

**Busting a Myth- Analyzing the Effectiveness of
Lean Amine / Sour Gas Differential Temperature Control**

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Introduction

In the vein of a popular TV series in which two “mythbusters” try to prove or debunk urban legends and other folklore, we begin this journey with our own industry “Myth” to either be proven true or busted. Fortunately, we will not need gel-like people to take bullets or be smashed in falling elevators to examine the circumstances around our topic. For the sake of our “prove or bust a myth” theme however, we offer one of the sacred truths of our industry as our “myth” and then ask-can it be “proved” or “busted”?

Let us take our “myth” directly from one of the leading authorities in gas treating, **Gas Purification**, 4th ed., Kohl and Riesenfeld, Gulf Publishing, Houston, 1985, pg. 125:

“Condensation of light hydrocarbon constituents of the feed gas can be avoided by keeping the temperature of the lean-amine solution about 10° to 15°F above that of the feed gas, thus ensuring that no phase change occurs.”

While this says light hydrocarbon, we will take this to mean any hydrocarbon with a boiling point greater than pentanes/pentenes at the conditions in an absorber.

From most texts or presentations on the subject, it is widely known that the accumulation of these higher boiling point hydrocarbons in amine systems leads to foaming, foaming leads to instability, and instability leads to lost production. To control foaming, the amine system operator must try to prevent these hydrocarbons from entering the amine solution, or remove them from the amine solution once they have entered.

To prevent high boiling point hydrocarbon contamination, liquid/vapor separation devices are placed upstream of amine absorbers first to separate the bulk phases, second to eliminate droplet entrainment, and third to eliminate aerosol droplet entrainment. After assuring the removal of entrained liquid, operators apply the technique of **lean amine/sour gas differential temperature control** to prevent condensation of any of the remaining high boiling point hydrocarbons in the vapor phase by keeping the hydrocarbons above their dew point. It is assumed that if the gas enters the amine from a liquid/vapor separation device, the vapor is at its dew point. Any drop in the temperature of the gas past that point will cause the condensation of material from the gas phase. Prevention of hydrocarbon liquid entrainment and condensation should insulate the amine system from foaming.

Data from several amine systems that have many, if not all of the hydrocarbon contamination prevention measures mentioned above in operation including lean amine/sour gas differential temperature control are examined. The data show significant foaming problems exist no matter what the differential temperature between amine and sour gas. It also shows a wide variety of dissolved hydrocarbons found in the amine solutions and regenerator overhead systems. Clearly, these data show our “myth” to be **busted**. After presenting the data for busting the myth, we delve into the reasons why the heavy hydrocarbons entered the amine and caused the foam problems, and look at the varying solutions to the hydrocarbon contamination problems in each case.

Operations Cases Used to Verify the Effectiveness of Lean Amine/ Sour Gas ΔT Control

Case One

Location: Refinery A

Treatment scenario: A 10-25 MMSCFD fuel gas absorber operating at 180 to 225 psig, up to 300 gpm of MDEA at 40-45 wt%. The entire system is composed of multiple absorbers and two regenerators including LPG, hydrogen, and other fuel gas treating services. The system is equipped with a slip stream regenerable hydrocarbon removal and particle filtration system. None of the other absorbers or the regenerators suffered from foaming during this period.

Treating specification: Required performance 160 ppm H_2S , nominal performance <20 ppm H_2S

Figure C1-A: Fuel Gas Treater and auxiliary equipment

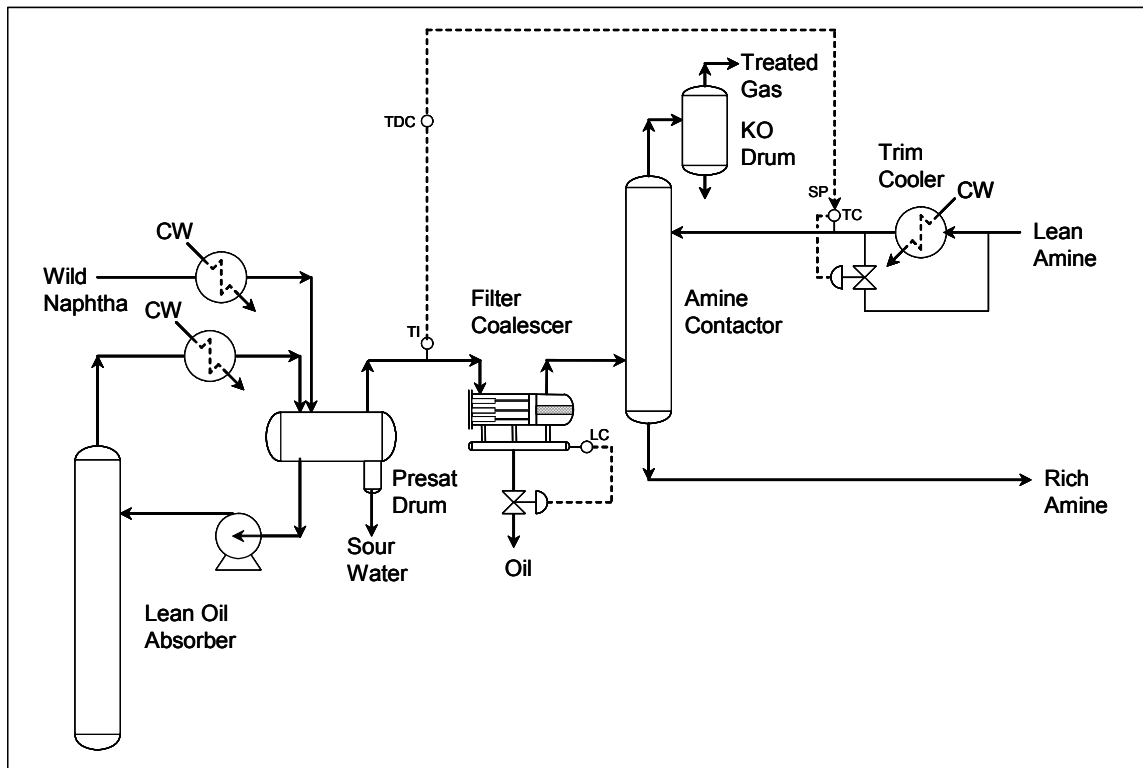
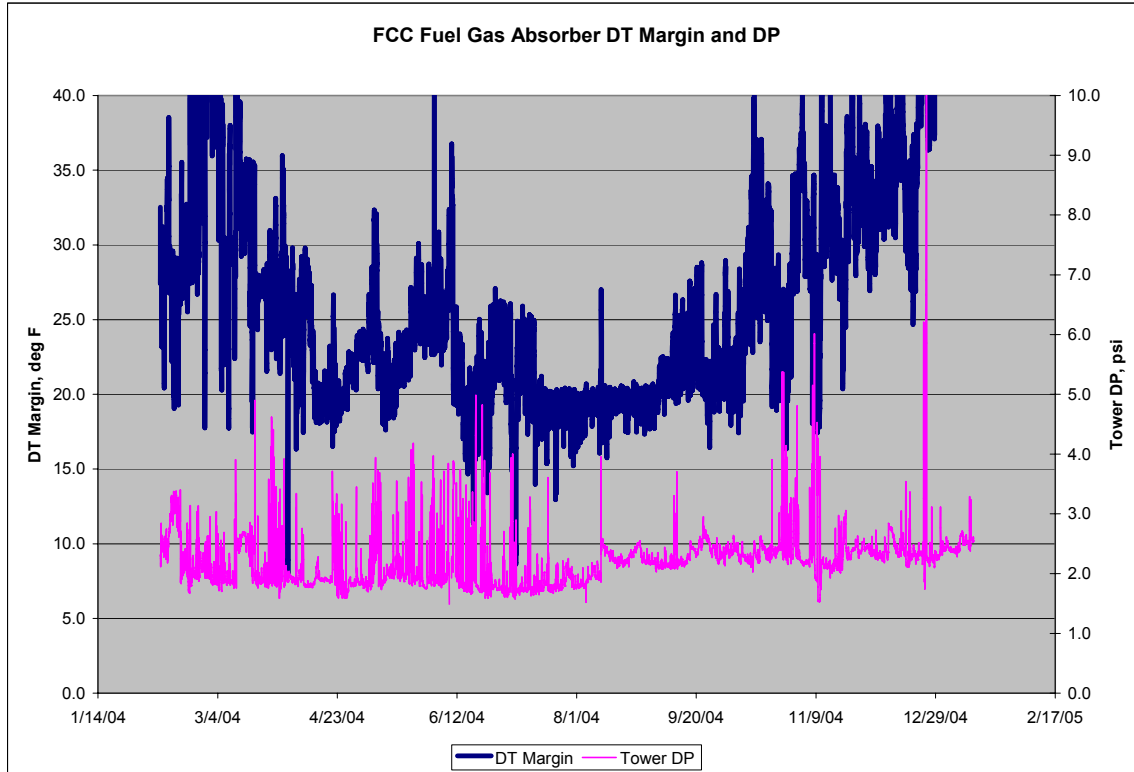


