Processing High-Ammonia Acid Gas in a SRU's

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Despite 60 years of progress in ammonia destruction technology, confusion, misconceptions and controversy persist. This paper addresses key issues in an effort to eliminate common operating problems and increase destruction capacity.

Ammonia-bearing acid gases have successfully been destroyed in SRUs for over 30 years but most of the experience has been with ammonia contents of 25% or less. This paper looks at what happens when the ammonia content is higher, as it is in acid gases originating from the processing of phenolic and non-phenolic heavy crudes. These increase the temperatures in the reaction furnace, reducing sulphur recovery and increasing the sensitivity of burner air control.

The economics of a two-stage sour water stripper with a smaller SRU for processing the likely range of ammonia concentrations are discussed among other options. Also discussed is whether and when it is economic to have one of the sour water strippers in a one-stage configuration for handling acid gases from phenolic crude processing and another in a two-stage configuration for non-phenolic acid gases to enable the SRU to accommodate the whole range of ammonia contents.

With the sulphur content of crude oil and natural gas on the increase and with the ever tightening restrictions on sulphur content in fuels, the refiners and gas processors will require additional sulphur recovery capacity. At the same time, environmental regulatory agencies of many countries continue to promulgate more stringent standards for sulphur emissions from oil, gas and chemical processing facilities. It is necessary to develop and implement reliable and cost-effective technologies to cope with the changing requirements.

Crude oil from several areas, including Brazil and Venezuela, contains high levels of nitrogen, which results in a very high level of ammonia in the sulphur recovery unit feed. We have seen ammonia contents of the overall acid gas as high as 55% by volume recently.

In some refineries, a purified ammonia stream is directly routed to a forced-draft incinerator, for instance, the NOxidizer™ system from John Zink. The NOxidizer™ system employs a two-zone furnace with a quench section between the two furnaces, the first zone operating under reducing conditions and the second zone operating under oxidizing conditions at a relatively low temperature to limit NOx formation.

It is well known that using oxygen enrichment will help with ammonia destruction in the Claus process. Oxygen enrichment raises the reaction furnace temperature which ensures complete destruction of heavy hydrocarbons and ammonia; reduces formation of COS and CS_2, and shortens gas residence time requirements for

contaminants destruction. However, for a very high level of ammonia the Claus unit has to operate on oxygen all the time and special provision must be made for preventing any plugging throughout the unit.

This paper will discuss the options for handling high ammonia feeds in SRUs.

The options include:

- two-stage sour water stripper design;
- oxygen-enrichment;
- ammonia destruction in incineration such as the NOxidizer™; Duiker
- Ammonia dissociater

A brief history

The first US refinery Claus plant was started up in Long Beach, California, in 1949 by the Hancock Chemical Co., a subsidiary of Hancock Oil, to process hydrogen sulphide (H_2S) from the adjacent Richfield Oil Co. and rich DEA from The Texas Co. two miles away. Understandably, the new technology was far from being well understood, and design features – direct contact sulphur condensers, for example – which looked good on paper proved dubious in practice. One early problem was chronic fouling of the amine acid gas headers with ammonium salts. In 1952, a small 10-tray foul water stripper was installed to re-strip H_2S and ammonia (NH_3) from the DEA regenerator overhead condensate (Fig. 1).
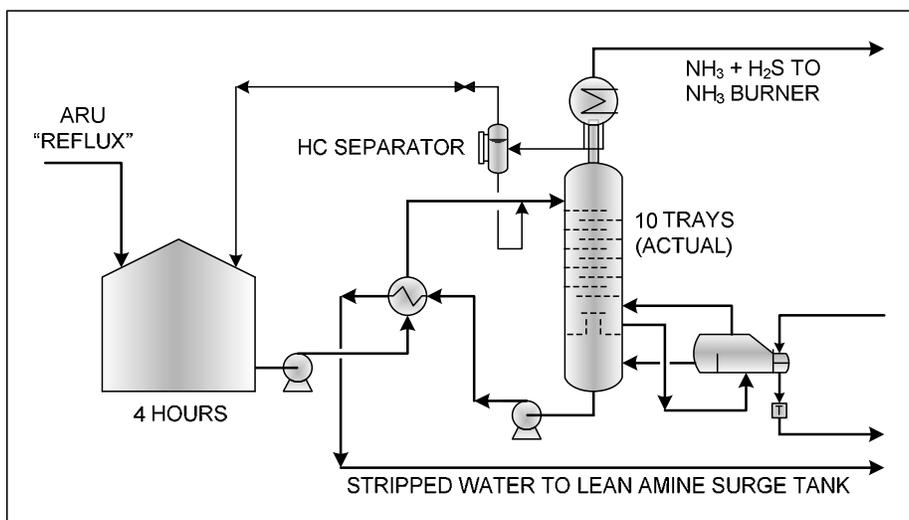

Fig. 1: DEA regenerator "reflux" stripper

Conventional thinking assumed that combustion of ammonia with excess air was necessary for complete destruction. The ammonia gas was thus fully oxidized in a dedicated burner in parallel with the Claus main amine acid gas burner. As can be expected, Claus condenser corrosion and catalyst deactivation were chronic problems.

