INTRODUCTION

RATE was recently awarded a contract to license a new grass roots sulphur recovery project including design of the Acid Gas Removal and the Sulphur Recovery Unit, Tail Gas Treating, water dew point control unit, Thermal Oxidizer (incineration system) and the Liquid Sulphur Degassing Units for a Sour Gas Field Development (SGFD) in Asia.

The challenge in this project was to deal with a wide range of feed gas compositions of H2S and CO2 while also dealing with a lot of impurities such as COS, heavy hydrocarbons, mercaptans, and BTEX where the operation of the SRU’s with the lean gas and impurities are difficult.

The Feed Gas Composition to the Acid Gas Removal Unit (AGRU) consists of number of cases. The design had to be able to operate in all of these cases while achieving a stable operation and meeting the performance guarantees and environmental regulations.

The H2S concentration to the AGRU varies from 2.3% to 5%, the CO2 concentration varies from 3% to 6%, Benzene from 40 ppmv to 90 ppmv, Toluene from 45 ppmv to 220 ppmv, Xylene from 20 ppmv to 150 ppmv, COS from 25 ppmv to 70 ppmv, and heavy hydrocarbons up to C50+, mercaptans from 15 ppmv to 50 ppmv plus many pseudo components according to the gas analysis from more than 100 wells. The H2S concentration to the sulphur Recovery unit varies from 30% to 47% and CO2 ranges varies from 58% to 43% respectively. The gas pressure was around 100 barg at 50°C.

The sales gas has to meet the following performance guarantees.

- H2S, 4 ppmv
- CO2 less than 1.7% max
- COS 4 ppm wt maximum,
- Organic sulphur 50 ppm- wt maximum, total sulphur 60 ppm wt maximum

The Sulphur Recovery, Tail Gas Treating, Thermal Oxidizer, and Liquid Sulphur Degassing have to meet the following performance guarantees.

- Sulphur Recovery minimum 96%,
- Overall SRU+ TGU less than 50 ppmv of SO2 at stack
- Liquid sulphur Degassing, less than 10 ppmw of H2S
In this paper, we will discuss the some design features of the units and reasons why they were selected. We will discuss the options that we considered to design this unit, including the solvent evaluation & selection, the impact of H2S /CO2 ratio on the AGRU and the SRU design, and the selected optimum scheme of the Sulphur Recovery Units, and the TGU to meet the performance guarantees for all 10 cases including the operating and the capital costs comparison.

**DETAIL DESCRIPTION**

RATE executed a new sour gas field development project by developing new schemes to meet all high H2S cases of the feed compositions and to achieve a stable operation by evaluating all commercial solvents and the most economical schemes by using the best possible technologies.

**ACID GAS REMOVAL** – the highest H2S case in the design basis concentration was selected as the design controlling case. In order to meet the specification of the treated gas that is shown below, we needed to select a proper solvent. We looked at all generic solvents such as MDEA and all formulated selective solvents chemical, physical, and hybrid solvents commercially available to meet the following specifications.

- H2S, 4 ppmv
- CO2 less than 1.7% max
- COS 4 ppm wt maximum,
- organic sulphur 50 ppm- wt maximum, total sulphur 60 ppm wt maximum

As part of the acid gas removal scheme configuration, we also evaluated **Hot Flash configuration, 2-stage regeneration system and lean /semi lean configuration.**

The hot flash gas configuration evaluation was conducted by heating the amine further before entering the flash drum. This feature has been used to flash the amine and to recover the hydrocarbons. In this scheme after the first flash drum, the rich amine enters the lean/rich exchanger and then enters to another heater to heats up then enters to the second flash drum to flash the gas. The gas leaving the second flash drum is cooled and enters an additional flash absorber to separate the hydrocarbon using lean solvent. Adding several pieces of equipment to the amine unit, would result higher capital cost thus we concluded that the hydrocarbon recovery is not significant and it is not cost effective.

In addition, by selecting the proper solvent, there is no need for the 2-stage regeneration scheme and there is no need for lean /semi lean scheme which additional equipment would increase the capital cost.