Figure C-1B: Operating data

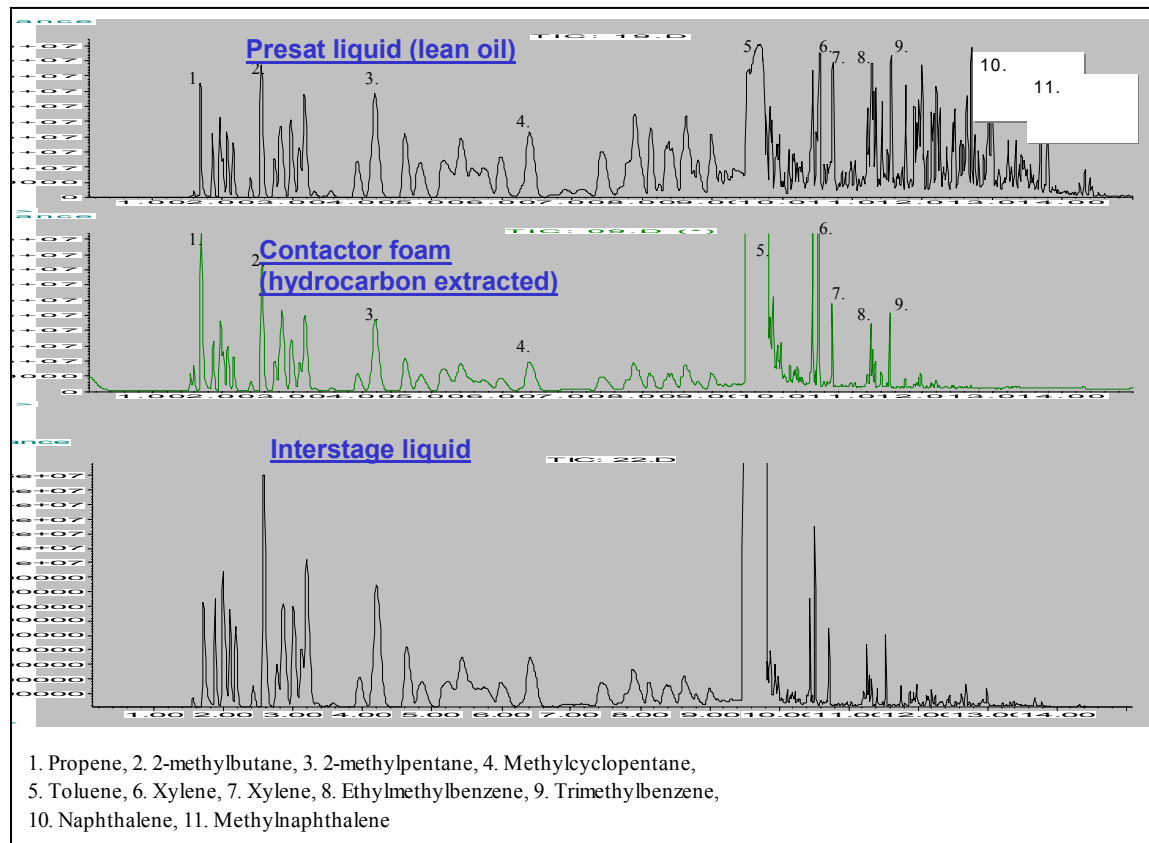


Operations situation: The performance of this tower was erratic at best. The tower would foam over even though it operated at as little as 25% of jet flood in the trays. At one point in the operation, amine entered the absorber at 80+ gpm for a period of several hours while the level control valve stayed shut on the rich amine outlet line. Approximately 10,000 gallons of amine solution were locked up in foam in the tower over that period of time, while operating with a lean amine/sour gas differential temperature of 20-25 degrees.

Table C1-1: Hydrocarbon Quantities in Lean & Rich Amine During an Upset

Sample	Hydrocarbon in Amine, ppmw
Lean Amine	50
Rich Amine from Foaming Contactor	5578

Figure C1-C: GC-Mass Spectrum Comparison of Foam Collected from Contactor Overhead KO Drum vs Several FCC Gas Plant Liquids



