A Multi-Layered Strategy to Manage Foaming in Ultra-Sour Gas Sweetening Units

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Abstract

Foaming is a major operational occurrence seen in gas sweetening units, causing production losses, off-spec product, solvent losses, equipment corrosion and excessive consumption of anti- foam chemicals. Foaming can therefore shutdown an entire unit if mismanaged.

This paper details a multi-layered bulletproof strategy to manage foaming while operating and designing ultra-sour gas sweetening units. Process simulation models of an operational amine unit at ADNOC Sour Gas (ASG) were built to showcase the positive impacts of each of these recommendations on the performance of the unit. This multi-pronged guideline that adds a safety-net around the upstream separation equipment, sour feed gas, lean amine, and absorber include: (1) Internal modifications to the upstream feed gas separator; (2) Fail-safe design practice around feed separator; (3) Recalculation & modification to the conventional 5 degC temperature approach rule for the lean solvent & (4) Robust differential pressure measurement across absorber.

The above recommendations demonstrated a CAPEX and OPEX saving of US$ 25 MM and US$ 0.6 MMPA respectively to ASG. In addition, reduction in process upsets and pressure drops in the gas circuit generated production related revenues of > US$ 2 million/yr.

<u>Keywords:</u> Middle East, Ultra Sour Gas Sweetening, Foaming, Process Simulation, Rules of Thumb, Operational Strategy

Introduction
Gas Sweetening

Undesirable contaminants are found in nearly all gas and liquid hydrocarbon streams. These must be removed before the end products can be effectively marketed as fuels. The undesirable components specifically considered in this paper are Hydrogen Sulfide (H2S) and Carbon Dioxide (CO2). These are collectively referred to as “sour gases” or “acid gases”. They are generally undesirable because of their toxicity, corrosivity, harmful environmental impacts and negative effect on fuel heating value.

Processes based on amine-based solvents are one of the most commonly used methods for removing acidic components from both gas and liquid streams. These are often referred to as “sweetening” processes since the end product is a stream that is purged of components that are “sour” in nature. The popularity of amine systems is largely linked by the system’s ability to regenerate the spent solvent while still achieving sufficient removal of the CO2 and H2S.

The general process flow for an amine sweetening unit is shown in Figure 1. The gas is fed to an absorber column, where it comes into contact with the solvent. The sour components are absorbed by the solvent, and the sweetened gas leaves the top of the column. The spent or “rich” solvent is then piped to a stripper column for regeneration. The regenerated, also called “lean”, solvent is recycled back to the absorber. Amine regeneration typically involves flashing the solvent down to lower pressure than the absorber, and/or increasing the temperature of the solvent. The purpose of the three main components in this process (absorber, flash tank, regenerator) is explained in further detail below.

The absorber (also called a contactor) is the primary equipment for the gas sweetening process. Its purpose is to bring the feed sour gas and amine solvent into contact so that the acid gases can be transferred from the vapor phase into the liquid phase. The transfer of sour gases to the solvent are generally favored at low temperatures and high pressures.

The regenerator strips out the absorbed acid gases from the solvent so that it can be recycled back into the absorber for further sweetening. Acid gases are more volatile compared to amine solvents; therefore, regeneration is typically favored at low pressures and high temperatures. The regenerator is typically operated at around 13 – 17 psig, and a reboiler (either steam or hot oil driven) is placed at the column bottom to provide boil-up and increase in column temperature. A condenser is often attached to the column top to induce reflux and reduce solvent / water losses.

At high feed gas pressures, amine solvents also end up absorbing some amounts of the hydrocarbon components on top of the target sour contaminants. These hydrocarbons are loosely bound to the solvent and can therefore be removed quite easily by dropping the pressure. The flash drum (positioned in between the contactor and regenerator) provides a convenient location to drop the pressure and recover the hydrocarbons by separating the vapor from the leftover solvent. This can considerably reduce the hydrocarbon losses from the regenerator overheads.

