Sulfur Recovery Unit
Expansion Case Studies

Robin Street
Technical Director – Sulphur Technology
WorleyParsons Europe Ltd
EMC² Tower, Great West Road
Brentford, Middlesex, UK TW8 9AZ
Robin.Street@worleyparsons.com

Mahin Rameshni, P.E.
Chief Process Engineer
(Sulphur Technology and Gas Treating)
WorleyParsons
125 West Huntington Drive
Arcadia, CA, 91007, USA
mahin.rameshni@worleyparsons.com
<table>
<thead>
<tr>
<th>Section</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Introduction</td>
</tr>
<tr>
<td>2</td>
<td>Claus Unit Description</td>
</tr>
<tr>
<td></td>
<td>2.1 Simplified Process Description</td>
</tr>
<tr>
<td>3</td>
<td>Tail Gas Treatment Unit Description</td>
</tr>
<tr>
<td>4</td>
<td>Amine Unit Configurations</td>
</tr>
<tr>
<td></td>
<td>4.1 Revamping or Modifying Existing Equipment</td>
</tr>
<tr>
<td></td>
<td>4.2 Oxygen Enrichment</td>
</tr>
<tr>
<td>5</td>
<td>Example of Increasing Capacity by Equipment Revamp</td>
</tr>
<tr>
<td>6</td>
<td>Example of a Medium Level Oxygen Enrichment Project</td>
</tr>
<tr>
<td>7</td>
<td>Example of a High Level Oxygen Enrichment Project</td>
</tr>
<tr>
<td>8</td>
<td>Summary</td>
</tr>
<tr>
<td>9</td>
<td>References</td>
</tr>
</tbody>
</table>
Section 1 | Introduction

Sulphur removal facilities are located at the majority of oil and gas processing facilities throughout the world. The sulphur recovery unit does not make a profit for the operator but it is an essential processing step to allow the overall facility to operate as the discharge of sulphur compounds to the atmosphere is severely restricted by environmental regulations.

Oil and gas producers are attempting to maximise production at minimum cost. This often means debottlenecking existing upstream facilities and may result in extra sulphur recovery capacity being required.

Oil refiners are also increasingly being forced to comply with legislation reducing the levels of sulphur in products. Combine this with the ability or need to process sourer crude oils and many refiners find that their existing sulphur recovery units do not have sufficient capacity.

Further more, in many countries environmental legislation is demanding higher recoveries from sulphur recovery units.

This paper discusses some of the options for increasing sulphur unit capacity from Claus units and Tail Gas Treatment units on the basis that the existing recovery is adequate. The examples given relate plants that are designed for a sulphur recovery of 99.8 - 99.9%.
The basic Claus unit comprises a thermal stage and two or three catalytic stages. Typical sulphur recoveries efficiencies are in the range 95-98% depending upon the feed gas composition and plant configuration.

The basic chemical reactions occurring in a Claus process are represented by the following reactions:

\[
\begin{align*}
\text{(1)} & : \quad \text{H}_2\text{S} + \frac{1}{2}\text{O}_2 & \rightarrow & \text{SO}_2 + \text{H}_2\text{O} \\
\text{(2)} & : \quad 2\text{H}_2\text{S} + \text{O}_2 & \leftrightarrow & \frac{3}{x} \text{S}_x + 2\text{H}_2\text{O}
\end{align*}
\]

Some of the \( \text{H}_2\text{S} \) in the feed gas is thermally converted to \( \text{SO}_2 \) in the reaction furnace of the thermal stage according to reaction (1). The remaining \( \text{H}_2\text{S} \) is then reacted with the thermally produced \( \text{SO}_2 \) to form elemental sulphur in the thermal stage and the subsequent catalytic stages according to reaction (2). Claus reaction (2) is thermodynamically limited and has a relatively low equilibrium constant for reaction (2) over the catalytic operation region.