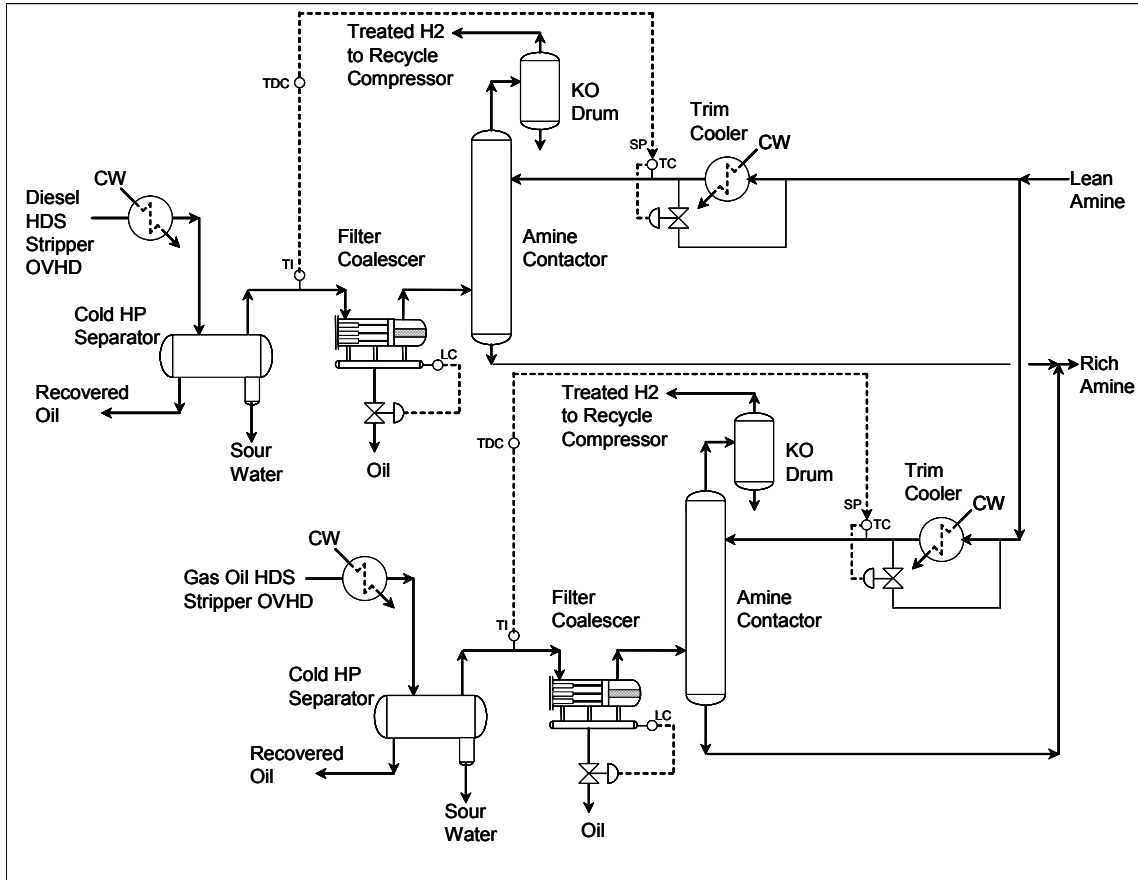
Case Two

Location: Refinery A

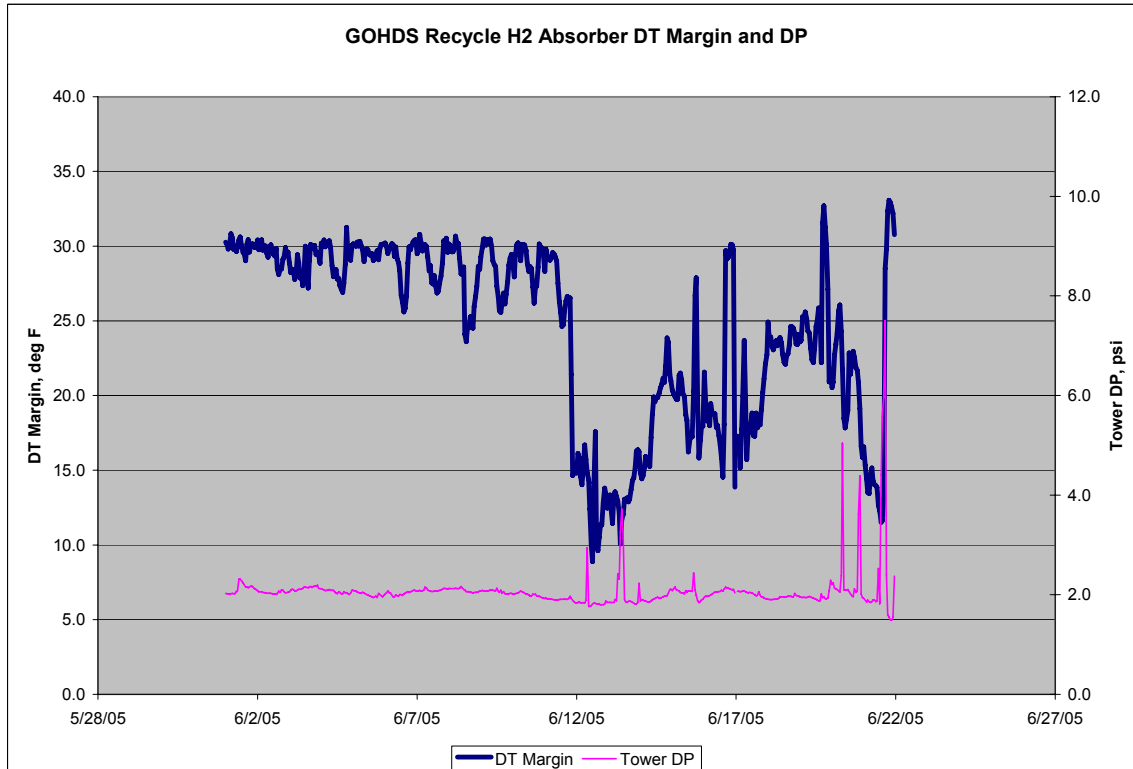
Treatment scenario: Two 50 MMSCFD hydrogen recycle absorbers operating at 1100 psig, approximately 160 gpm of MDEA at 40-45 wt% per absorber. The system was composed of the hydrotreater absorbers and a regenerator. The system was equipped with a slip stream regenerable hydrocarbon removal and particle filtration system. The regenerator did not suffer from foaming during this period.

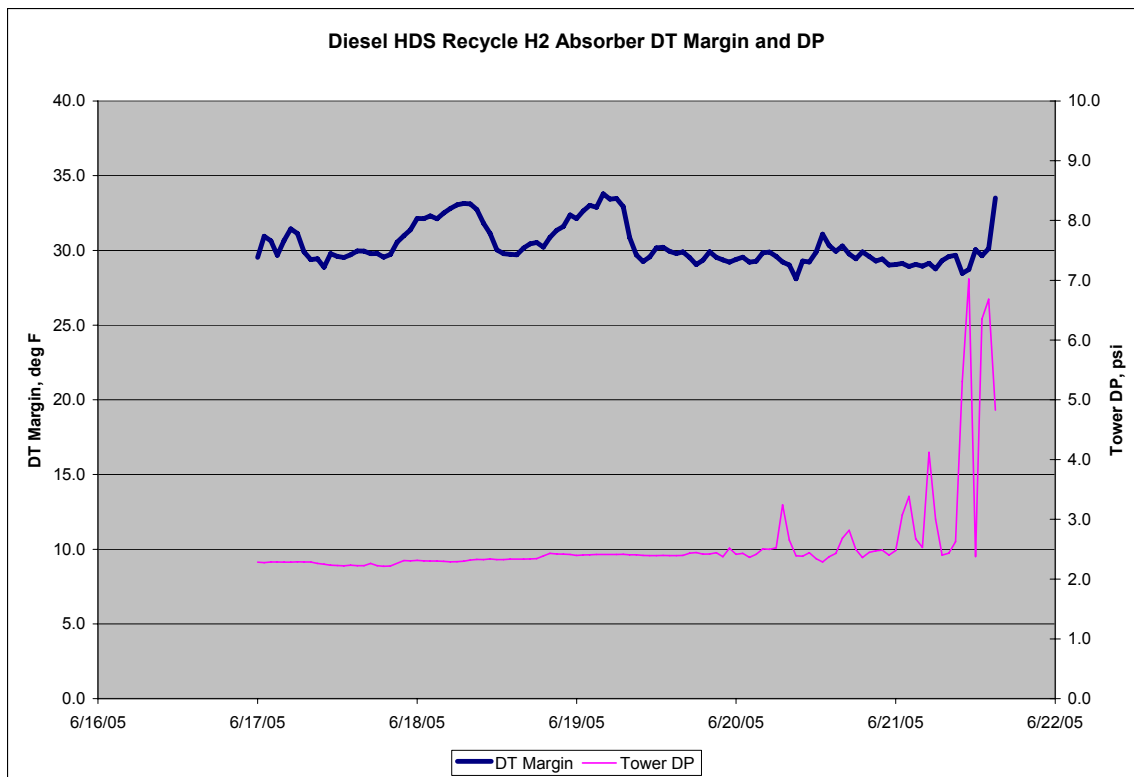
Treating specification: No hard specification, nominal performance <6 ppm H₂S

Figure C2-A: Recycle Hydrogen absorbers and auxiliary equipment



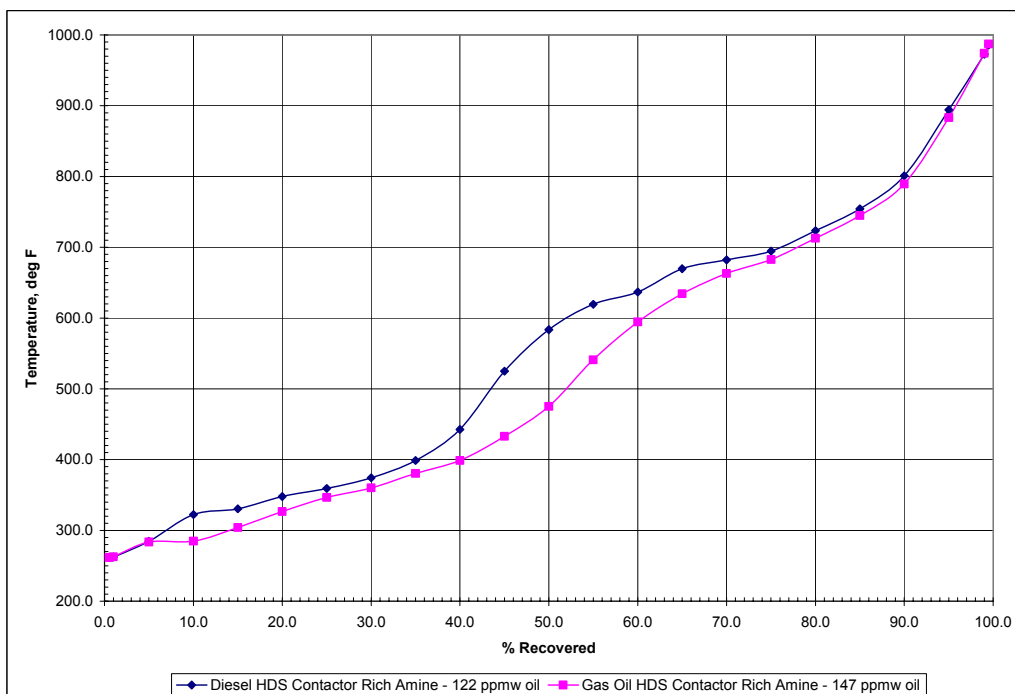
Figures C2-B: Operating Data





Operations situation: These absorbers suffered from a dose of liquid hydrocarbon carried into the amine during a startup. After fixing problems with the filter/coalescer units upstream of the amine contactors, data was taken on the amine from the tower during foaming conditions. The amine was extracted with toluene and the hydrocarbons were analyzed by simulated distillation (GC). Distillation data of the oil found in the rich amine samples is presented in Figure C2-C. The lean amine was found to contain 33 ppmw oil at this time.