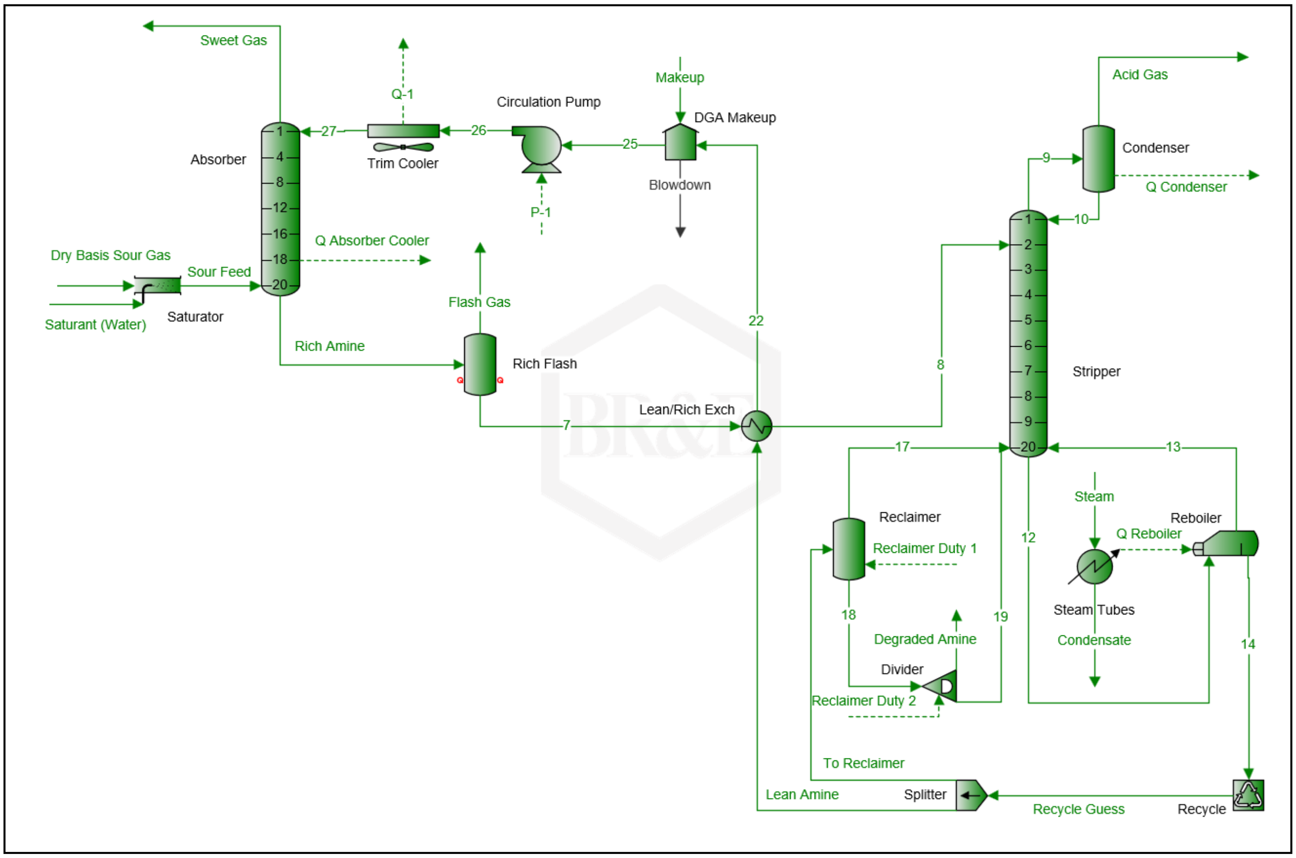


Figure 1: Typical Amine Sweetening Unit Process Flow (DGA)

Historically, majority of the gas sweetening units around the world operated at around 35 deg C, 30 bar and contained feed acid gas concentration of 5%. However, these conditions are becoming less common with the development of new ultra-sour gas fields, especially in the Middle East.

ADNOC Sour Gas Plant – Process Overview

ADNOC Sour Gas are one of Abu Dhabi National Oil Company’s gas producers. Considered to be a pioneer in sour gas field development, it has the capacity to produce over 1 billion cubic feet of sour gas per day.

ADNOC Sour Gas is a joint venture between ADNOC (with a 60% share), and Occidental Petroleum (with a 40% share). It was established over a decade back in 2010 to exploit the Shah Gas Field. Since its commencement of operations, it has built a global reputation as a reliable supplier of clean gas and one of the world’s largest sulfur manufacturers.

The sour gas processed at ADNOC Sour Gas’ Shah plant is anything but typical. It has greater temperatures (~60 deg C), pressures (~70 bar) and acid gas content (25% H2S + 10% CO2), compared to the average gas plant.

The Shah plant’s sweetening unit has two identical trains, designed to process 1 BCFD of ultra-sour gas and later upgraded to 1.28 BCFD. The amine solution used in the unit is an aggressive primary amine, DGA®, that removes almost 99.9% of the acid gas (both H2S and CO2).

The sour gas is preconditioned by first passing through an inlet condenser with the main purpose of condensing heavy hydrocarbons in the feed stream, followed by a separator that is designed to knock out these heavy liquids. Post that, the gas is heated up slightly in a cross exchanger to keep the feed gas temperature much above HC condensation temperature before being fed into the absorber where it is contacted by the lean DGA® solvent. The process flow and conditions of theamine absorption unit is given in the figure below.