After extensive technical and cost evaluation, we concluded that formulated based MDEA chemical solvent can meet all of the project requirements and will remove CO2, H2S and COS to meet requested specifications to less than 150 ppmv CO2, H2S less than 1 ppmv and COS to less than 4 ppmv in the treated gas. We performed the extended evaluation on the solvent selection from generic to formulated or selective solvents by all solvent suppliers and our recommendation was formulated MDEA based chemical solvent versus hybrid solvent where this solvent has the flexibility to meet the product specification for all cases.
The proposed solvent unit resembles a conventional amine-type acid gas removal flow scheme, but utilizes a formulated MDEA-based solvent (45%wt), and employs proprietary RATE technology. The thermal or flash regenerated solvent process removes CO$_2$, H$_2$S and other sulphur components. Hydrocarbon losses are minimal in the process due to their low solubility in the selected solvent.

The advantages of the formulated MDEA based chemical solvent versus hybrid are described below:

- Meet the project specification for all cases
- No need for polishing unit before or after amine unit
- Less circulation and less capital cost
- A Proven technology and optimized design based on over 50 years operating experience
- Ability to handle a wide range of inlet H$_2$S and CO$_2$ concentrations with minor operating adjustments and no hardware modifications
- No need for 2-stage regeneration, and no need for lean/semi-lean configuration, no need for hot flash configuration
- System designed for high reliability, flexibility and required design turndown without the need for hardware modifications
- Availability of the solvents at regional supply facilities for fast response to customer needs
- Capability and availability of sharing operational data from other facilities using the same solvent

Figure 1 and 2 represents the simplified Amine unit process flow diagram.
Figure 1 represents a typical amine unit where the sour gas flows through a gas filter prior entering the absorber. The bottom of the absorber rich solvent flows to the rich amine flash drum through a control valve to reduce the pressure prior entering the flash drum to separate the hydrocarbon. The flash gas flows to the incinerator where the rich solvent flows to L/R Exchanger prior entering the tail gas regeneration system.

Figure 2 represents a typical regeneration system to separate the H2S and recycle back to the reaction furnace. Air cooler is the only media is used to cool the lean solvent.
SULPHUR RECOVERY & TAIL GAS TREATING

The duty specification for the SRU represents different feed compositions where the H2S varies between (31-68) % and CO2 varies between (58-21) % respectively.

We also have mercaptans and possible hydrocarbons that can have a significant impact on the sulphur recovery operation.

BTEX / Mercaptans are removed from acid gas feed upstream are fed to the sulphur plant. The low concentration of H2S will make it difficult for the reaction furnace of the sulphur plant to attain and sustain a stable flame. In addition, the high concentration of hydrocarbons in the acid gas feed would require a flame of high temperature to ensure complete hydrocarbon destruction. The actual amount of these contaminants will depend on the type of acid gas removal solvent to be used upstream of the sulphur plant. These hydrocarbon contaminants tend to crack in low flame temperatures, and causes carbon lay down problem in the catalyst bed of the first sulphur converter. Carbon laydown associated with cracking of heavy hydrocarbons has caused a significant number of unplanned shutdowns. A minimum temperature of 1050 °C is required to assure that the mercaptans are destroyed; therefore, our design reflected such requirements.
We assumed the following features as part of our design.

1. **Three stage versus two stage Claus**

   It was required to achieve 96% sulphur recovery for all 10 cases. Due to the lean H2S concentration, in order to achieve and to guarantee 96% recovery, the calculated sulphur recovery should be above 96%; therefore, we needed a three stage Claus to meet the recovery for all cases. Please refer to the sulphur recovery chart as shown above.

2. **Start of Run (SOR) and End of Run (EOR)**

   As the sulphur recovery operates, the catalyst ages and become less active, therefore, throughout the life of the Claus catalyst (4-6 years) the sulphur recovery will be decrease from SOR to EOR. In order to maintain a reliable operation, it is our normal practice to design the tail gas unit based on EOR operation of the sulphur plant assuming more H2S enters the tail gas unit.

3. **Acid gas feed preheat and combustion air preheat**

   Preheating acid gas feed and combustion air will certainly help to increase the flame temperature of the Claus reaction furnace. However, these design measures are not capable of raising the flame temperature to the desired level. A combination of these techniques with another technology is required to achieve the desired stable and high temperature flame in accommodating the specified acid gas feeds of various levels of H2S and heavy hydrocarbons contents.