Around 1970, it was recognized that undue Claus unit corrosion was the result of sulphur trioxide (SO_3), which condenses as sulphuric acid in the sulphur condensers. Accordingly, the concept of oxidizing ammonia in a reducing atmosphere in the first zone of a two-zone (or "front/side split") reaction furnace was developed, while splitting the ammonia-free amine acid gas between the two zones to ensure sufficient excess H_2S in zone 1 to minimize SO_3 according to Fig. 2, without unduly lowering the temperature considered necessary for complete ammonia destruction (Fig. 3) [1].
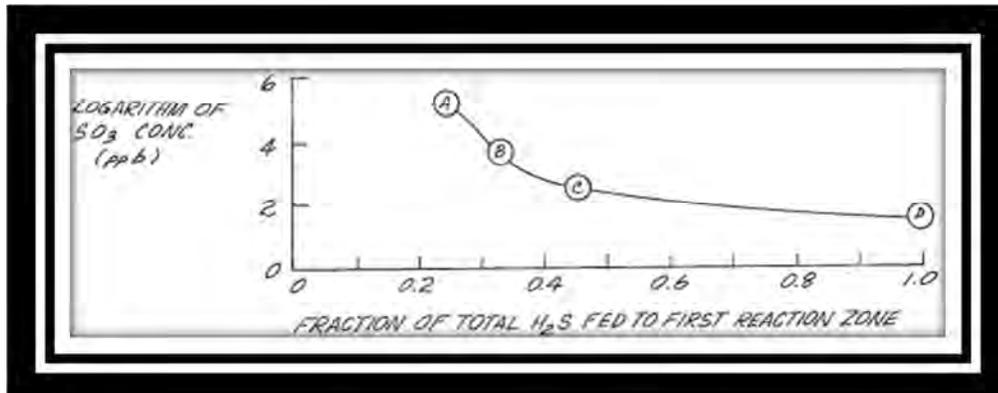
Fig. 2:   SO$_3$ as a function of amine acid gas split
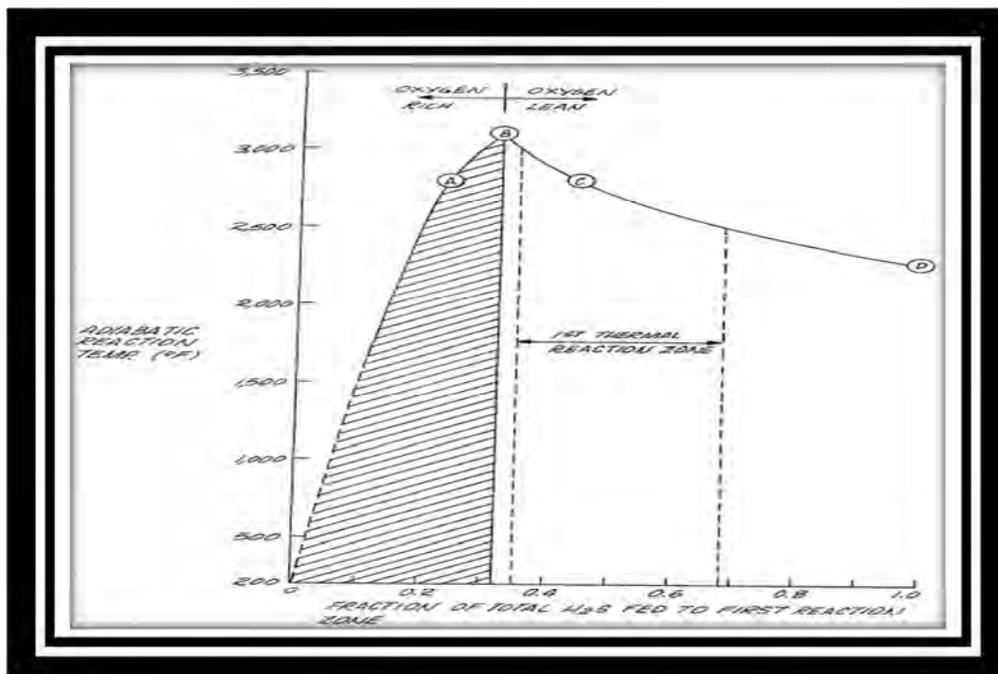
Fig. 3   Impact of amine acid split on Zone 1 temperature