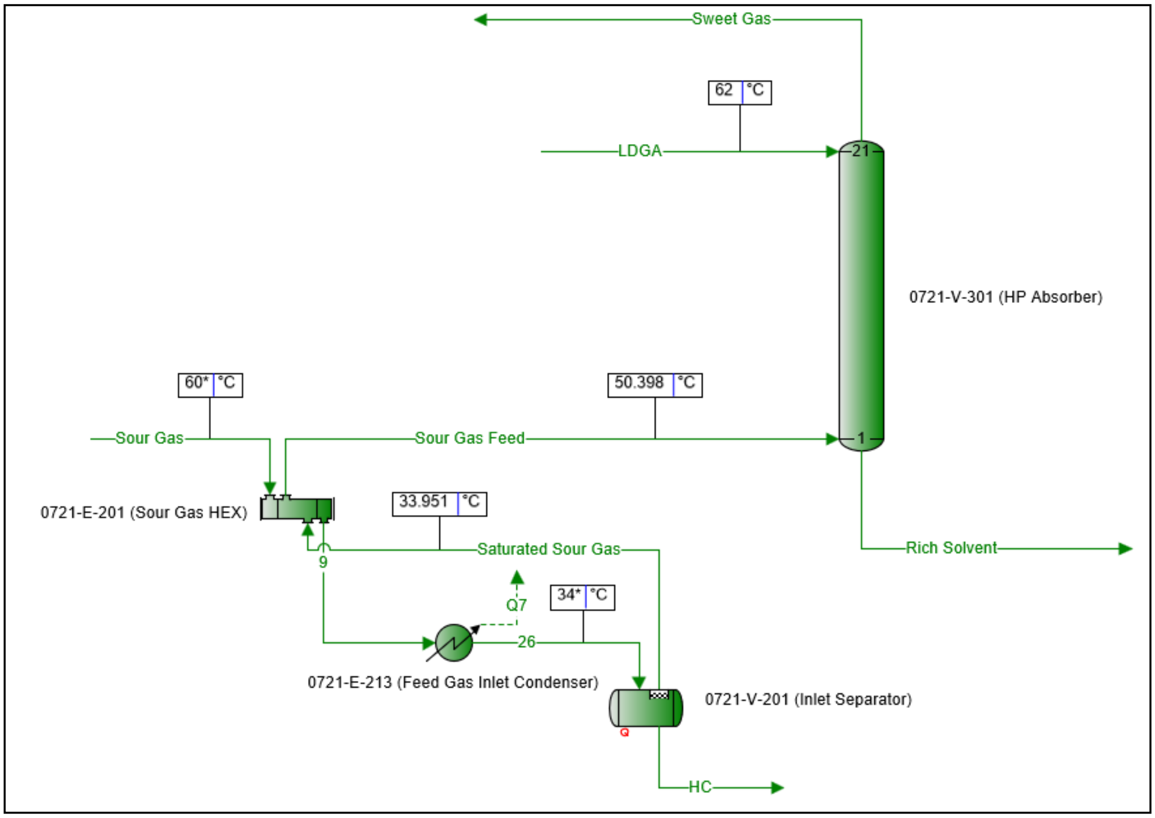


Figure 2: ASG Current HP Absorber Scheme

Challenges

One of the biggest advantages of amine sweetening processes is its ability to regenerate the spent solvent and recycled back to the absorber making it an overall OPEX light process. However, such processes also come with its own sets of cons and challenges.

The presence of hydrocarbon liquids in amine systems can result in a phenomenon known as foaming. Foaming is a phenomenon caused due to the mechanical incorporation of gas into a liquid phase,where the liquid ends up surrounding a volume of gas, subsequently creating bubbles. It is a major operational occurrence seen in gas sweetening units that can negatively impact product quality and can result in production losses, solvent losses, equipment corrosion and excessive consumption of anti-foam chemicals. Foaming can therefore shutdown an entire unit if mismanaged and negate the OPEX savings that a sweetening unit expects to achieve.

During the normal operation and during the work towards expanding the plant capacity to 1.45 BCFD, ASG noticed significant amounts of foaming in its HP absorber and potential for furtherdeterioration after expansion. It was concluded that high amounts of liquid carry over from the inlet separator is causing condensation of hydrocarbon in the absorbers resulting in foaming.

To salvage the plant from foaming and manage the situation, ASG initially deliberated over three (3) potential short-term decisions:

- Additional anti-foam injection into the absorbers at an overhead of US$ 0.6 MMPA
- Run the sweetening unit at lower throughput reducing overall plant production
- Install new sour gas filter coalescers worth US$ 25 MM at the absorber inlet

Long-Term Strategies to Mitigate Challenges

To effect more long-lasting, bullet proof and cost-effective solutions to the problem of foaming,four strategies were devised. This was done by looking at potential improvements & safety margins that can be added to major participants involved in the amine absorption process:

- Inlet separator
- Sour feed gas
- Lean amine
- Absorber column

Internal Modifications to the Inlet Separator

While the purpose of an inlet separator is to knock out heavy hydrocarbons before the sour gasenters the absorber, an analysis done on its expected performance at the target 145% plant capacity showed severe limitations and significant liquid carryover in the sour gas to downstream HP absorber. To resolve this issue, internal alterations were made to the upstreamfeed separator such as vane type inlet device, mesh coalescer and demisting cyclones at the outlet. These recommended modifications demonstrated up to 99.9% reduction of fine mist carry-over (potential foaming triggers) to the gas phase.