As the feed acid gas normally contains other compounds, which could include carbon dioxide, hydrocarbons, mercaptans and ammonia, the actual chemistry in the furnace is very complex. The latest analysis of this has been presented by Borsboom and Clark. (reference 1).

### 2/3 Stage Claus Process

![Diagram of 2/3 Stage Claus Process]
The hot combustion products from the furnace at 1000-1300°C enter the waste heat boiler and are partially cooled by generating steam. Any steam level from 3 to 45 bar g can be generated.

The combustion products are further cooled in the first sulphur condenser, usually by generating LP steam at 3 – 5 bar g. This cools the gas enough to condense the sulphur formed in the furnace, which is then separated from the gas and drained to a collection pit.

In order to avoid sulphur condensing in the downstream catalyst bed, the gas leaving the sulphur condenser must be heated before entering the reactor.

The heated stream enters the first reactor, containing a bed of sulphur conversion catalyst. About 70% of the remaining H₂S and SO₂ in the gas will react to form sulphur, which leaves the reactor with the gas as sulphur vapour.

The hot gas leaving the first reactor is cooled in the second sulphur condenser, where LP steam is again produced and the sulphur formed in the reactor is condensed.

A further one or two more heating, reaction, and condensing stages follow to react most of the remaining H₂S and SO₂.

The sulphur plant tail gas is routed either to a Tail Gas treatment Unit for further processing, or to a Thermal Oxidiser to incinerate all of the sulphur compounds in the tail gas to SO₂ before dispersing the effluent to the atmosphere.
Tail Gas Treatment units are designed to increase the overall sulphur recovery by processing the gas from the Claus unit final sulphur condenser. The process discussed in this paper is generally referred to as the “SCOT” process although there are many versions of this type of process. Parson’s version of the process is known as BSR/Amine which has been operating in commercial plants for over 25 years. The use of this process can increase the sulphur recovery to over 99.9%.

The process has two sections, firstly the hydrogenation section where all sulphur compounds are converted back to hydrogen sulphide according to the following equations.

$$S + H_2 \rightarrow H_2S \quad (3)$$

$$SO_2 + 3H_2 \rightarrow H_2S + 2H_2O \quad (4)$$

$$CS_2 + 2H_2O \rightarrow 2H_2S + CO_2 \quad (5)$$

$$COS + H_2O \rightarrow H_2S + CO_2 \quad (6)$$

**BSR Hydrogenation Section**

**Simplified Process Description**

- In the RGG, natural gas is burnt substoichiometrically to produce some reducing gas H₂ and CO. These supplement H₂ present in the Claus tail gas from cracking of H₂S in the Reaction Furnace. The RGG combustion products are
mixed the Claus tail gas to bring it to the correct temperature for hydrogenation reaction.

- In the Hydrogenation reactor, all sulphur compounds are converted to H₂S by reaction 3-6 above. The reactions are exothermic and heat is removed from the gas in the reaction cooler, which produces LP steam at 3 – 5 bar g.

- The gas is cooled further in a Direct Contact Condenser (or Quench Tower) by a circulating water stream down to a suitable temperature for amine treatment and sour water is condensed from the stream.

The second section of the process is the removal of H₂S from the gas by amine treatment. The gas contains more CO₂, produced by combustion of hydrocarbons in the acid gas feed and the RGG, than H₂S. In addition, the acid gas itself may contain CO₂. Hence, the amine used must be selective for H₂S over CO₂. MDEA is often used for this application although alternative amines are available.

**Amine Section**

**Simplified Process Description**

- Gas is contacted with lean amine solution in the absorber. The amine absorbs the H₂S and some of the CO₂. The treated gas is sent to the Thermal Oxidiser where residual H₂S is converted to SO₂ before discharge to atmosphere.