It was further further described a reaction furnace divided into two zones by a refractory choke ring (Figs 4, 5) which (1) enhanced zone 1 mixing by reducing short-circuiting, (2) enhanced zone 2 mixing by virtue of increased turbulence resulting from the temporary velocity increase, (3) reflected radiation back for increased zone 1 temperatures, and (4) partially shielded the waste heat boiler inlet tube sheet from radiant heat.
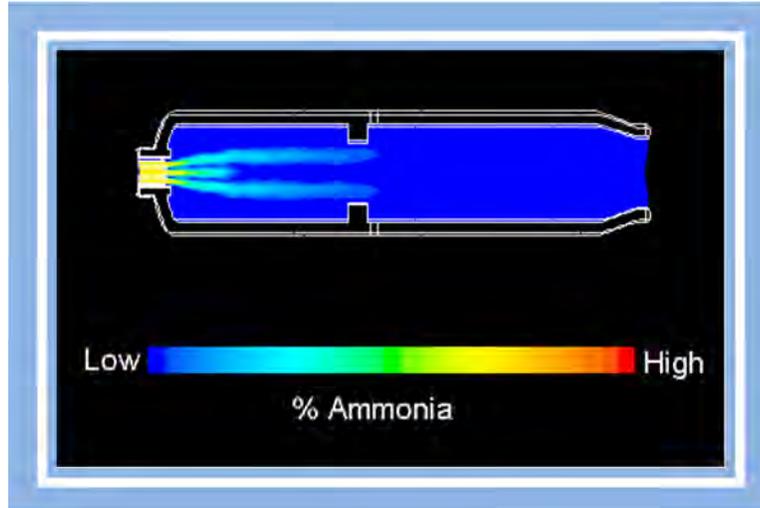
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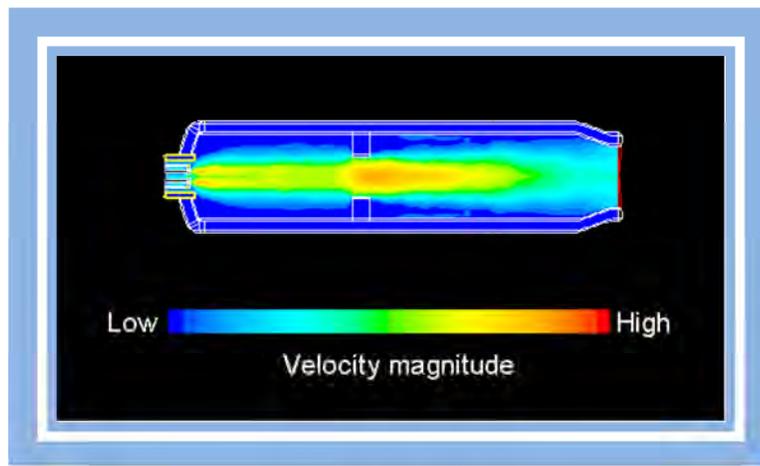**Fig. 4:   Choke ring effect on NH$_3$ distribution**

Fig. 5:   Choke ring effect on velocity

We continue to employ the same basic two-zone reaction furnace concept that has proved so efficient and robust, albeit typically with high-intensity burners which have become the industry standard.

The alternative to a choke ring in competing two-zone furnace designs is a checkerwall. In general, we consider checker walls more prone to failure and inferior in terms of zone 2 mixing. Choke rings also have the advantage of facilitating zone 2 access for maintenance.

Beginning in 1969, Comprimo (now Jacobs), usually in licensing partnership with Goar, Allison & Associates, offered a system for ammonia destruction in a single furnace chamber, contingent on suitable temperature, utilizing the LD Duiker high-intensity burner.

Other engineering companies followed suit with either single-chamber or two-zone designs. While most proponents of the two-zone scheme adopted the RATE concept of a zone 1 reducing environment, a notable minority clung to recommending ammonia destruction in an oxidizing environment as late as the 1990s, possibly

even to the present. That is decidedly ill-advised in our experience, not only because of the SO_3 problem but also because Claus reaction stoichiometry limits the amount of sour water H_2S to one third of the total for ammonia destruction in an oxidizing environment. With reducing conditions in zone 1, by comparison, sour water H_2S can account for at least 50% of the total.

For many years, ammonia destruction mechanisms were generally presumed to be a combination of simple oxidation – equation (1).

$$2NH_3 + 1.5\ O_2\ \rightarrow\ N_2 + 3H_2O \qquad (1)$$

and thermal dissociation with subsequent oxidation of hydrogen – equations (2) and (3).

$$2NH_3\ \rightarrow\ N_2 + 3H_2 \quad (2)$$

$$2H_2 + O_2\ \rightarrow\ 2H_2O \quad (3)$$
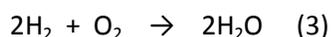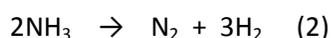
In an apparent effort to rationalize the virtually complete ammonia destruction that could be achieved, it was often repeated in the literature – at least until the mid 1990s – that ammonia and hydrocarbons were selectively oxidized in the presence of H_2S because they burned so much faster. ASRL disagreed, in part because the H_2S auto-ignition temperature is so much lower, and concluded from testing that relative oxidation rates are actually $H_2S > HCs > NH_3$ [2].

ASRL further concludes that thermal dissociation of ammonia is strongly inhibited by H_2S and water – both abundant in the Claus furnace. It follows, and is now generally accepted, that the predominant route of ammonia oxidation is by reaction with SO_2, ostensibly as in equation (4) [3].

$$2NH_3 + SO_2\ \rightarrow\ N_2 + H_2S + 2H_2O \quad (4)$$
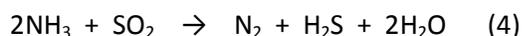
Of course, nominal stoichiometric oxygen demand is still defined by the above reaction of ammonia with oxygen.

Single chamber vs two-zone

Furnace temperatures of 2,250-2,350°F (1,232-1,288°C) are generally considered necessary for complete destruction of ammonia. For air-only plants, the obvious advantage of a two-zone furnace is generally higher zone 1 temperatures – particularly important during turn-down. Single-chamber designs more commonly employ acid gas and/or combustion air preheat (usually with indirect HP steam or "hot oil") to boost furnace temperatures. Beyond that, the relative merits are a subject of debate.