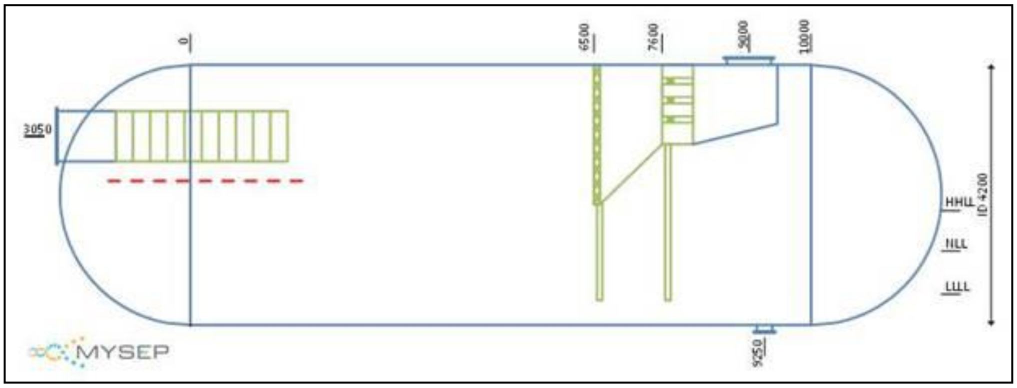


Figure 3: Internal Modifications to Inlet Separator

A simulation verification was also conducted to validate the benefits of reduced liquid carry-overto the sour gas downstream of the separator. The results indicated that a drop in liquid entrainment in the vapor phase lowers the hydrocarbon dew point of the sour gas entering the absorber, adding a process safety-net, and thereby reducing the chances of any potential condensation.

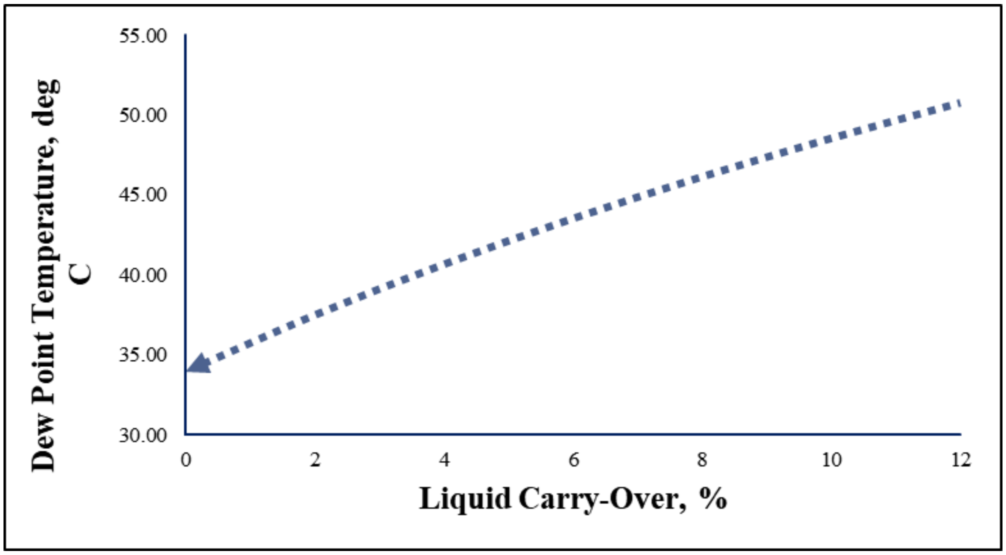


Figure 4: Effect of Liquid Carry-Over on Sour Feed Dew Point Temperature

Fail-safe design practice around feed separator

To aid in the knockout of heavy liquids in the feed separator and vaporize any potential carryover, it is recommended to sub-cool the sour fluid before the inlet separator and superheatthe gas coming out of it before it enters the absorber.

The advantages of “subcooling + superheating” the sour gas can be demonstrated when it iscompared to a scenario when this technique isn’t followed.

A process simulation model for both cases show that with subcooling + superheating of the sourgas, the final feed stream to the absorber is around 50 deg C but additionally the dew point temperature is only 30 deg C, almost 20 deg C lower than feed temperature. This gives a factor of safety in a scenario of any potential hydrocarbon condensation.