- The rich amine is sent to the regenerator after being heated in the Lean/Rich exchanger by the hot lean amine from the bottom of the regenerator.
Section 3

Tail Gas Unit Description

- In the regenerator, the acid gases are released from solution by heating the solution in the reboiler. The overhead from the regenerator is cooled and the condensate returned to the column (not shown in drawing). The cooled, water saturated, acid gas is recycled to the Claus Unit.

- The hot lean amine is cooled firstly by heating the rich solution and then in the lean amine cooler before entering the absorber.
Section 4  

Amine Unit Configurations

Two options for increasing plant will be studied with reference to revamps that have been implemented.

4.1 Revamping or Modifying Existing Equipment

The first option to be considered is whether a plant’s capacity can be increased by changing current operating modes or partially revamping or replacing equipment. These are some of the topics that can be investigated:

i. The capacity of a sulphur recovery unit is governed by the pressure available in the acid gas, determined by the operating pressure of the upstream amine regenerator. This pressure is normally in the range, 0.5 – 1.0 bar g

Increasing the operating pressure of the upstream amine unit absorber and reducing the pressure drop of control valves and lines to the sulphur unit will give an immediate capacity boost. For instance, a pressure increase from 0.6 to 0.9 bar g will give around a 20% increase in capacity.

However, it will also be necessary to increase the pressure available from the combustion air blowers. It may be possible to revamp these by replacing impellers and/or motors depending upon the type of blower. The depth of the sulphur seal legs will also have to be checked.

ii. Continuing with the upstream amine units; where the acid gas contains a lot of CO₂ consideration may be given to replacing primary amines, MEA or DEA, with a selective amine such as MDEA. Some of the CO₂ can then be rejected in the treated gas, if this is acceptable. The removal of an inert gas from the feed will allow an increase in unit capacity.

iii. Sulphur units are often designed for end of run conditions, this means that allowances are made in the design for deactivation of the catalyst and fouling of the catalyst and equipment. It may be possible to revamp the unit to take advantage of this inherent capacity increase that is available.

The acid gas burner may have been designed for a high pressure drop, 0.08 bar is not unusual. A burner with a lower pressure drop could be considered.

iv. If a plant has fired heaters for reheat and these foul the catalyst then replacement with indirect (steam) reheat will help.

v. If the tail gas treatment unit is designed for an end of run 94% recovery in the Claus unit and the start of run recovery is 96% then replacement of alumina catalyst with titania catalyst could prolong the catalyst activity.
vi. Tail Gas Treating units are usually designed for a lower recovery in the Claus unit than it can actually achieve. For instance, a typical two reactor Claus can give a sulphur recovery of 96.5 – 97% but the TGTU may be designed for 94% recovery. Advantage can be taken of this “extra” capacity for a revamp but this does remove some of the safety factor incorporated into the original design for process upsets and catalyst deactivation.

In practice, it takes a combination of some of the options suggested to give any substantial increase in capacity. Experience shows that this approach can be implemented for some plants as will be seen by the example given later.

4.2 Oxygen Enrichment

The second option available for plant expansion is oxygen enrichment, which has been widely accepted throughout the world as a means of achieving increased sulphur recovery unit capacity. Parson’s designed units are operating in the U.S.A., Europe and Japan.

The principle behind the use of oxygen enrichment is the replacement of some or all of the air needed for combustion by oxygen thus removing inert nitrogen from the system. This allows the volume of acid gas processed to be increased for the same unit pressure drop.

The following tables illustrate the reduction in flow that can be achieved by using oxygen. The tables show that at a total oxygen concentration of 37% (air + oxygen) the same quantity of acid gas can be processed with almost a 30% reduction in mass flow rate. Hence, the same plant could theoretically process over 40% more acid gas purely on a pressure drop basis.