Figure C2-C: Distillation data for oil extracted from rich amine samples



Case Three

Location: Refinery B

Treatment scenario: Two FCC fuel gas absorbers, a hydrogen absorber, and regenerator. The absorbers did not suffer from foaming during this period.

Figure C3-A: FCC Gas Treating Arrangement

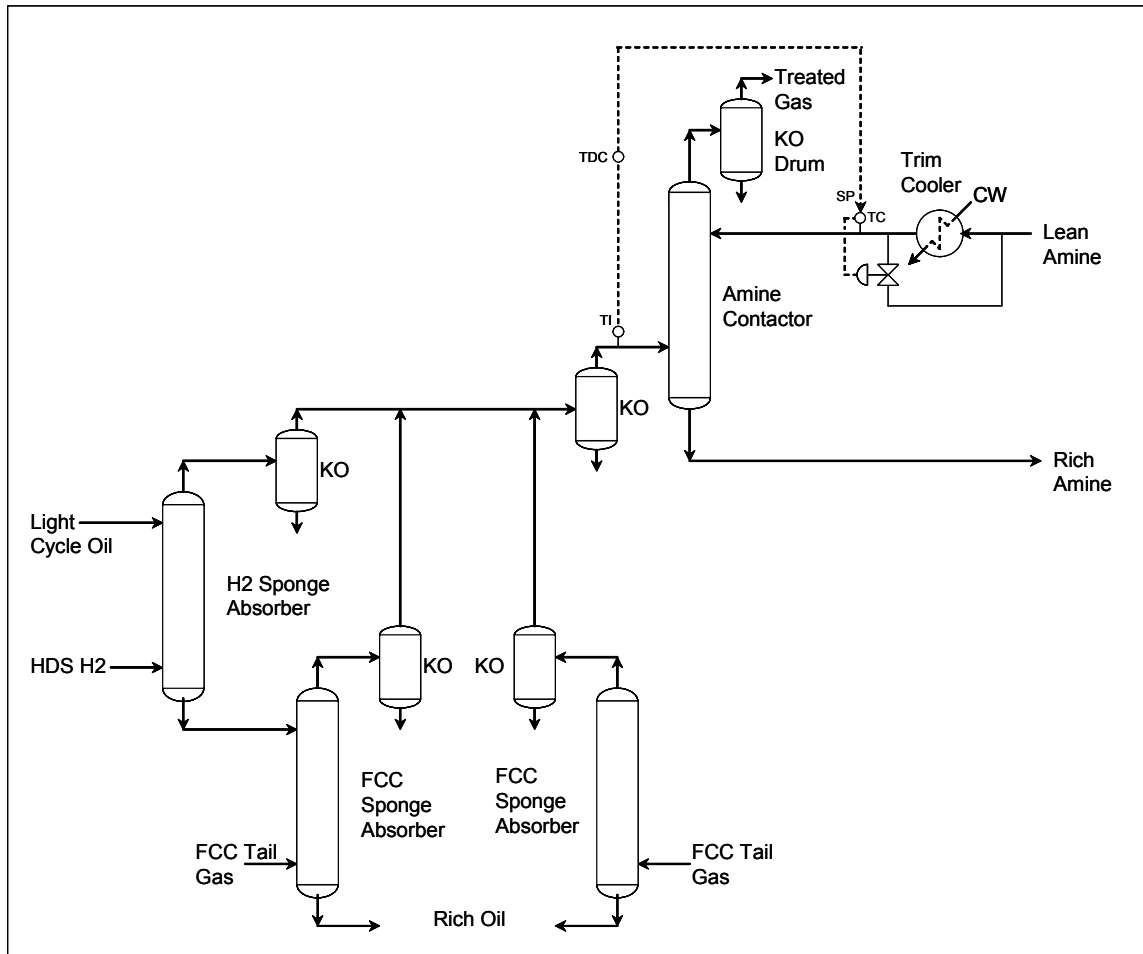
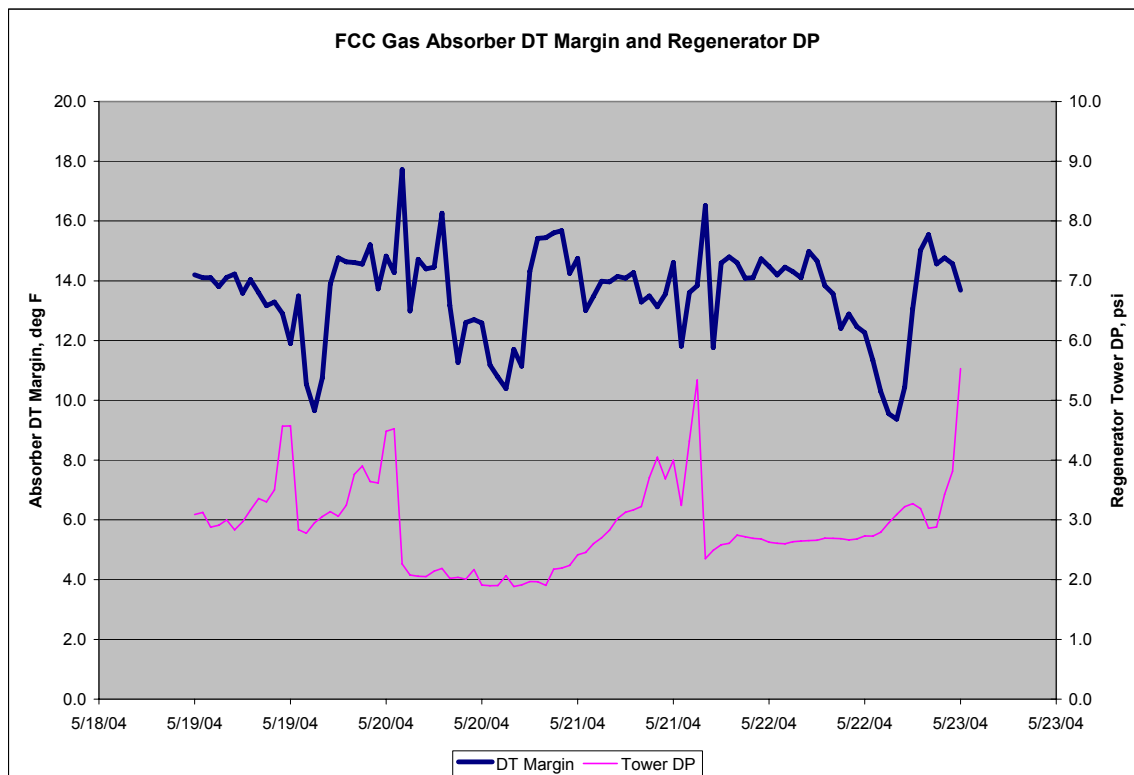


Figure C3-B: Operating data during a typical upset



Operations situation: The regenerator would foam with varying frequency. The amine unit is equipped with slipstream lean amine charcoal filtration but does not contain a rich amine flash drum. The regenerated lean amine from this unit has never historically showed abnormally high levels of hydrocarbon.