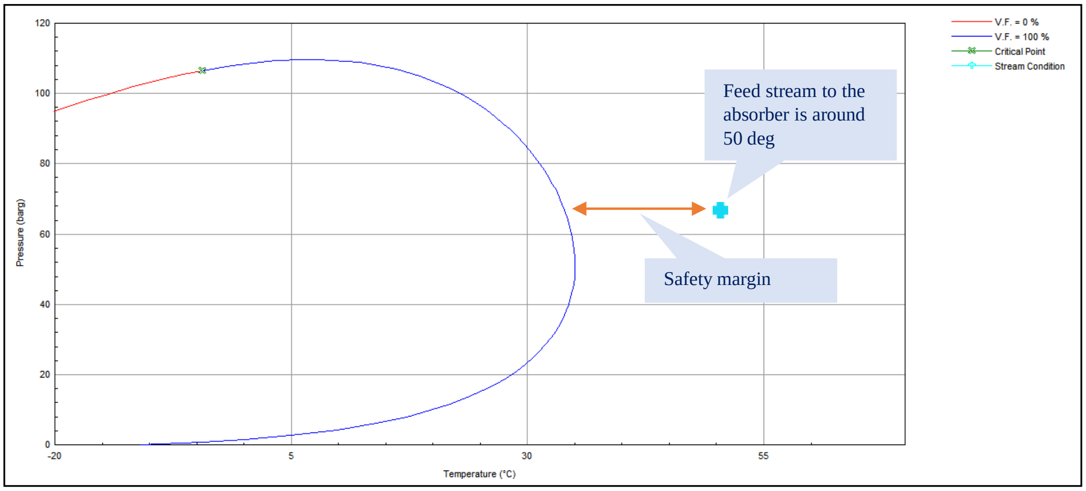


Figure 5: Dew Point Safety Margin Achieved in Feed Gas (Scenario: Sub-Cooling + Superheating)

With no subcooling + superheating of the sour gas, the final feed stream to the absorber is saturated and therefore its temperature is equal to its dew point (~60 deg C). In such a scenario, there is a high possibility of hydrocarbon condensation since the sour gas' dew point is relativelyclose to the amine inlet temperature (62 deg C). This can potentially increase the chances of foaming in the absorber too. Furthermore, the amount of hydrocarbon liquid knocked out in the inlet separator is also almost 60% lower in this scenario.

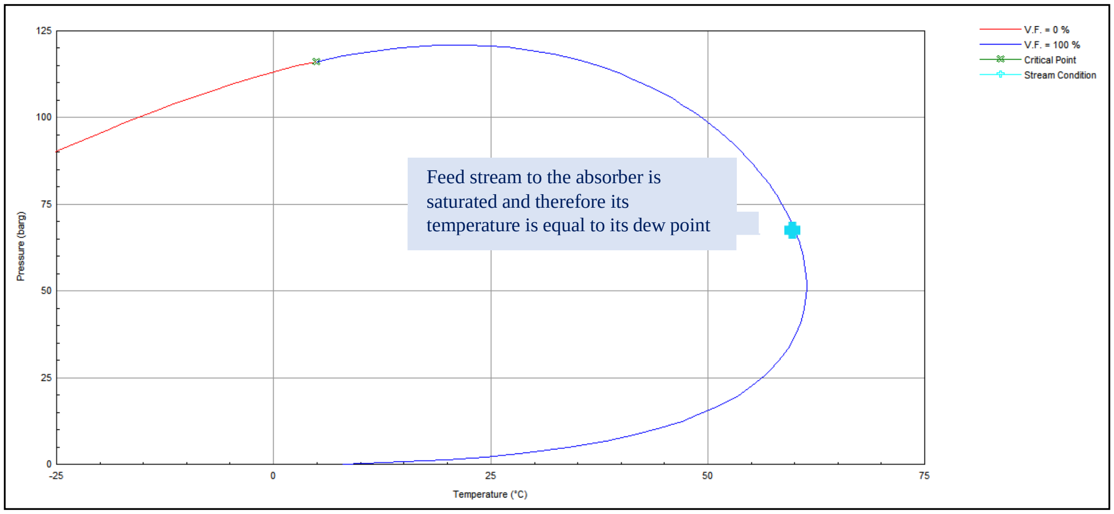


Figure 6: No Dew Point Safety Margin (Scenario: No Sub-Cooling + Superheating)