Interestingly, the hydrogen production, from H₂S cracking, is increased due to the higher furnace temperature resulting in the total oxygen demand falling by 5%.
Table 1

<table>
<thead>
<tr>
<th>Component, kg mole/h</th>
<th>Acid Gas</th>
<th>Combustion Air</th>
<th>Combustion Products</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrogen Sulphide</td>
<td>133.20</td>
<td>21.70</td>
<td></td>
</tr>
<tr>
<td>Sulphur Dioxide</td>
<td></td>
<td>15.06</td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>12.21</td>
<td>3.07</td>
<td>130.56</td>
</tr>
<tr>
<td>Oxygen</td>
<td>76.58</td>
<td>76.58</td>
<td></td>
</tr>
<tr>
<td>Nitrogen</td>
<td>285.51</td>
<td>290.92</td>
<td></td>
</tr>
<tr>
<td>Ammonia</td>
<td>10.81</td>
<td>10.81</td>
<td></td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>10.88</td>
<td>0.12</td>
<td>12.18</td>
</tr>
<tr>
<td>Hydrogen</td>
<td></td>
<td>21.23</td>
<td></td>
</tr>
<tr>
<td>Carbon Monoxide</td>
<td></td>
<td>5.40</td>
<td></td>
</tr>
<tr>
<td>Carbonyl Sulphide</td>
<td></td>
<td>0.02</td>
<td></td>
</tr>
<tr>
<td>Hydrocarbons, C₃</td>
<td>2.20</td>
<td>2.20</td>
<td></td>
</tr>
<tr>
<td>Sulphur, S₂</td>
<td></td>
<td>48.10</td>
<td></td>
</tr>
<tr>
<td><strong>TOTAL</strong></td>
<td>169.30</td>
<td>365.28</td>
<td>545.17</td>
</tr>
<tr>
<td>Mass Flow Rate, kg/h</td>
<td>5519.4</td>
<td>10510.1</td>
<td>16029.5</td>
</tr>
<tr>
<td>Temperature, ºC</td>
<td>40</td>
<td>40</td>
<td></td>
</tr>
</tbody>
</table>
Table 2

<table>
<thead>
<tr>
<th>Component, kg mole/h</th>
<th>Acid Gas</th>
<th>Combustion Air</th>
<th>Oxygen</th>
<th>Combustion Products</th>
</tr>
</thead>
<tbody>
<tr>
<td>Hydrogen Sulphide</td>
<td>133.20</td>
<td></td>
<td>16.37</td>
<td></td>
</tr>
<tr>
<td>Sulphur Dioxide</td>
<td></td>
<td></td>
<td>17.28</td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td>12.21</td>
<td>1.32</td>
<td></td>
<td>120.28</td>
</tr>
<tr>
<td>Oxygen</td>
<td>32.89</td>
<td>39.80</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Nitrogen</td>
<td>122.64</td>
<td>0.20</td>
<td>128.24</td>
<td></td>
</tr>
<tr>
<td>Ammonia</td>
<td>10.81</td>
<td></td>
<td></td>
<td>8.42</td>
</tr>
<tr>
<td>Carbon Dioxide</td>
<td>10.88</td>
<td>0.05</td>
<td>35.08</td>
<td></td>
</tr>
<tr>
<td>Hydrogen</td>
<td></td>
<td></td>
<td></td>
<td>9.10</td>
</tr>
<tr>
<td>Carbon Monoxide</td>
<td></td>
<td></td>
<td></td>
<td>0.01</td>
</tr>
<tr>
<td>Carbonyl Sulphide</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrocarbons, C₃</td>
<td>2.20</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Sulphur, S₂</td>
<td></td>
<td></td>
<td></td>
<td>49.72</td>
</tr>
<tr>
<td><strong>TOTAL</strong></td>
<td>169.30</td>
<td>156.90</td>
<td>40.00</td>
<td>384.51</td>
</tr>
</tbody>
</table>

| Mass Flow Rate, kg/h | 5519.4 | 4514.4 | 1279.2 | 11313.0 |
| Temperature, ºC      | 40     | 40     | 40     | 1500    |

The limiting factor in the use of oxygen enrichment is the furnace temperature. Maximum refractory design temperatures are in the range 1750 - 1800ºC. Worley Parsons normally limit the maximum operating temperature to 1500 - 1550ºC.