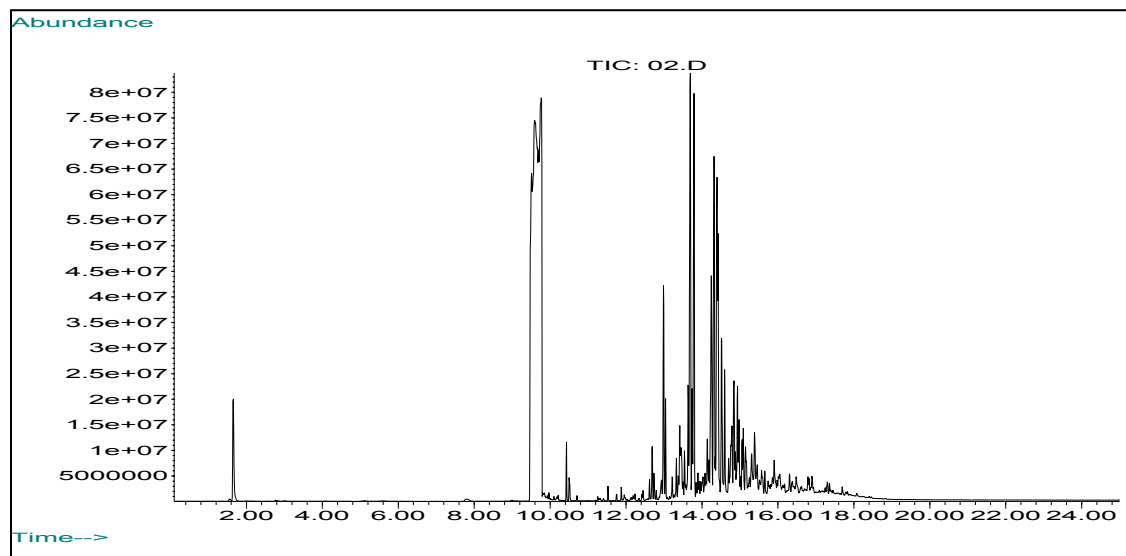
Samples of several streams were taken when the regenerator was foaming. These included the lean amine, rich amine and the reflux water from the overhead system.

The reflux water showed an unusually high concentration of hydrocarbons as shown in Table 3C-1. Further analysis using GC mass spectroscopy identified the majority of the oil present in the reflux water to be naphthalene substituted aromatics as indicated in Figure C3-C.

Table 3C-1: Oil content during a regenerator foaming episode

Sample	Hydrocarbon in Amine, ppmw
Lean Amine	2
Rich Amine	7
Regenerator Reflux Water	1625

Figure C3-C: GC Mass Spectrum of Oil Extracted from Regenerator Reflux Water



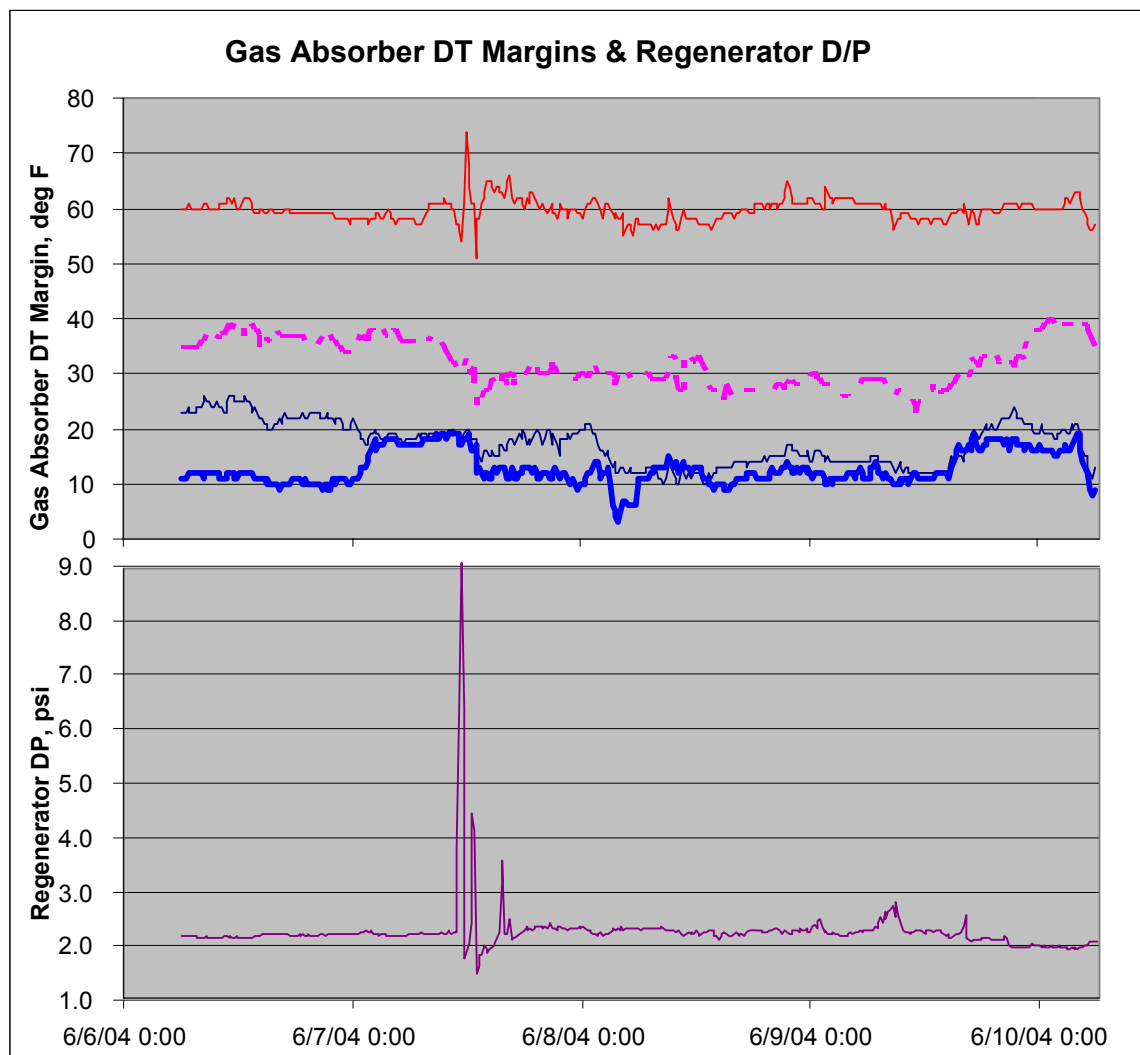
Reflux H₂O Contaminants: Naphthalene; methyl-naphthalene isomers, dimethyl-naphthalene isomers, trimethyl-naphthalene isomers

Case Four

Location: Refinery C MDEA Regenerator

Treatment scenario: FCC fuel gas absorber, hydrogen absorber, and LPG absorber connected to one regenerator. None of the absorbers suffered from foaming during this period. The absorbers are all packed, the regenerator is trayed.

Figure C4-B: Typical operating data



Operations situation: The regenerator would foam with varying frequency. Lean amine from the regenerator did not show an abnormally high level of hydrocarbon. The foaming problem prevented purging the reflux for oil contamination on a regular basis.

We sampled several streams when the regenerator was foaming. These included the lean amine and the reflux water from the overhead system.

The reflux water showed an unusually high concentration of naphthalenes as shown in Table C4-1.

Table C4-1: Hydrocarbon Content

Sample	Hydrocarbon in Amine, ppmw	Hydrocarbon Characterization
Lean Amine	19	C14-C30 n-paraffins
Reflux Water	46	C9-C40 naphthalenes

Now that we have seen four widely varying instances where lean amine/sour gas differential temperature control was in use and saw that in all the cases some part of the amine system started foaming, we can say with confidence that using this method of control will not prevent the accumulation of high boiling hydrocarbons from the gas in the amine. We have sufficient evidence now to say our “myth” is busted.

Why the “Myth” is Busted

Having seen the operations data, we now dig deeper to find out why this technique did not afford the protection advertised to the industry for so many years. On the case by case basis that we have presented the data, we will examine the underlying reasons that our ‘Myth’ has become unraveled.

Case One

We sampled several streams when the absorber was foaming. These included the lean amine, rich amine from the absorber, foam liquid from the overhead knock out drum on the absorber, the pre-saturation drum (gas source) liquid, and gas plant compressor interstage knock out drum liquid.

We can see that the lean amine contained a low amount of hydrocarbon, but the rich amine had a considerable amount of dissolved hydrocarbon, mostly aromatic gasoline components in the C9-C14 boiling range. The chromatograms in Figure C1-C also show the oil in the foam closely matched the liquid knocked out of the gas phase in the compressor inter-stage and did not have the heavier components from the liquid in the pre-saturation drum. This was a clear indication that the hydrocarbons found in the amine were not simply entrained liquid.