Single-chamber proponents claim better potential oxidation of organics in amine acid gas because the entire stream is processed in the main burner. On the other hand, sour water is arguably more likely to contain emulsified organics – particularly where there is a coker – which are more likely to be destroyed in zone 1 of a two-zone furnace, where the approach to oxidizing conditions is much greater.

A seemingly ignored aspect of two-zone furnaces is the increased SO_2/NH_3 ratio. The consensus seems to be that the ammonia oxidation reaction is not equilibrium-limited, and the same temperatures are therefore required in a two-zone furnace as in a single chamber. At the least, it is arguably reasonable to conclude that a two-zone furnace tends to compensate for lower burner intensity or residence. Still, it seems odd that this key consideration has not been rigorously explored. It also suggests that generalizations about ammonia destruction should maintain a distinction between single-chamber and two-zone systems.

Single-chamber proponents claim simpler process control. While it is true that automatic adjustment of the amine acid gas split in a two-zone furnace can be successfully achieved, the potential for inadvertent oxidizing conditions is a downside. Unfortunately, the control operator is less likely to appreciate the long-term adverse impact of oxidizing conditions than the process engineer or maintenance supervisor.

We discourage operator emphasis on maximizing the zone 1 temperature because it tends to encourage oxidizing conditions. It is arguably better to calculate the amine acid gas split based on measured flow rates and typical H_2S concentrations. Our practice is routing 33-66% of total H_2S to zone 1, and preferably 33-50%. From a practical standpoint, however, 33% is essentially oxidizing, and the zone 1 share may actually be less than 33% on account of meter error or compositional variations. To provide a safety margin, a target of 45% (based on a range of 40-50%) of total H_2S to zone 1 is typically used, resulting in the following formula with reference to Fig. 6.

$$A*X + C*Y \quad = \quad 0.45*(A*X + B*Y) \qquad (5)$$

$$C \quad = \quad 0.45*B - 0.55*A*X/Y \qquad (6)$$

Such calculation is only required if necessary to increase the zone 1 temperature in cases where SWS gas accounts for a significant share of total H_2S. Otherwise a simple 50/50 split of amine acid gas will suffice. Figure 7 compares predicted flame temperatures for (1) 45% of total H_2S routed to zone 1 and (2) a simple 50/50 amine acid gas split for SWS gas rates ranging from 10% to 50% of total H_2S.
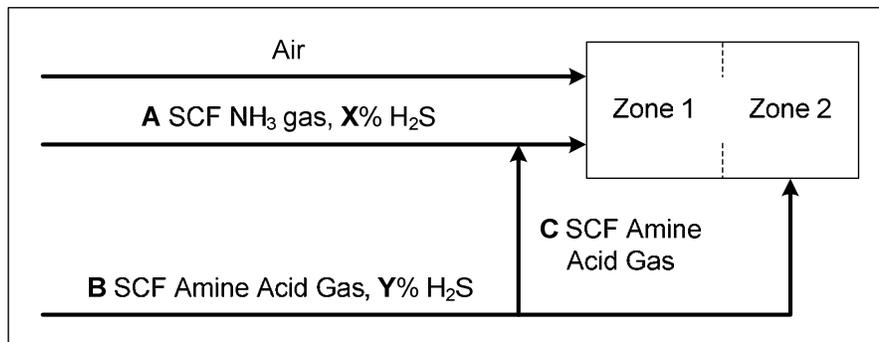
Fig. 6   Calculation of amine acid gas split
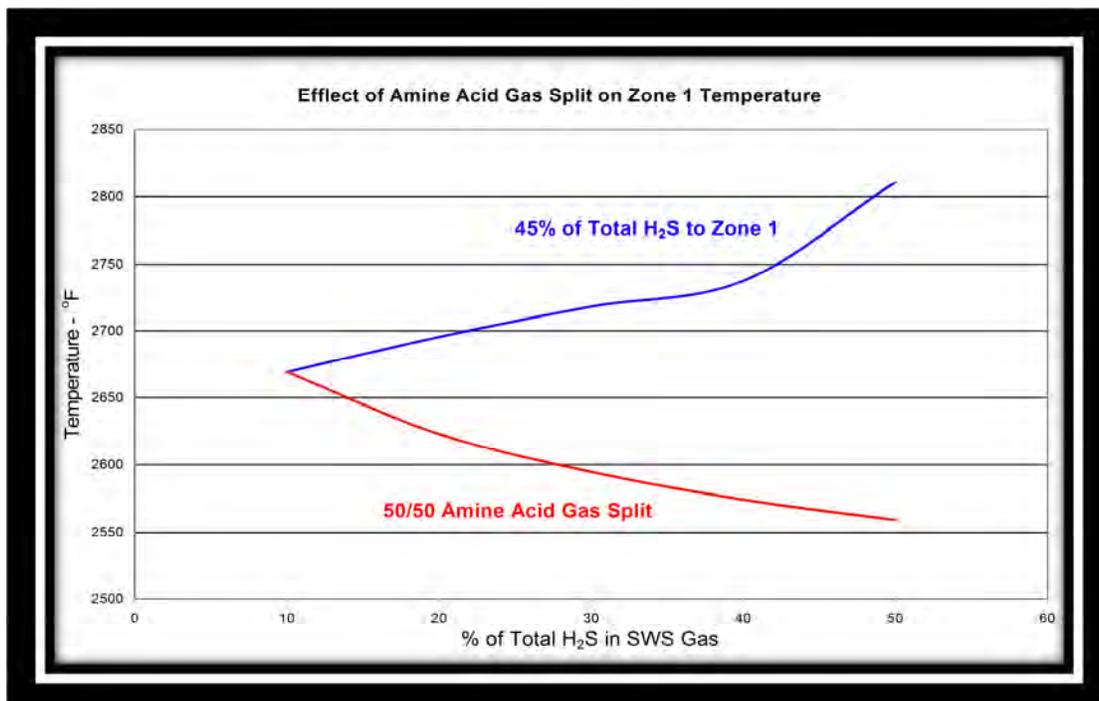
Fig. 7: **Temperature impact of optimized vs. 50/50 amine acid gas split**