Three levels of oxygen enrichment are normally considered.

**Low level oxygen enrichment**

Oxygen is mixed with the combustion air to attain an oxygen concentration of up to 28%. This is the limit of oxygen in air that does not require special materials. Capacity increases of about 20 – 25% over the original design capacity can be obtained via this technique.

**Medium level oxygen enrichment**

Oxygen is introduced into a proprietary burner, independently of the combustion air, to attain an oxygen concentration between 28 – 45%. Capacity increases up
to 75% over the original design capacity are possible limited only by the furnace temperature.

High level oxygen enrichment

Oxygen is introduced into a special burner, independently of the combustion air, to attain an oxygen concentration between 45 – 100%. Capacity increases up to 150% greater than the original design capacity are achievable. The acid gas cannot be burnt directly with the enriched air stream as the combustion temperature is too high. A number of technologies are commercially available to overcome this problem. WorleyParsons/BOC's Double Combustion “SURE” process is one these. The use of this technology is given in one the following examples.

Figure 1 shows the effect of oxygen concentration and capacity increase versus % oxygen for a typical refinery acid gas.

The Double Combustion process is designed to overcome the temperature limitations shown above. The process can be configured in two ways. Either a new combustion chamber/waste heat boiler can be installed upstream of the existing boiler or new furnace plus two pass waste heat boiler can be installed. These options are shown in figures 2 and 3.
Figure 2

Figure 3

Reaction Furnace #2

oxygen
A large oil and gas facility selected WorleyParsons to study the possibilities of increasing the capacity of their sulphur recovery unit to meet the demands of higher processing capacity that was being installed upstream. After an examination of all the options, it was concluded that the required capacity could be achieved by revamping the plant without the use of oxygen.

Two sulphur unit trains were revamped, both of which, comprised a two reactor Claus unit followed by a SCOT tail gas treatment unit. The two trains were designed to operate in parallel at 70% capacity equivalent to 681 tpd sulphur production per train at 99.9% sulphur recovery. Hence, the design capacity of each train is calculated to be 973 tpd.

Attempts to operate above the normal capacity had proved problematic. Some of the major problems encountered were waste heat boiler tube failure, excessive steam carryover from the waste heat boiler and sulphur condenser accompanied with severe vibration, loading of the Claus reactor beds with sulphur and vibration of the reducing gas generator. Also, the H₂S content of the acid gas was 76% compared to the design basis of 81%. Consequently, the operating capacity of the trains was limited to around 750 tpd.

The project awarded to WorleyParsons was to increase the capacity of each train to 1300 tpd whilst improving the reliability of the unit. This was a 33% increase above the original design capacity and 73% above the operational maximum.

The following is a list of the major changes made to the plant:

- A hydraulic check of unit showed that much of the equipment and lines had been generously sized. However, to enable the unit to process the extra gas it was necessary to reduce the pressure drop where ever possible. Several measures were taken to accomplish this. These comprised replacing the burner on the reaction furnace, replacing the trays in the quench tower with random packing, replacing the trays the tail gas unit absorber with structured packing.

- The combustion air blowers could not provide sufficient flow at the correct pressure for the revamped plant. Analysis of the blower curves showed that if two blowers were operated at 50% of the flow rate each then the output pressure could be obtained. A new control scheme was installed to enable safe operation of the two blowers.

- A new single pass waste heat boiler was installed to WorleyParsons design. The new boiler eliminated film boiling which was the cause of tube sheet failures. The original boiler operated a hot gas bypass system of reheat. This
Section 5

Example of Increasing Capacity by Equipment Revamp

was eliminated in the new design and a new fired heater installed for the first Claus reactor reheat. The new boiler design also included a new large steam drum to overcome the water carry over and vibration problems of the original design.