We simulated the operation of the pre-saturation drum upstream of the fuel gas contactor by back-blending the feed using GC data collected on the gas and data from a SimDist of the liquids from the pre-saturation drum using both the Peng Robinson Stryjek-Vera and Kabadi Danner property packages. The results are shown in Table C1-2 below.

Table C1-2: Back-blended FCC Contactor Feed Gas Reconstructed from Feeds

Temperature		122 F					
Pressure		236.2 psia					
Flowrate		11.9 MMSCFD					
Peng Robinson Stryjek-Vera Method				Kabadi-Danner Method			
Component	Mole Fr.	Component	Mole Fr.	Component	Mole Fr.	Component	Mole Fr.
H2S	0.021	n-Hexane	0.008951	H2S	0.021	n-Hexane	0.008957
CO2	0.015	Benzene	0.001605	CO2	0.016	Benzene	0.001592
CO	0.008	NBP[0]193*	0.001570	CO	0.008	NBP[0]193*	0.001518
H2O	0.008	NBP[0]213*	0.001335	H2O	0.008	NBP[0]213*	0.001284
Nitrogen	0.045	NBP[0]239*	0.000518	Nitrogen	0.046	NBP[0]239*	0.000494
Hydrogen	0.262	NBP[0]264*	0.000269	Hydrogen	0.266	NBP[0]264*	0.000254
Methane	0.352	NBP[0]290*	0.000155	Methane	0.357	NBP[0]290*	0.000145
Ethane	0.112	NBP[0]315*	0.000084	Ethane	0.115	NBP[0]315*	0.000078
Ethylene	0.070	NBP[0]340*	0.000047	Ethylene	0.071	NBP[0]340*	0.000043
Propane	0.009	NBP[0]365*	0.000023	Propane	0.008	NBP[0]365*	0.000021
Propene	0.034	NBP[0]390*	0.000010	Propene	0.023	NBP[0]390*	0.000009
i-Butane	0.015	NBP[0]415*	0.000005	i-Butane	0.002	NBP[0]415*	0.000004
1-Butene	0.002	NBP[0]438*	0.000001	1-Butene	0.003	NBP[0]438*	0.000001
i-Butene	0.002	NBP[0]466*	0.000000	i-Butene	0.001	NBP[0]466*	0.000000
n-Butane	0.009	Total Heavies	0.014573	n-Butane	0.003	Total Heavies	0.014402
i-Pentane	0.011			i-Pentane	0.014		
1-Pentene	0.001			1-Pentene	0.016		
n-Pentane	0.006			n-Pentane	0.004		
Total Light Ends	0.985			Total Light Ends	0.982		

We can see that there is a significant portion of gasoline and heavier components in the vapor phase. In liquid flow terms, the heavy end C₆⁺ tail is equivalent to 150-160 BPD. It is interesting to note that the gas chromatograph only picked up about 1/3 of the total C₆⁺ equivalent concentration. We then took this stream and looked at what happens during condensation of this stream with 2 degrees of heat loss. Results are shown in Table C1-3.

Table C1-3: Three Phase Flash Results for FCC Contactor Feed Gas

Phase	Overall	Vapor	Hydrocarbon	Aqueous
Molar Flow, lbmol/hr	1303.76	1303.08	0.122	0.560
Component	Mole Fract	Mole Fract	Mole Fract	Mole Fract
H2S	0.020583	0.020593	0.008804	0.000276
CO2	0.015446	0.015454	0.002319	0.000057
CO	0.008188	0.008192	0.000194	0.000000
H2O	0.008013	0.007588	0.002797	0.999234
Nitrogen	0.045251	0.045275	0.000910	0.000003
Hydrogen	0.260645	0.260781	0.003354	0.000011
Methane	0.352023	0.352205	0.021607	0.000084
Ethane	0.114559	0.114616	0.033191	0.000056
Ethylene	0.070520	0.070556	0.014214	0.000010
Propane	0.009440	0.009444	0.008426	0.000002
Propene	0.033667	0.033682	0.026734	0.000010
i-Butane	0.015252	0.015257	0.030385	0.000001
1-Butene	0.002294	0.002295	0.005410	0.000001
i-Butene	0.001882	0.001882	0.004324	0.000001
n-Butane	0.009209	0.009212	0.025351	0.000002
i-Pentane	0.011163	0.011163	0.068077	0.000001
1-Pentene	0.001330	0.001330	0.008867	0.000001
n-Pentane	0.006133	0.006132	0.048168	0.000002
n-Hexane	0.008955	0.008942	0.195512	0.000004
Benzene	0.001595	0.001591	0.045284	0.000158
NBP[0]193*	0.001518	0.001512	0.067621	0.000038
NBP[0]213*	0.001284	0.001277	0.083535	0.000030
NBP[0]239*	0.000494	0.000490	0.053087	0.000010
NBP[0]264*	0.000254	0.000250	0.045664	0.000004
NBP[0]290*	0.000145	0.000141	0.043594	0.000002
NBP[0]315*	0.000078	0.000075	0.039869	0.000001
NBP[0]340*	0.000043	0.000040	0.036839	0.000000
NBP[0]365*	0.000021	0.000018	0.029525	0.000000
NBP[0]390*	0.000009	0.000007	0.022057	0.000000
NBP[0]415*	0.000004	0.000003	0.014741	0.000000
NBP[0]438*	0.000001	0.000001	0.006685	0.000000
NBP[0]466*	0.000000	0.000000	0.002853	0.000000

The simple cooling and separation computation on the gas stream leaving the pre-saturations drum shows that two phases will form from the gas as it condenses, a hydrocarbon liquid phase and a water phase. The outcome of this computation would tend to support the use of the lean amine/sour gas differential temperature control. However, the water phase contains some dissolved hydrocarbons, but more notably shows that aromatic gasoline components accumulate in the aqueous phase in greater quantity as was noted in the chromatograms in Figure C1-C.

Thus, the simulation points out something that we as an industry simply miss or ignore, that is that many of the hydrocarbon components in these gases are simply soluble in the water in the amine solution. Even worse, data collected and presented in an article by Richon et al¹ shows that the presence of amine enhances the solubility of these components by a factor of 3-10 times. Using the lean amine/sour gas differential temperature control, we certainly would have prevented bulk hydrocarbon condensation, but this control strategy appears to have no effect on simply dissolving these materials from the gas phase directly into the amine solution.

As an industry, we ignore this largely because hand computations or even computer simulations of these systems are done using pseudo-components with poorly defined water solubility, or because we tend to lump that last little fraction of hydrocarbon that is heavier than pentane into “C₆⁺”, generally simulated as hexane for simplicity. A particular reason for this is that most gas chromatographs that are employed for stream analysis for this type of service concentrate on the hydrocarbon economic value of the stream (purity, heating value, recoverable liquids) and less on the hydrocarbon impurity tail. We tend to look only at non-hydrocarbons as the impurities and measure them to sometimes excruciatingly low ppm or ppb levels while measuring hydrocarbon impurities to that “utility system” down to a mere 0.1 mole % (1000 ppm) and then putting them into a “lump”.