**FEATURES EVALUATED BUT NOT USED**

We performed the simulation for all cases in order to achieve the desired combustion temperature; we had to come up with the most cost effective option as part of our evaluations.
There are key elements that we considered and we explain the reasons what and why these features were not selected.

- **Conventional Acid Gas Enrichment** - Acid gas amine type enrichment unit upstream of SRU common regeneration unit with the TGU/amine unit will consist of a common acid gas pretreatment step with selective solvent based MDEA formulated solvent, and a three stage Claus units followed by a common tail gas/amine tail gas treating unit, to enrich the gas to the SRU and to reduce the SRU equipment. The acid gas enrichment works very well where the H2S concentration are very low, for this project, the H2S concentration for high H2S cases varies from 46% to 68% and the CO2 concentration varies from 43% to 21%, corresponding from all cases H2S varies between (31-68)% and CO2 varies between (58-21)%.

  Acid gas enrichment is very effective where the H2S/CO2 ratio is low and will achieve high CO2 slip. The H2S is enriched and SRU unit will be smaller. In this project, we had high H2S/CO2 ratio and CO2 slip will decline therefore, the SRU will not be smaller and the acid gas enrichment are not very effective. In other words, after a detailed evaluation, we concluded that adding conventional acid gas enrichment would be not as effective as generally expected.

- **Supplemental fuel gas burning** - Supplemental natural gas burning is a technique often employed to raise the flame temperature while processing acid gas feed of low H2S concentrations in a sulphur plant, where we did not have natural gas available in this project, and using fuel gas will cause soot formation.

- **Oxygen Enrichment** - Oxygen enrichment technology is an ideal technology for raising the flame temperature of the Claus reaction furnace especially for acid gas of low H2S concentration, however, Oxygen was not available in the project.

- **Using cooling water** - Using cooling water is not practical in summer due to hot climate weather temperature; therefore, water cooler usage in the tail gas unit would not help in summer.

### Proprietary Acid Gas Enrichment & SRU Schemes

We developed a unique acid gas enrichment process called (Rich “S-MAX”), and a unique proprietary scheme of the reaction furnace which are described below.

We used a proprietary “S-MAX” 2-zone reaction furnace, which is different with the conventional scheme, with high intensity burner. In the conventional scheme of the 2-zone reaction furnace, hydrocarbons and mercaptans will be bypassed to the second zone where the combustion temperature would not be adequate for the heavy impurities destruction, which will cause soot formation and catalyst deactivation.

The combustion temperature is always higher in the first zone than the second zone. From the above evaluation, we concluded that we could use exclusive schemes by RATE – For the high H2S cases, where the H2S concentration is higher as described below. In addition in order to meet the tight stack
SO2 specification, the tail gas unit should be designed with formulated selective solvent MDEA based formulated solvent.

We split the acid gas from the amine unit where up to 75% of the amine gas entered the first zone of the reaction furnace and up to 25% of the acid gas is routed to the tail gas absorber in addition to the quench overhead that flows to the tail gas absorber. The tail gas amine unit is designed with the much higher amine loading similar to the amine unit, so in Summary:

- 25% of the amine acid gas is sent to the tail gas absorber
- 75% of the amine acid gas is sent to the FIRST ZONE OF THE REACTION FURNACE
- The tail gas absorber operates at higher rich loading (0.2-0.3 mol/mol)
- The tail gas recycle from the tail gas regeneration unit is recycled to the SRU but not to the first zone, instead:
  - The acid gas from the tail gas regeneration column, which is hydrocarbon / mercaptan free, is recycled back to the SRU. It is preheated and flows to the second zone of the reaction furnace. The combusted gas from the zone 1 reaction furnace flows to the second zone through choke ring where the temperature is above ignition temperature, and burn the acid gas in the second zone and the combusted temperature leaving the reaction furnace is 990°C according to our simulation for the high H2S cases.
- Our tail gas absorber was designed with 0.2 to 0.3 mol/mol loading.
  - The acid gas loading in the tail gas absorber is normally 0.1 mol/mol maximum, and the acid gas loading for the amine absorber is normally 0.3 mol/mol, it means there is significant free amine in the tail gas absorber to process the portion of the acid gas. The tail gas absorber acts not only as a tail gas absorber but also as an enriched absorber without adding significant cost to the project. This scheme also removes the hydrocarbons / mercaptans, which cause problems in the second zone of the reaction furnace. As H2S concentration increases for case 2 through case 6, the 25% slipstream from the SRU feed to the tail gas absorber may be reduced as long as we achieved 1100°C -1150°C combustion temperature in the first zone of the reaction furnace.