- The steam drums on the sulphur condensers and the tail gas unit reactor effluent cooler were modified, installing new risers and down comers. This stopped the vibration occurring in these vessels.

- A new steam reheater was installed for the second Claus reactor reheater as the original was not large enough to handle the new duty. The new reheater was also designed to operate with a lower pressure drop.

- The final sulphur condenser was modified to allow the steam side to be operated at 1 bar g. This involved installing an air cooler to condense the steam produced with the condensate then being recycled to the condenser in a closed loop operation. This enabled the process gas outlet temperature to be reduced from around 160°C to 130°C thus condensing out more of the sulphur with a resultant increase in recovery in the Claus unit of 0.6%. In this way the load on the tail gas treatment unit was also reduced.

- The lower gas temperature from the final sulphur condenser increased the duty of the reducing gas generator; this unit was due for replacement as the original item was suffering from severe vibration problems.

- Extra catalyst was installed in the Claus and Tail gas hydrogenation reactors.

- The air coolers on the quench tower were already having problems keeping the circulating water temperature down in the summer. Extra water circulation and cooling were determined to be necessary for the revamp. A new circulating water pump was installed so that two pumps could be operated in parallel with one spare. Extra air cooling was also installed in parallel with the existing coolers.

- As already stated the absorber trays were replaced by packing as otherwise flooding of the columns would have occurred. However, the major change that enabled the amine section of the tail gas unit to operate at the higher capacity was the replacement of the MDEA solvent with Flexsorb.

- Flexsorb is a hindered amine developed by ExxonMobil for selective removal of H₂S. The use of Flexsorb allowed the amine circulation rate for the revamp to be around 80% of the original circulation rate whilst processing 33% more tail gas. This meant that no modifications were required on the regenerator and overhead system.
The lean/rich exchanger is a plate and frame exchanger. New plates were installed as Flexsorb needs special gasket material and replacement of the plates was deemed to be more economic than attempting replace the gaskets on the existing plates.

A water wash section was installed at the top of the amine absorber to minimise solvent losses.

ExxonMobil have appointed WorleyParsons as sub-licensors of processes using Flexsorb. Further details on the properties and use of Flexsorb can be found in a paper by Fedich, McCaffrey and Stanley (reference 2).
An Italian refinery needed to increase the capacity of one of their sulphur recovery units by 60% from 161 to 257.5 tpd. The unit was again a two reactor Claus with a SCOT tail gas treatment unit.

With an oxygen enrichment revamp the level of oxygen used is tailored to ensure that the unit pressure drop does not increase. For this project the total oxygen concentration was 33% with a consumption of 3 t/h. Oxygen at 93% purity was supplied by a dedicated VSA unit.

The changes that were carried out to the unit were as follows;

- New larger knock out drum for acid gas and sour water gas were installed but the existing lines were found to be large enough to handle to be increased flows.
- New BOC oxygen burner for the reaction furnace suitable for using oxygen and air.
- New oxygen supply system.
- Modifications to the control scheme to incorporate the new oxygen stream and replacement of the flow control valves for gas and sour water stripper gas flows.
- The reaction furnace refractory was replaced with a higher grade material suitable for the higher operating temperature. Modifications were also made to the furnace, to give a two zone configuration, for sour water stripper gas burning during air only operation.
- The steam line from the first sulphur condenser was increased from 4” to 6”.
- The trays in the Quench Column were replaced with structured packing.
- The Circulating Cooling water pump capacities increased by installing the maximum sizes impellors and new motor drives.
- An additional air cooler was supplied for the circulating water circuit.
- The solvent in TGTU amine section was changed to Ucarsol 101.