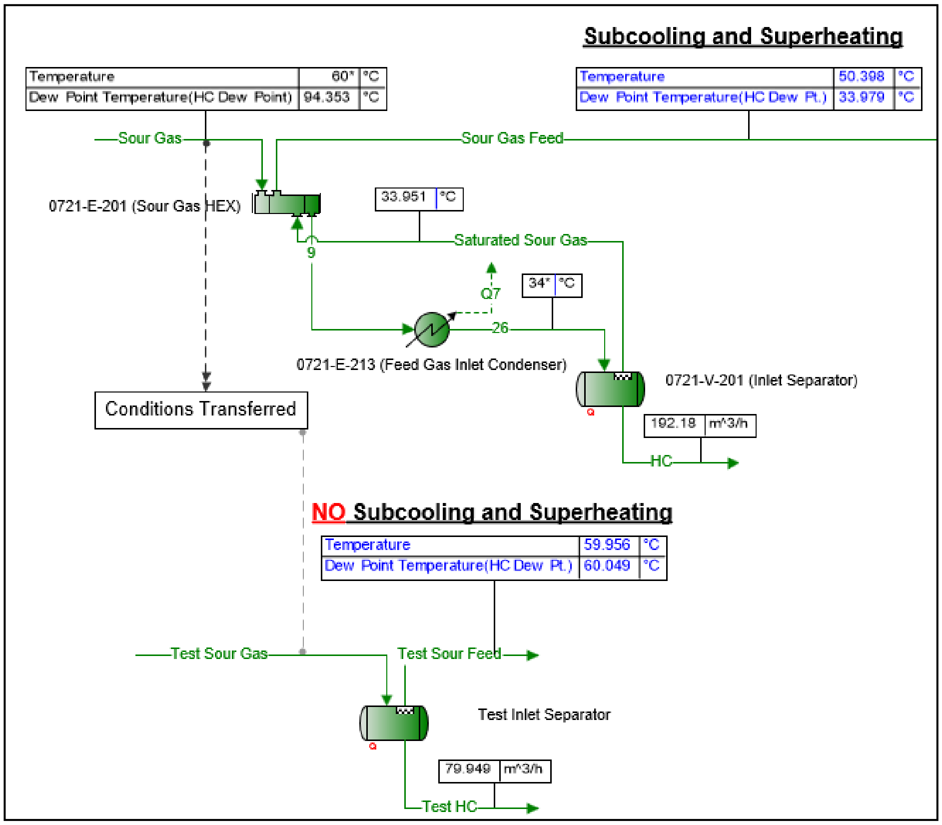
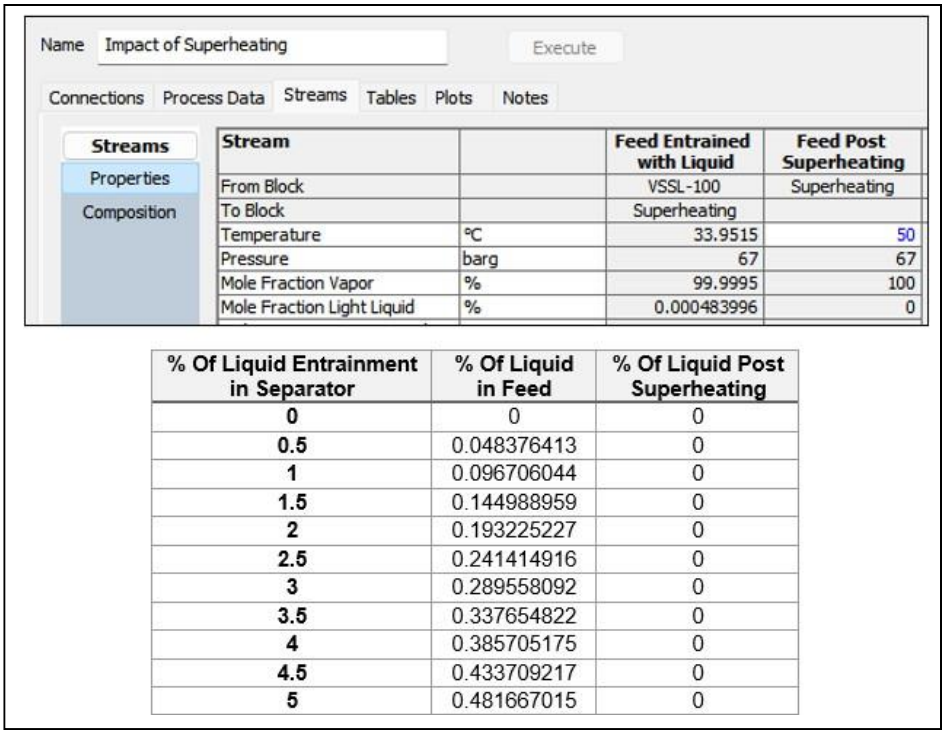


Figure 7: Comparing Scenarios of "Sub-Cooling + Superheating" & "No Sub-Cooling + Superheating"

Another reason why superheating the sour gas is a good design practice is because, in reality, separators (even with internal modifications) cannot achieve ideal / pure separation of phases. Therefore, in live plant operation there will always be some amount of liquid entrainment into the vapor phase resulting in the actual dew point of the sour feed gas stream to be slightly higherthan the theoretical dew point reported by thermodynamic calculations of perfect separations. Superheating the gas can therefore correct this by vaporizing any potential liquid carryover fromthe inlet separator.

A simulation assessment was conducted to validate the benefit of superheating the sour gas before routing to HP absorber in an indicative scenario of imperfect separation. Trace amounts ofliquid entrainment was manually added into the separator model and the results indicated that no liquid fraction was observed in the gas stream at the exchanger outlet even if the liquid carryoveramount increased.