Of course, there are limits to the accuracy of simulated temperatures, given the complexity of Claus furnace chemistry. One source suspects that pyrometer readings generally tend to be 50-150°C (90-270°F) lower than predicted [2]. Another source suspects that higher temperatures predicted at higher NH_3 concentrations tend to be offset by increased endothermic dissociation [4].

Based on the nominal stream compositions in Table 1, Table 2 presents a simplified sensitivity analysis of individual +/– 20% deviations in flow meter accuracy or H_2S concentrations. For the arbitrary case of equimolar ARU and SWS gas streams, actual zone 1 H_2S will range from 40-52% of total H_2S, equating to errors ranging from –11% to +16% of the 45% target (F). In short, the deviation is acceptable. But multiple simultaneous errors do complicate the analysis.

Table 1 Hypothetical Compositions		
Component	Mole %	
	ARU gas	SWS gas
H_2S	80	35
CO_2	13	
NH_3		35
H_2O	7	30
	100	100

| A	B	C	X	Y	F
1.00	1.00	0.21	0.35	0.80	0.45
0.80	1.00	0.21	0.35	0.80	**0.41**
1.20	1.00	0.21	0.35	0.80	**0.48**
1.00	**0.80**	0.21	0.35	0.80	**0.52**
1.00	**1.20**	0.21	0.35	0.80	**0.40**
1.00	1.00	0.21	**0.28**	0.80	**0.41**
1.00	1.00	0.21	**0.42**	0.80	**0.48**
1.00	1.00	0.21	0.35	**0.64**	**0.49**
1.00	1.00	0.21	0.35	**0.96**	**0.42**

The chemistry of unwanted byproducts

The fundamental issue is how to destroy ammonia in the Claus reaction furnace without causing downstream acid corrosion or deposition of ammonium salts.

Barring localized metal skin temperatures at or below the sulphur freezing point, acid corrosion is obviously the result of SO_3, which essentially exists as sulphuric acid vapour downstream of the waste heat boiler. Most of that acid vapour which does not either sulphate the catalyst or combine with residual ammonia to deposit salts condenses in the sulphur condensers, particularly the last. In severe cases liquid acid floats on the sulphur until trapped within the dip leg.

As stated, our rationale for destroying ammonia in a reducing atmosphere was to avoid by-product oxidation of SO_2 to SO_3 (Fig. 2). However, he may have been right for the wrong reason. Sulphur trioxide is considered unstable above 800°C (1,472°F), and dispute therefore persists over the possibility of SO_3 surviving the highly reducing conditions of zone 2 (assuming decent mixing and residence) [5].

NOx offers an alternative route to SO. Nitrous oxide (NO) is a potential by-product of ammonia oxidation by molecular oxygen:

$2NH_3 + 2.5O_2 \rightarrow 2NO + 3H_2O$
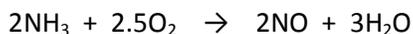
If all free oxygen were initially consumed by H_2S, NO would not form. Conversely, if the amine acid gas split results in an excess of air in zone 1, some NO will most likely be made.

In the real world, some ammonia will undoubtedly beat H_2S to the oxygen in any case. Under reducing conditions, however, NO initially formed may still be reduced back to ammonia:

$NO + 2.5H2 \rightarrow NH_3 + H_2O$

$2NO + 5H_2S \rightarrow 2NH_3 + 2H_2O + 2.5S_2$
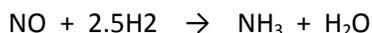
Under oxidizing conditions (local if not bulk), some NO will oxidize to nitrogen dioxide (NO_2):

$$NO + 0.5O_2 \quad \rightarrow \quad NO_2$$
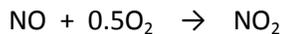
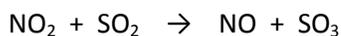Nitrogen dioxide in turn can oxidize SO_2 to SO_3, where the reaction is favoured at < 500°C (932°F):

$$NO_2 + SO_2 \quad \rightarrow \quad NO + SO_3$$

The implication, then, is that at least some NO_2 survives zone 2 to form SO_3 in the waste heat boiler and downstream. As an aside, it follows that it is particularly important to minimize NH_3 in the amine acid gas in plants with acid-gas fired auxiliary reheat burners, as substantial NOx will undoubtedly be made despite bulk excess H_2S.