The sulphur pit had an air sparge type degassing system. In order to degass the extra sulphur produced it was necessary to add another air sparger. The air from the pit was sent by a steam ejector to the incinerator. The line sizes were increased to reduce the pressure drop and allow the ejector to handle the increased flow.
The final example of a revamp situation is an American refinery using oxygen enrichment incorporating the SURE Double Combustion technology, as licensed by WorleyParsons and BOC. As previously discussed, this process allows higher levels of enrichment, including the use of pure oxygen, and is design to overcome the problems of high combustion temperatures.

Of the two Double Combustion options available, the one selected for this unit was to install a new reaction furnace waste heat boiler upstream of the existing unit.

There are two sulphur recovery trains and the objective of the project was to enable the refinery, to increase the sulphur capacity of each train from 213 tpd to 426 tpd. Each sulfur plant has the BSR tail gas unit, which was designed for 213 tpd equivalent sulfur production.

Each train is designed to produce up to 426 LTPD sulfur while using oxygen and operating in the Double Combustion mode, when the other train is out of operation, and all the acid gas and SWS gas must be routed to one train.

The turndown capacity, while using oxygen is 100 LTPD. The normal operating conditions with both Sulfur trains operating, are 213 LTPD with oxygen enrichment.

To accomplish this capacity increase required plant modification together with the addition of new equipment including:

- **First Stage Reaction Furnace & Burner.** The Reaction Furnace is sized based on WorleyParsons technology by allowing enough residence time for complete destruction of ammonia in the acid gas feed.

- **First Stage Waste Heat Boiler and Steam Drum**

- **Modifying Existing Reaction Cooler with Steam Drum**

- **A common Contact Condenser Air Cooler for the tail gas unit to provide enough cooling duty for the existing Contact Condensers for the new process condition. Circulating water piping to be arranged so that water from a single train is directed to only half of the bays when the train is operating at 213 LTPD sulfur production.**

- **Lean amine air cooler in the amine tail gas unit** In order to maintain the temperature of Lean MDEA to the existing MDEA Contactor more cooling duty is required

- **Oxygen supply**
Example of a High Level Oxygen Enrichment Project

- Oxygen piping, and oxygen lances to the second stage reaction furnace
- Revised Instrumentation and control system for burner management to handle air and oxygen
- All the equipment, piping, and instrumentation were evaluated for the higher sulfur production and the necessary changes made.
- The amine circulation rate was kept the same as the current operation, however, the concentration was increased from 35%wt to 50%wt.
This paper has presented three different examples of sulphur recovery plant expansion. Each of the plants has been successfully started and is in operation.

For anyone wishing to increase the capacity of their sulphur recovery unit how should it be evaluated? It should be stated that revamps are almost always cheaper than new plants and major capacity increases without the use of oxygen are rare. Depending upon the extent of the capacity increase required the revamp options may fall readily into one of the categories discussed.

The first step to be taken should be to commission a conceptual study based on the original design documents. However, it should not be assumed that the nameplate capacity is the actual capacity.

Assuming that the revamp is to proceed then the next step is to carry out a test run on the unit. If there is more than one unit, then it may be possible to complete a capacity test on one unit by turning down the other units. Where there is only a single unit it sometimes difficult to test at full capacity due to lack of feedstock and the actual capacity has to be evaluated from the test data.

The use of an independent testing firm will give a full analysis of all the units stream and may help to identify any bottlenecks or under performing sections.

The test run data will establish the actual plant capacity and this can then be used to determine which revamp route to take, carry out preliminary design work and evaluate the actual cost of the project.
## References

1. Borsboom H.; Clark P.; 2001

   “New Insights into the Claus Thermal Stage Chemistry and Temperatures”, Brimstone2002 Sulfur Recovery Symposium, May 2-10, 2002, Banff, Alberta, Canada


   “Advanced Gas Treating to Enhance Producing and Refining Projects using FLEXSORB® SE Solvents”, Encuentra y Exposicion Internacional de la Industria Petrolera - Meeting and International Exposition of the Petroleum Industry (E EXITEP 2003), Veracruz, Mexico