Name: Impact of Superheating

Stream		Feed Entrained with Liquid	Feed Post Superheating
From Block		VSSL-100	Superheating
To Block		Superheating	
Temperature	°C	33.9515	50
Pressure	barg	67	67
Mole Fraction Vapor	%	99.9995	100
Mole Fraction Light Liquid	%	0.000483996	0

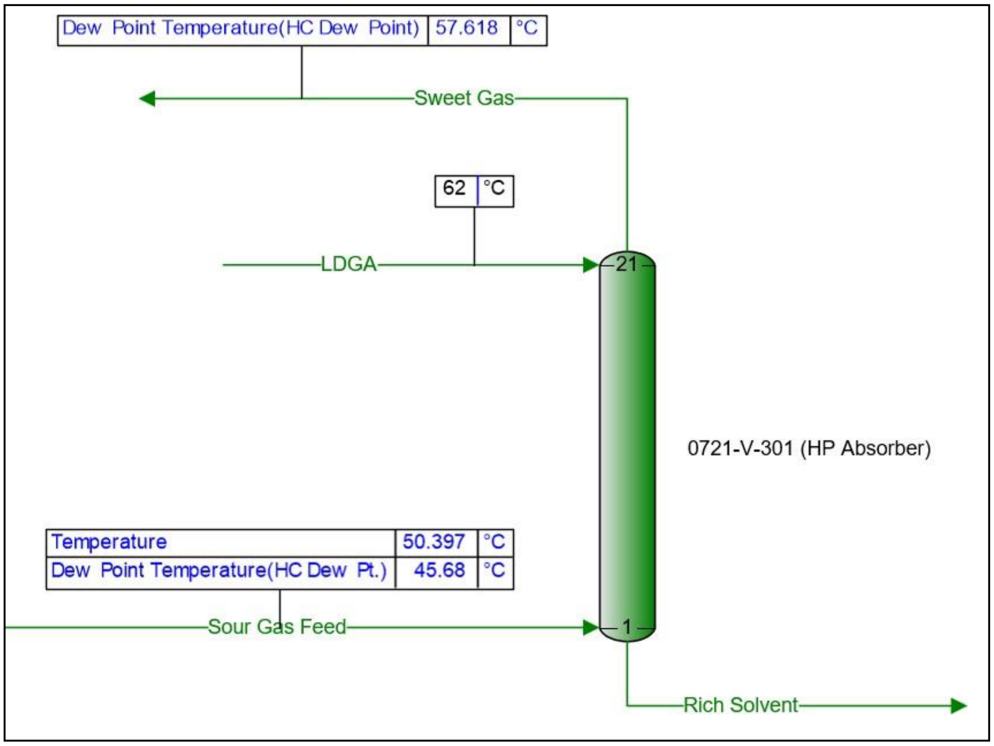
% Of Liquid Entrainment in Separator	% Of Liquid in Feed	% Of Liquid Post Superheating
0	0	0
0.5	0.048376413	0
1	0.096706044	0
1.5	0.144988959	0
2	0.193225227	0
2.5	0.241414916	0
3	0.289558092	0
3.5	0.337654822	0
4	0.385705175	0
4.5	0.433709217	0
5	0.481667015	0

Figure 8: Benefit of Superheating in Scenario of Imperfect Inlet Separation

Recalculation & modification to the conventional 5 deg C temperature approach rule

During the operation and design of amine contactors, it is common practice to maintain a minimum temperature approach of 5 deg C. Temperature approach here is defined as the difference in temperature between the sour feed gas at the bottom and the lean amine solvent at the top. The primary motivation behind this guideline is to prevent hydrocarbon condensation inthe contactor and potential foaming.

When treating ultra-sour feed gases however, the hydrocarbon dew point of the sweet gas leaving the absorber is usually significantly higher than the sour feed entering it. The reason forthis increase is the high quantity of acid gas that is being treated (25% H_2S + 10% CO_2 in ASGoperations). As the acid gas is removed, the remaining hydrocarbons are left at a much higher concentration (partial pressure) in the treated gas than they were in the feed gas. This causes thephase envelope to shift to the right thereby raising the hydrocarbon dew point of the gas and increasing the probability of condensation as it bubbles up through the tower Therefore, this strategy prevents the direct and blanket application of the conventional 5 deg C temperature approach rule. It rather suggests calculating the hydrocarbon dew point temperature of the sweet gas (with low acidic content) and set the lean solvent to be 5 deg C warmer than thattemperature instead. This guideline adds an additional layer of protection by preventing the formation of a second liquid phase in all sections of the contactor, including the top (where the probability of hydrocarbon condensation is the maximum).