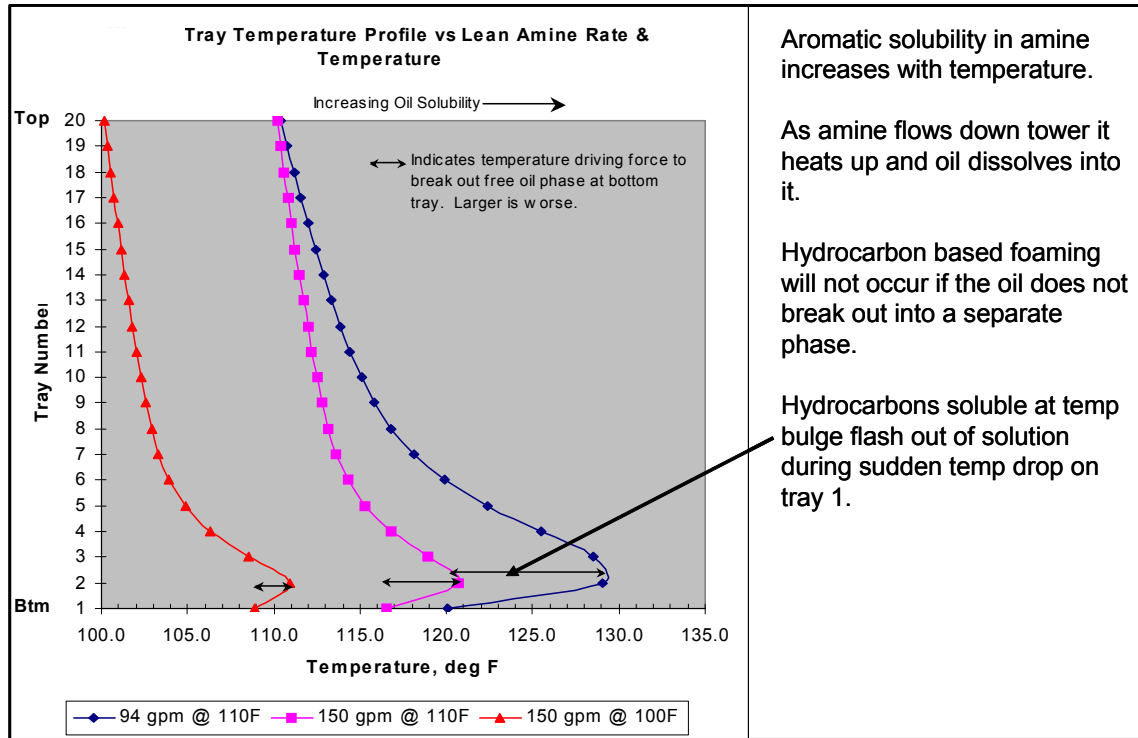
Let’s consider the effect of this choice not to measure or differentiate the “C₆⁺” in gas streams on an amine system. Suppose that in an amine treater in a natural gas service of 300 MMSCFD of gas that there is 10 ppm of benzene. Some simple math shows that we have $(300 \times 10^6) \times (10 \times 10^{-6})$ SCFD of benzene in the gas, or 3,000 SCFD. On a daily basis this is 7.9 lb. moles of benzene in the gas per day going to the amine contactor or 616 lb per day or about 2 BBL/day. When it comes to an amine system or a sulfur plant downstream, picking up this much hydrocarbon can have a significant impact on operations from amine foaming to Claus catalyst deactivation to air demand swings in the unit. If we depend on carbon picking this hydrocarbon up, remember that most activated carbon will hold about 5% of its weight in hydrocarbon. If the carbon had to remove 1/10th of the total of this hydrocarbon, we would need 1232 lb of carbon a day to keep the system free of the effects of hydrocarbon accumulation.

Expecting 6 months run length on that carbon bed? It will have to be about 225,000 lb of carbon or a vessel about 17 ft. diameter by 34 ft. length!

If we now consider water solubility and amine enhancement of water solubility for some heavier hydrocarbons, and we consider that these materials follow the general laws of solubility, namely that higher temperatures increase solubility, it should be evident that ***DT control can add to the problem.*** Our nature usually tells us that if some is good, than more should be better. However, as alcohol consumption and anti-foam addition have taught mankind over the years, more is not always better.

Reviewing Figure C1-D, we can see the effect of inlet amine temperature on the “bulge”, the rapid increase and decrease of temperature inside of a typical amine absorber caused by the rapid reaction of the acid gas with the amine near the bottom of the absorber. Some may make the assumption that running lean amine hotter will put a larger cushion between the condensation temperature and temperature inside the column. What appears to make things worse in this case is that the increase in temperature increases the solubility of contaminant hydrocarbons in the amine, especially where we see the “bulge”. As the amine cools off from the peak of the bulge, the solubility of the hydrocarbon should decrease. If the hydrocarbon that is dissolved reaches a point where it begins to concentrate near the gas/amine solution interface, the concentration gradient set up at the interface can make the property changes in the amine solution surface required to support rapid, stable foam growth.

Figure C1-D: Effect of Lean Amine Temperature & Flowrate on Hydrocarbon Solubility



Aromatic solubility in amine increases with temperature.

As amine flows down tower it heats up and oil dissolves into it.

Hydrocarbon based foaming will not occur if the oil does not break out into a separate phase.

Hydrocarbons soluble at temp bulge flash out of solution during sudden temp drop on tray 1.

Case Two

We can see in Figure C2-C that we have a wide boiling range of hydrocarbons found in the amine. We again looked to a simulation of the separation drum upstream of the amine unit to see why these heavy oils were being transported to the amine. One would expect at high pressures and low temperatures that heavy hydrocarbons would tend to stay in the condensed phase.

There are two key issues identified in this simulation as demonstrated in Table C2-2. First, there is a significant “tail” of heavy hydrocarbons in the hydrogen going to the absorbers. The tail in the gas oil hydrotreater hydrogen extends into the diesel boiling range. Following our math above, we can see that 2090 ppm of heavy oil in 48.1 MMSCFD of hydrogen treated can yield over a 98 BBL/day of oil for the amine to try to absorb. Because these hydrotreaters operate on cracked stocks, and the potential for more water soluble aromatic compounds exists, we see that there is considerable potential for picking up the more water soluble heavy hydrocarbons available. Second, we see in Table C2-3 that cooling the gas produces water as the primary condensate, not hydrocarbon. Both these key issues point to the ineffectiveness of the lean amine/sour gas differential temperature control in this situation.

Table C2-2: Back-blended GOHDS Contactor Feed Gas Reconstructed from Feeds

Temperature	94	F	
Pressure	1127	psia	
Flowrate	48.1	MMSCFD	
Component	Mole Fr.	Component	Mole Fr.
H2S	0.006	n-Hexane	0.001942
Nitrogen	0.131	NBP 169F	0.000035
Hydrogen	0.704	NBP 194F	0.000029
H2O	0.001	NBP 220F	0.000022
Methane	0.098	NBP 241F	0.000017
Ethane	0.042	NBP 268F	0.000016
Propane	0.012	NBP 292F	0.000012
i-Butane	0.001	NBP 317F	0.000009
n-Butane	0.002	NBP 342F	0.000004
i-Pentane	0.001	NBP 367F	0.000002
n-Pentane	0.001	NBP 392F	0.000001
		NBP 416F	0.000001
Total Light Ends	0.998	Total Heavies	0.002090

Table C2-3: Three Phase Flash Results for GOHDS Contactor Feed Gas

Phase	Overall	Vapor	Aqueous
Molar Flow, lbmol/hr	5285.90	5285.59	0.31
Component	Mole Fract	Mole Fract	Mole Fract
H2S	0.005969	0.005970	0.000415
Nitrogen	0.130629	0.130636	0.000045
Hydrogen	0.703503	0.703544	0.000092
Methane	0.098390	0.098396	0.000090
Ethane	0.041959	0.041962	0.000083
Propane	0.011829	0.011830	0.000009
i-Butane	0.000962	0.000962	0.000000
n-Butane	0.001892	0.001892	0.000002
i-Pentane	0.000879	0.000879	0.000000
n-Pentane	0.000846	0.000846	0.000001
n-Hexane	0.001942	0.001942	0.000004
H2O	0.001053	0.000994	0.999256
NBP 169F	0.000035	0.000035	0.000000
NBP 194F	0.000029	0.000029	0.000001
NBP 220F	0.000022	0.000022	0.000001
NBP 241F	0.000017	0.000017	0.000000
NBP 268F	0.000016	0.000016	0.000000
NBP 292F	0.000012	0.000012	0.000000
NBP 317F	0.000009	0.000009	0.000000
NBP 342F	0.000004	0.000004	0.000000
NBP 367F	0.000002	0.000002	0.000000
NBP 392F	0.000001	0.000001	0.000000
NBP 416F	0.000001	0.000001	0.000000

Case Three

In this case we felt that the system should have been protected from the water soluble hydrocarbons because all the absorber gases were being cleaned by sponge oil that was designed to remove heavy oils from the gas phase.

Referring to Figure C3-A, we see that the sponge oil for the one of the FCC tail gas absorbers, light cycle oil or LCO from the FCC, goes first to the hydrogen absorber. Apparently, it was felt that the sponge oil could do “double duty” by eliminating the small amount of heavy oil in the hydrogen before sponging the gasoline out of the FCC tail gas. We did not observe the gasoline aromatics in the reflux water that we saw in Case One; rather, we observed the naphthalenes and substituted naphthalenes shown in Figure C3-C.