Ironically, the above NO reduction reactions also mean that NOx formed in zone 1 will tend to revert to ammonia in zone 2. Sulphuric acid vapour and ammonia will invariably combine downstream to form ammonium sulphate. At best, gradual catalyst deactivation will result from plugged pores. At worst, salt deposition within exchanger tubes, catalyst beds and demisters will restrict process gas flow.

It is reasonable to expect higher ammonium salt deposition rates in plants with acid gas auxiliary reheat burners due to their propensity for generating SO_3. Contrary to a common belief, however, ASRL concludes that ammonium sulphate and thiosulphate can deposit in the absence of SO_3 [5]. Nevertheless, it is arguably reasonable to expect that ammonium salt deposition is slower in the absence of SO_3.

Residual ammonia levels of <150 ppmv are generally considered good practice [6]. Minimum detection levels are estimated to be 25-30 ppmv [5, 6].

Often, deposits assumed to be ammonium salts are, in fact, iron sulphate corrosion products. Whereas both salts should be water-soluble, combined sulphur often renders them otherwise.

Clarification of terminology

Once it is determined how to destroy conventional quantities of ammonia without downstream acid corrosion or salt deposition, often the next issue is how much ammonia can be processed before those problems become unavoidable. Unfortunately, anecdotal evidence and rules-of-thumb are often clouded by inconsistent and undefined interpretations of the "% NH_3" in Claus feed.

Consider, for example, the nominal stream compositions in Table 1. Assuming a combined Claus feed where SWS H_2S accounts for 50% of the total, the molar breakdown will be as shown in Table 3.

Table 3 Relative Moles in Claus Feed						
Stream	**Moles**				**Total**	
	H_2S	**CO_2**	**NH_3**	**H_2O**	**Dry**	**Wet**
ARU gas	1.00	0.16		0.09	1.16	1.25
SWS gas	1.00		1.00	0.86	2.00	2.86
Total	2.00	0.16	1.00	0.95	3.16	4.11

There are at least three possible definitions of "% NH_3" in the Claus feed. (Table 4) Definition 1 (molar wet basis) is presumed to be the most prevalent, but uncertainty remains. Practical upper limits of "25-30% NH_3" are generally cited, but often subject to interpretation [7].

Table 4 Possible Expressions of Relative Molar NH_3	
Definition	**Mol %**
1) NH_3 / total (wet basis)	24
2) NH_3 / total (dry basis)	32
3) NH_3 / (NH_3 + H_2S)	33

In addition, "% NH_3" may occasionally be expressed as the effective concentration to zone 1 of a two-zone furnace. This amounts to the proverbial apples-and-oranges and is potentially misleading, as the relative air in zone 1 is far different for a single chamber than with a two-zone furnace with an amine acid gas split.

Expanding ammonia destruction capacity

In the last 20 years there has been a general trend toward higher NH_3/H_2S ratios due variously to higher-nitrogen crudes and increased hydrotreating activity. In some cases, relative ammonia quantities exceed what can be conventionally processed in the Claus unit, as limited by the ability to avoid excessive NOx. Even at the comfortable limit of 25% NH_3 in the total Claus feed (molar wet basis), SRU sulphur production capacity is reduced by 50%. The typical solution is to separate H_2S and ammonia in a two-stage sour water stripper and incinerate the ammonia external to the Claus unit.

The most prevalent example of two-stage sour water stripping is the two-stage process, originally developed in the 1960s to process relatively concentrated sour wash water from hydrocrackers and hydrotreaters (Fig. 8). Hydrogen sulphide is selectively stripped at ~ 125 psig in the first tower, with subsequent recovery of ammonia at normal single-stage SWS pressures in the second. The stripping steam requirement is nominally twice that of a conventional SWS. Process enhancements can yield marketable anhydrous ammonia, although consistently meeting product specification has reportedly been a problem in many cases. More often, the ammonia is incinerated.
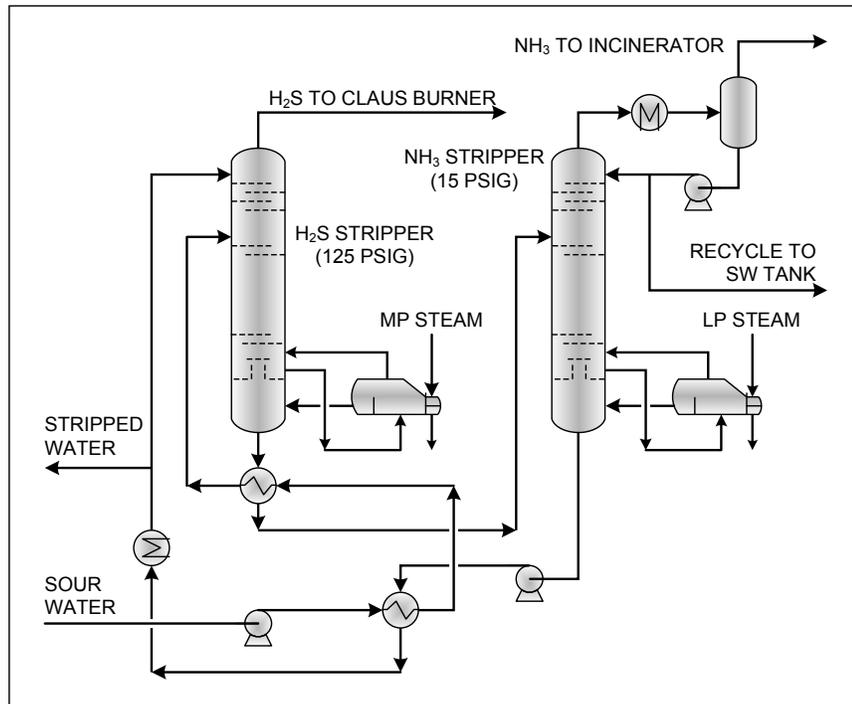
Fig. 8: Two-stage SWS design