Dew Point Temperature(HC Dew Point)	57.618	°C

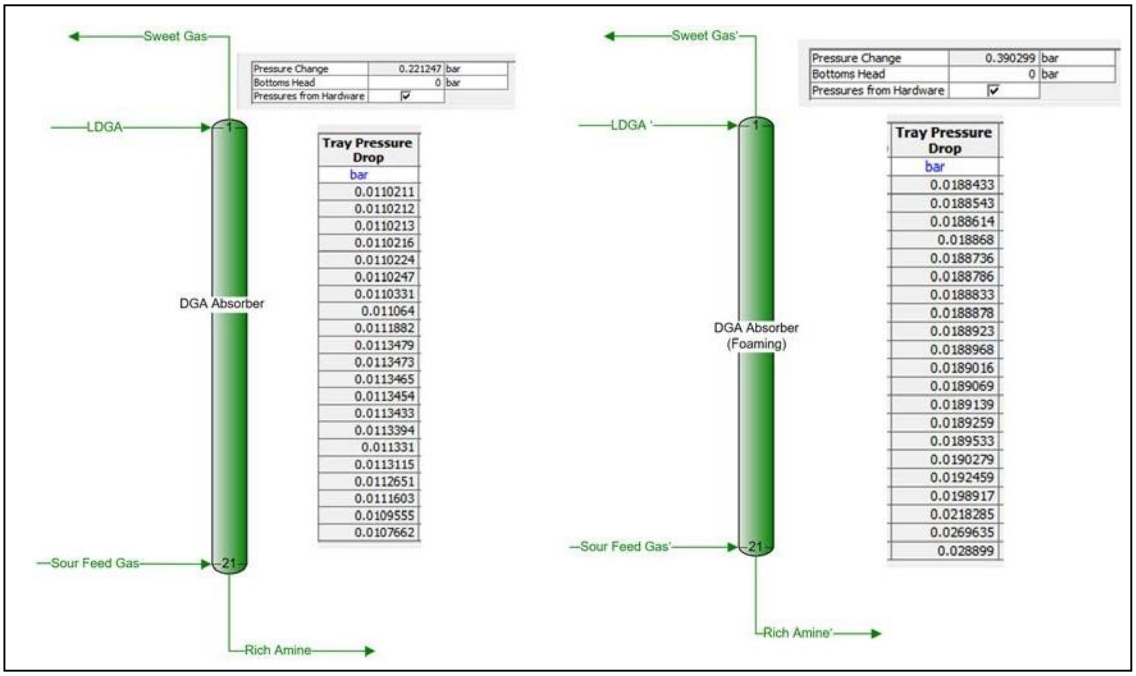
62	°C

Temperature	50.397	°C
Dew Point Temperature(HC Dew Pt.)	45.68	°C

Figure 9: Increase in Gas Dew Point from Bottom to Top of Absorber

Robust differential pressure measurement across absorber

The empty volume occupied by the bubbles formed during foaming leads to a higher pressure drop in a column. In fact, when a column generates a stable foam, the void space generated decreases the column capacity as well. A process simulation model confirms how a column withfoaming and reduced capacity calculates a higher pressure drop.



Pressure Change	0.221247	bar
Bottoms Head	0	bar
Pressures from Hardware	☑	

Tray Pressure Drop
bar
0.0110211
0.0110212
0.0110213
0.0110216
0.0110224
0.0110247
0.0110331
0.011064
0.0111882
0.0113479
0.0113473
0.0113465
0.0113454
0.0113433
0.0113394
0.011331
0.0113115
0.0112651
0.0111603
0.0109555
0.0107662

Pressure Change	0.390299	bar
Bottoms Head	0	bar
Pressures from Hardware	☑	

Tray Pressure Drop
bar
0.0188433
0.0188543
0.0188614
0.018868
0.0188736
0.0188786
0.0188833
0.0188878
0.0188923
0.0188968
0.0189016
0.0189069
0.0189139
0.0189259
0.0189533
0.0190279
0.0192459
0.0198917
0.0218285
0.0269635
0.028899

Figure 10: Spike in Pressure Drop in Foamed Column

One of the most common diagnoses therefore for confirming initiation of foaming is a pressuredrop spike across an absorber. Hence, if the pressure drop reading is wrong it can be difficult tounderstand when to take foaming linked precautionary measures.

The industry usually sticks to installing a digital pressure drop measurement (2 pressure transmitters at top & bottom of the column with a calculated DP reading) device, which is a cheaper, easy install but highly unreliable option. To understand why this is the case, one needs to revisit a few facts. First, the absorbers discussed in this paper operate at high pressures (~70 bar). Second, the usual pressure drop of a normal operating column is around 0.3 bar. For this operating case, range of digital pressure transmitters are 0 - 80 bar. If we apply a 0.2% error factor, it gives an error value of ~0.2 – 0.3 bar per transmitter. This means that the calculated column DP during normal operation may be in the range of 0.1 – 0.5 bar, even when the columnis not foaming. This can make it difficult to detect whether there is actual foaming is occurring or not if one solely relies on the measurements off a digital pressure transmitter.

Thus, to accurately detect small & sensitive pressure drop fluctuations, it is therefore recommended to invest in hard-piped pressure drop transmitter measurements across the column that gives error-free results. These transmitters are considered much more reliable measuring devices with an error percentage that is considerably lower that 0.2%, in relation to its pressure ranges.

Conclusions

The multi-layered bulletproof strategies mentioned in this paper to manage foaming enabledADNOC Sour Gas to finally eliminate the need for inlet coalescers & excessive anti-foam injection – options that were both being originally considered. This resulted in a CAPEX andOPEX saving of US$ 25 MM and US$ 0.6 MMPA respectively.

In addition, the implementation better foaming management strategies around the amine absorberallowed ASG to save on around ~0.3 barg of pressure drop (that would have otherwise been lost due to foaming), corresponding to a reduction in process upsets and delivery of sweet gas at higher pressure to the NGL facility. This resulted in a higher ethane recovery by ~0.5%, thereby generating production related revenues in excess of US$ 2 MMPA.

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