	Sweet cond	H ₂ SO ₄	Water	Water + Sweet cond	Sweet cond	Water
Flow rate	Q	m ³ /d	447.0	954.0	1431.0	477.0	954.0	2385.0	2862.0	477.0	2385.0
Pipe diameter	D	in	4.0	6.0	6.0	4.0	6.0	8.0	8.0	4.0	8.0

Table 7
Specifications of process pumps.

Pump Name	Q(m ³ /h)	H(m)	ΔP _{pump} (kPa)	W _{pump} (kW)
Sour condensate pump	19.9	5	42.1	0.27
Acid pump	39.8	5	93.8	1.22
First mixer pump	59.6	10	152.7	2.98
Recycle acid pump	39.8	10	184.3	2.39
Sweet condensate pump	19.9	5	42.1	0.27
Water pump	99.4	5	51.4	1.67
Second mixer pump	119.3	10	97.9	3.81
Recycle water pump	99.4	5	51.4	1.67
Product pump	19.9	15	123.3	0.80

used to calculate the optimum stream velocity (McCabe et al., 2005),

$$V_{opt}(ft/s) = \frac{12\dot{m}^{0.1}}{\rho^{0.36}} \quad (1)$$

where \dot{m} is the mass flow rate (lb/s) and ρ is fluid density (lb/ft³). The optimum pipe sizes (internal diameters) are readily computable by dividing actual flow rate by above optimal velocity ($D = \sqrt{4Q/\pi V_{opt}}$). As it can be seen in Table 6, all pipe sizes are in the range of 4–8 inches in diameter which is quite acceptable.

Table 7 provides the volumetric flow rates, required heads, total pressure drops encountered and corresponding powers of various centrifugal pumps with efficiency of $\eta = 0.85$. Evidently, the required head of each pump is selected from the level of liquid in the downstream tank and pressure drops are due to skin friction

Table 8
Process equipment (fixed) costs.

Type	Equipments	No.	Unit price (2012)	Total cost (2012)		
Pump	Feed pump	1	15,000	15,000		
	Acid pump	1	15,000	15,000		
	First Mixer pump	1	15,000	15,000		
	Recycle Acid pump	1	15,000	15,000		
	Sweet Condensate pump	1	15,000	15,000		
	Water pump	1	16,000	16,000		
	2nd Mixer pump	1	16,000	16,000		
	Recycle Water pump	1	16,000	16,000		
	Product pump	1	15,000	15,000		
	Agitator	Agitator	4	23,000	92,000	
		Electromotor + Coupling	Feed pump	1	1000	1000
			Acid pump	1	1000	1000
			First Mixer pump	1	1000	1000
Recycle Acid pump	1		1000	1000		
Vessel	Sweet Condensate pump	1	1000	1000		
	Water pump	1	1000	1000		
	2nd Mixer pump	1	1000	1000		
	Recycle Water pump	1	1000	1000		
	Product pump	1	1000	1000		
	Agitators	4	2000	8000		
	Mixer	4	32,000	128,000		
	Acid-Condensate Decanter	2	60,000	120,000		
	Water-Condensate Decanter	2	59,000	118,000		
	Sum			613,000		

because each process step operates at atmospheric pressure. The following equation is used in SI units to compute the required powers:

$$\text{Power}(W) = [\Delta P_H + \Delta P_f] \frac{Q}{\eta} = \frac{\Delta P_{\text{Pump}} Q}{\eta} \quad (2)$$

The computed values for pump powers varies between 0.27 kW (0.36 hp) and 3.81 kW (5.1 hp) which require single phase electromotor drivers.

4.3. Economical aspects

Tables 8 and 9 summarize various economical aspects of the entire project based on the designed parameters found in the previous section (by using the experimental data collected in this work).

All prices are computed from reference (Walas, 1990) and corrected for the second quarter of 2012 using appropriate cost indexes (Chemical Engineering and Aug, August 2012). As it can be seen, the entire project requires around 2.5 million dollars investment based on second quarter of 2012 prices, however this value is very preliminary and subject to significant development and requires contingency and allocation of an accuracy range.

Table 9
Other capital investments.

Items	% of total	Cost \$ (2012)
Purchased equipment	25	611,000
Purchased-equipment installation	6	147,000
Instrumentation (installed)	10	244,000
Piping (installed)	8	196,000
Electrical (installed)	5	122,000
Buildings (including services)	5	122,000
Yard improvements	2	49,000
Service facilities (installed)	13	318,000
Land	1	24,000
Engineering and supervision	7	171,000
Construction expense	10	244,000
Contractor's fee	2	49,000
Contingency	6	147,000
Total cost	100	2,444,000

Table 10
Annual process variable (current) costs.

Items	Quantity	Unit price	Total cost (\$ per annum)
Process water (treated)	71.55 (m ³ /day)	5 (\$/m ³)	118,000
Process sulfuric Acid (98 wt %)	28.62 (m ³ /day)	178.6 (\$/m ³)	1,686,000
Electricity	1200 (kW-hr/day)	0.12 (\$/kW-hr)	43,000
Manpower	4	2000 (\$/month)	96,000
Depreciation	10% (Based on ten years)	244,000 (\$)	244,000
Maintenance	2% of capital cost	–	49,000
Total	–	–	2,236,000

Table 10 illustrates all current cost required to produce 3000 bbl/day sweet condensate for 330 active days. Total water and acid circulation rate for the entire process are 2385 and 954 cubic meters per day, respectively. Assuming about 3% water and acid make up in each day, the consumption rates of fresh water and acid will be about 72 and 29 cubic meters per day. According to Table 10, around 2.26\$/bbl is required to process the sour condensate. A small profit can be anticipated because the sweet condensate is much more saleable than the sour one.

It should be emphasized that the main objective of this project is its environmental and safety (HSE) impact rather than economical benefits. Furthermore, sweetening of sour condensate can significantly alleviate the corrosion problems from gas refinery to consumption point.

5. Conclusion

Production of sour condensate in Khangiran gas field has been tripled in the last six years due to several technical reasons. Transportation, storage and consumption of these condensates have led to severe health, safety and environmental issues. Efficient desulfurization of sour condensates can dramatically improve HSE standards. Various experimental scenarios were considered in this research using different reagents, temperatures, volume ratios and concentrations. The oxidation process which used sulfuric acid as the oxidative agent provided the most effective and economical procedure for deep desulfurization which was able to entirely remove the hydrogen sulfide and mercaptans while reducing the total sulfur content from over 8000 ppms to less than 700 ppms. The proposed desulfurization method required around 2.26\$/bbl to process such a sour condensate. Also, small profit can be anticipated because the sweet condensate is much more saleable than the sour one. It should be emphasized that the HSE aspects of this process is much more desirable than its financial aspects.

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