It appears that the mechanism of lifting heavy oils into the gas phase by high hydrogen partial pressure caused some of the material from the LCO sponge oil used in the hydrogen absorber to enter the hydrogen before going to the amine absorber. Using a solubility correlation reference 2, we would expect a fairly substantial solubility of naphthalenes and substituted naphthalenes in water (50-150 ppmw) at the conditions present in the amine system. Thus, driving force would exist for the amine to remove some of this material from the hydrogen feed gas stream. Entering as dissolved materials in the rich amine to the regenerator, they were steam stripped up the column into the condenser. Here they were condensed and returned to the regenerator, building up until an oil concentration gradient began to form at the amine solution surface to trigger the foaming episodes.

Case Four

In this case, we found naphthalenes again in the regenerator reflux water. LCO was not being used on the hydrogen stream, but was being used as sponge oil for the FCC tail gas prior to amine treatment. A mechanism similar to that for Case 3 could be postulated based upon this evidence.

Going beyond the “Myth”-lessons learned and learning applied

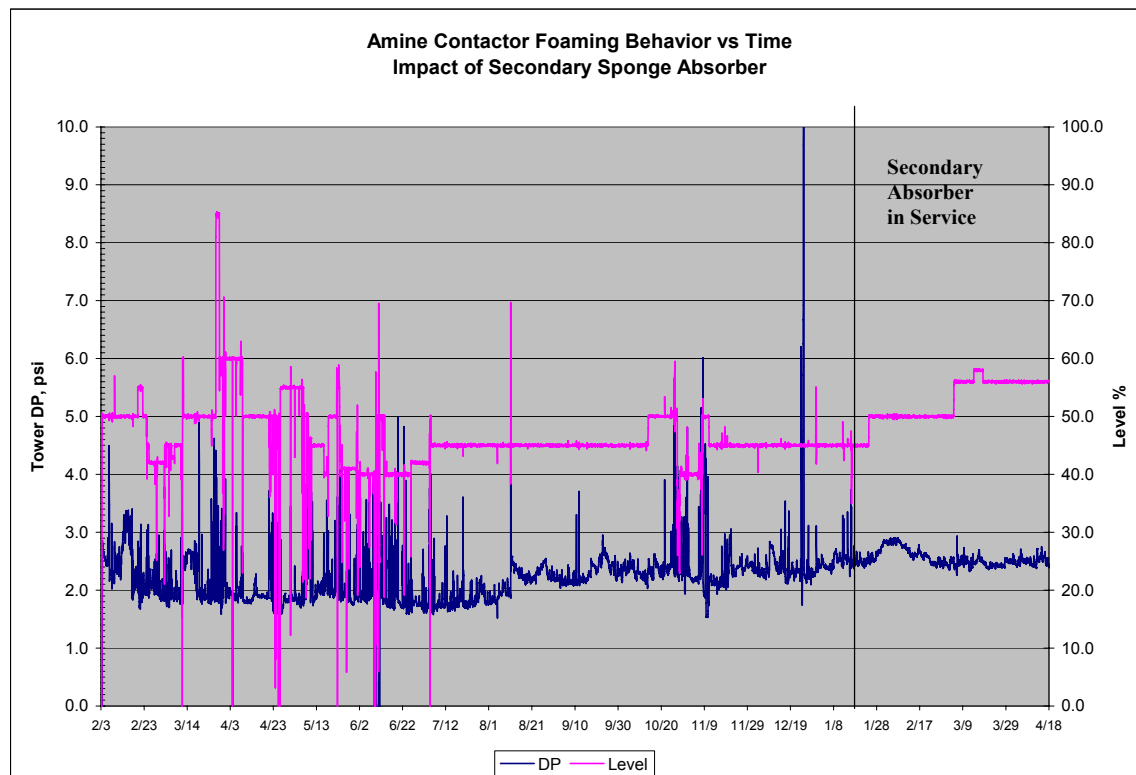
Even though we had learned, installed, trained on, and implemented lean amine/sour gas differential temperature control for a variety of gas treaters, heavy hydrocarbon in the gas phase still entered the amine and ended up causing foaming somewhere in the system. After finding out which hydrocarbons were causing the problems, we had to act to stop the hydrocarbon incursion by other means, because lean amine/sour gas differential temperature control just did not stop the oil.

Case One

After simulation and consultation with internal fractionation experts, it was decided to commission an unused tower to function as a secondary absorber using LCO from the FCC. At first the refinery, with high hopes, put the tower in service without the sponge oil hoping the large vessel volume would assist in removing C_6^+ hydrocarbons. The foaming continued with the same frequency. The plant then completed installation of the remaining piping, pumps, and controls to provide sponge oil to the absorber.

In Figure C1-E we can clearly see the effect of completely removing the gasoline and heavier hydrocarbons from the inlet gas using the sponge oil tower. The foaming stopped dead immediately and did not return.

Figure C1-E: Secondary Sponge Absorber Eliminated Foaming



Cases Two and Three

The data from Case Three shows that we cannot apply the learning from Case One directly to Case Two. Treating hydrogen with trace heavy oils in the gas oil boiling range and the high partial pressure of hydrogen acting as a “stripping gas” on a typical diesel boiling range sponge oil directs us away from the sponge oil concept.

In Case Two, the key to prevent foaming became a two-pronged effort. The first prong was to insure the filter system captured as much of the dissolved oil as possible. This meant going from long rinse/back flush cycles on the oil filter media due to sour water generation concerns to aggressive rinse/back flush cycles. Carbon filtration would have required extensive filter outages to replace the carbon; the regenerable oil removal system gave the plant a sufficiently fast way to renew the surface of the oil removal media. The second prong was limiting anti-foam additions. Since anti-foam overdose caused these absorbers to become inoperable at any rates, the plant had to adopt stringent anti-foam permissives to prevent recurrence.

In Case Three some extension between foaming episodes was observed when the plant increased reflux purging on the regenerator. Unfortunately this could only be done as a batch process and not continuously, so even though the foaming was occurring occasionally, the frequency was significantly reduced.

Case Four

When the plant increased sponge oil circulation rate, some relief in the regenerator foaming frequency was observed. Apparently, the additional cooling of the FCC contactor feed gas was somewhat effective in reducing the oil incursion rate.

Conclusions

“MYTH” Busted! Lean Amine/Sour Gas differential temperature control for gas treaters in amine service does not prevent the accumulation of foam causing, higher boiling point hydrocarbon contamination in the amine solution. Lean Amine/Sour Gas differential temperature control will only prevent condensation from occurring. *It's still good to do; it just isn't perfect!*

In using Lean Amine/Sour Gas differential temperature control, don't overdo it. There may be a hydrocarbon solubility trap waiting.

C₆⁺ hydrocarbons need to be effectively removed from the gases upstream of an amine absorber to prevent water soluble components of the gas stream from being dissolved in the amine. Aromatic compounds are usually the worst components because they have higher solubility than do other hydrocarbon types. Sponge oil towers and silica gel beds³ have been used effectively depending on the amount of hydrocarbon needs to be removed.

Try to collect real data from real plants on the C₆⁺ content of the gases to be treated and include real compound data in simulations to understand their impact. Don't totally rely on pseudo-components or use guessed components. Simulate to see which of the components of the gas will condense-hydrocarbon, water, or mixtures of both.

Be careful when mixing towers with trays and towers with packing in gas absorption or regeneration the same amine system. Heavy liquid hydrocarbons will tend to accumulate at the vapor/liquid interface on the trays due the mass transfer surface generation mechanism (bubbling) of trays. Packed towers will allow contaminants to pass to the tower bottoms due to the mass transfer taking place on the film covering a fixed area. Hydrocarbons that slip through the packed tower can migrate to the trayed tower and get trapped.

Purging reflux from a regenerator not only prevents ammonia buildup and subsequent acid gas accumulation and corrosion, but it also removes hydrocarbon that becomes trapped by steam stripping and condensation in the regenerator. This will help prevent regenerator foaming.

Good analytical chemists are a blessing-an acknowledgement for the work Jim Thompsen has done in helping discover the compounds that make our life in amine treating interesting!

References

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