In general, the days of simple ammonia incineration are gone because of the high NOx emissions. More sophisticated systems are required, such as the John Zink NOxidizer™, and Duiker low NOx incinerator. In one version of the Zink NOxidizer™ or similar like Duiker (Fig. 9), supplemental fuel and ammonia are initially oxidized under reducing conditions. The stream is partially quenched with flue gas or Claus tail gas, followed by addition of secondary air for subsequent oxidation of residual carbon monoxide and hydrogen in the absence of ammonia, which would be conducive to NOx formation. Waste heat steam is usually recovered.
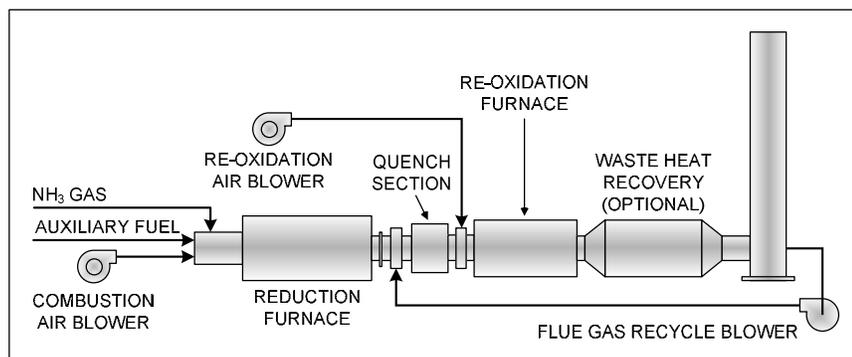
Fig. 9:   John Zink NOxidizer™

Oxygen Enrichment

Oxygen enrichment technologies in Claus sulphur recovery plants have gradually gained worldwide acceptance as a cost-effective method to expand SRU capacity, though there are other reasons for using oxygen enrichment

besides merely increasing capacity. In general, commercially available technologies offer three levels of oxygen enrichment: low-level, medium-level, and high-level providing additional capacity of about 25%, 75% and 150% respectively.

Effects of oxygen use on ammonia and hydrocarbon destruction

If not properly destroyed, hydrocarbons in the acid gas feed often cause carbon lay-down on the catalyst and generation of undesirably high concentrations of COS and CS_2. In addition, ammonia in the acid gas feed often causes deposition of complex ammonia/sulphur salts in cooler parts of the plant. These undesired phenomena can cause unscheduled plant shut-downs, reduce sulphur recovery and shorten catalyst life. Oxygen enrichment raises the reaction furnace temperature, which ensures complete destruction of heavy hydrocarbons and ammonia, reduces formation of COS and CS_2, and shortens the gas residence time needed to destroy contaminants.

In the case of lean acid gas feed contaminated with high levels of heavy hydrocarbons, oxygen enrichment provides an inexpensive and simple solution to circumvent a problem that is otherwise insoluble without recourse to costly processing technology.

Quantitatively, based on practical experience, the "SURE" burner has proved to be very effective in destroying ammonia and hydrocarbons in Claus plants. Outside their application in Claus plants, oxy-fuel burners are widely used in the metals and minerals, chemical and refining industries to burn a wide range of fuels, including gases, liquids, and pulverized solids. One of their most attractive features is their ability to burn heavy residual hydrocarbons cleanly.

Two major effects in using oxygen or oxygen-enriched air in place of air for combustion are higher temperatures and higher flame speeds. The degree of change depends on the degree of oxygen enrichment, but in the case of pure oxygen, temperatures may increase by 1,000°C and flame velocities by 10 times in round numbers. The combination of these two effects is to produce a hotter, shorter, more intense flame much better suited to the rapid destruction of combustible materials.

The destruction of individual feed components in a Claus unit cannot be considered in isolation, since there is considerable molecular interaction. Both hydrogen sulphide and ammonia dissociate quite readily and the higher the temperature, the higher the level of dissociation. The result is that when oxygen is used, the hydrogen level in the reaction furnace increases greatly over that achieved in air-based systems. Most of this hydrogen will subsequently recombine with sulphur in the waste heat boiler (WHB), including hydrogen produced from ammonia dissociation. So the ammonia must be burnt effectively, even if the mechanism of destruction is initially dissociation, in order to preserve the Claus stoichiometry downstream of the WHB.

It is possible to speculate that the hydrogen remaining in the gas after the WHB will be higher if the level in the reaction furnace before the boiler is higher. This must be true if the quench rate in the boiler remains constant. The effect may be small, however, and in the case of up rating with oxygen, where the WHB sees a higher load, a fall in quench rate may reduce it still further.

Heavy hydrocarbons such as BTX can be present in the feed to Claus units in certain cases; and their propensity for cracking thermally to produce carbon is well understood. Little, if any, published data is available on the effect of oxygen on BTX destruction in a Claus environment; however, it is known that oxygen is very efficient in burning these materials in other environments.

It is also possible to design the "SURE" Claus burner so that feed stream contaminants are preferentially destroyed, making use of the much higher reaction rates which are available even in plants operating with only moderate enrichment levels. Qualitatively, therefore, oxygen should have a beneficial effect on the destruction of heavy hydrocarbon contaminants such as BTX.

Three general parameters may be said to control the destruction of feed stream contaminants in Claus units: temperature, mixing and residence time. Temperature may be the most important parameter; and mixing is an essential parameter to ensure that all the contaminant molecules reach a high enough temperature and promote reaction where appropriate. A study of the Claus system using a kinetic CFD model indicates that residence time is important, particularly to sulphur forming reactions, but while oxygen is present, the system is dominated by hydrogen combustion. In such a system, molecules like ammonia compete poorly for oxygen and the initial step in its destruction is likely to be dissociation. The higher temperatures generated with oxygen use clearly favour this; and since the ammonia is often contained in a stream separate from the main Claus feed stream, it is possible to maximize the benefit through the burner design.

The key parameter considered when applying the Claus process is to maintain a stable flame at the burner. However, if the feed gas contains contaminants, a much higher flame temperature will be required to destroy the undesirable compounds in the furnace so that they do not cause operational difficulties downstream.

Others (cyanide, mercaptans)

These contaminants can be destroyed in the Claus furnace based on the same considerations as given for aromatic hydrocarbons above. They are detrimental to the wet oxidation and non-regenerative processes due to spent chemical disposal and odour problems.

It cannot be overemphasized that a well-designed burner and reaction furnace, which promote good mixing of the reactants, is essential for complete destruction of undesirable feed contaminants.

Economics

When a refinery is equipped with integrated SWS facilities for both phenolic and non-phenolic sour water, the incremental capital cost is primarily due to the addition of an extra stripper and reboiler and the incremental operating cost is due to the higher steam consumption. If the refinery SWS uses separate strippers for non-phenolic and phenolic sour water, it is often best to keep the phenolic sour water stripper as a single-stage stripper because a far greater proportion of the total ammonia in the sour water is associated with the non-phenolic sour water. Economically, it is best to provide an integrated SWS or, if separate SWS systems are in place or preferred for water reuse and management, to only use a two-stage stripper on the non-phenolic sour

water. With a two-stage stripper SWS, the ammonia from the sour water no longer goes to the Claus unit. This reduces considerably the mass throughput in the Claus plant and, consequently, the tail gas unit as well. That more than offsets the higher cost of the two-stage sour water stripper. Sulphur recovery is also increased by because the SWS stream to the Claus unit has been eliminated.

As shown in Table 7, a two-stage SWS looks attractive, achieving the 99.9% recovery target. It could be optimized according to what technology for ammonia destruction is being selected.

Table 7 Capital Cost Comparison		
	Single–stage SWS plus SRU	Two-stage SWS plus SRU
SWS (BASE)	50	80
SRU Claus 2-stage	82	55
BSR-TGU/ MDEA	68	45
Incinerator with heat recovery	20	24
Total cost	220	204

Conclusion

Ammonia destruction in a Claus tail gas treating unit is an alternative solution where the ammonia content of the feed to be processed exceeds 30-35% of the total Claus feed on a wet basis. Crude oil from several areas contains levels of nitrogen which result in ammonia levels of 50-60% of the total acid gas, well beyond the capability of conventional sulphur plants. The following criteria should be considered.

1. For ammonia burning in the Claus unit, the industry standard achieving stable combustion temperature is essential to prevent any plugging from unconverted ammonia.
2. A high-intensity burner has to be selected with possibility of CFD modelling to assure ammonia destruction.
3. Based on CFD modelling results, a two-zone reaction furnace with a choke ring is always recommended to ensure ammonia destruction, uniform gas distribution and adequate combustion temperature, and to prevent cold spot in the furnace, condensation and sulphuric acid formation and, ultimately, plugging in downstream equipment.
4. If oxygen is available it would be the most economic option which would not be requiring having a two-stage SWS.
5. The NOxidizer™ or Duiker or similar process requires a two-stage SWS operating perfectly so that the ammonia entering the unit is free of H_2S in order not to violate any emissions regulations. That is a difficult commitment.
6. Even though a two-stage SWS is more expensive than a single-stage SWS, because ammonia does not have to be processed in the sulphur recovery plant, the larger reduction in the cost of the smaller-sized Claus unit and BSR/TGU gives an overall cost reduction.
8. The economics of a single stripper with a larger SRU against a two-stage stripper and a smaller SRU depend on several factors such as flow, quantity of ammonia, SRU recovery and ammonia destination.

9. A two-stage SWS is more economical in larger units processing maximum ammonia contents. The overall recovery is also at a maximum.
10. For each project, the selection of the best option must be made on the basis of a careful study taking account of every applicable parameter in addition to cost considerations. These include factors such as capacity, ammonia content, availability of oxygen and utilities, environmental regulations, operability, constructability, and site constraints.

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