For cost estimate purposes we also evaluated if the tail gas absorber would be designed with 0.1 mol/mol rich loading instead of 0.3 and it is obvious that we needed to have much higher solvent circulation and much higher capital cost.

In addition to above challenges, due to hot temperature in summer using cooling water is not adequate to achieve low H2S concentration less than 10 ppmv from the absorber overhead therefore we evaluated 2 options tail gas design with or without chiller.
As part of our evaluation, we compared chemical solvent versus hybrid solvent with or without chiller, with or without partial enrichment which translate 0.1 mol/mol rich loading versus 0.2-0.3 mol/mol rich loading as summarized below in table 1.

**TABLE 1 – SOLVENT COMPARISON**

<table>
<thead>
<tr>
<th>Vendor ---TGU – Evaluation</th>
<th>Formulated</th>
<th>Formulated Co2 based</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solvent</td>
<td>Chemical solvent</td>
<td>Hybrid and other solvents</td>
</tr>
<tr>
<td>Circulation rate</td>
<td></td>
<td></td>
</tr>
<tr>
<td>TGU only – with chiller, 10 ppm H2S</td>
<td>1400 gpm</td>
<td>1210 gpm</td>
</tr>
<tr>
<td>TGU + partial Enrichment – with chiller, 10 ppm H2S</td>
<td>2500 gpm</td>
<td>4900 gpm</td>
</tr>
<tr>
<td>TGU only – without chiller</td>
<td>1600 gpm</td>
<td>1450 gpm</td>
</tr>
<tr>
<td>TGU + partial Enrichment – without chiller 0.1 rich loading</td>
<td>2400 gpm</td>
<td>NA /(high)</td>
</tr>
<tr>
<td>TGU + partial Enrichment – without chiller 0.25 rich loading, 80 ppm</td>
<td>1800 gpm</td>
<td>NA / (high)</td>
</tr>
</tbody>
</table>

As noted in above table, we selected the tail gas design as a partial enrichment, using chemical solvent without any cooling water trim cooler nor chiller and we achieved 80 ppmv of H2S in the treated gas from the tail gas absorber overhead.

As shown above, our evaluation on the tail gas unit was very comprehensive considering four schemes.

We looked at the design of the tail gas to achieve typical 0.1 mol/mol rich loading including processing the amine acid gas as partial enrichment with or without chiller.

We looked at the design of the tail gas unit to achieve higher loading 0.3 mol/mol rich loading including processing the amine acid gas as partial enrichment with or without chiller.

As noted above, we needed 2400 gpm if we used 0.1 mol/mol rich loading with partial enrichment while we needed 1800 gpm of the solvent circulation with partial enrichment and both case without the chiller which means basically the partial enrichment would not add additional cost to the scheme but also there was a cost reduction due to lower circulation rate.

The partial enrichment Rich "S-MAX" shows significant improvement of the sulphur recovery unit as well as reduction cost compare to conventional tail gas treating design.

The following figures 3 and 4 represent the unique proprietary design of the reaction furnace in the sulphur recovery unit.

Figure 3 shows the proprietary design of the reaction furnace where the acid gas is recycled from the tail unit is preheated before entering the second zone of the reaction furnace.

Figure 4 shows a typical 3-stage Claus unit to meet the requirements of 96% minimum recovery. It is important to point out the first reactor contains Titanium catalyst to improve the hydrolysis of COS and CS2 due to high CO2 content and to improve the overall sulphur recovery.
FIGURE 3 – PROPRIETARY DESIGN OF THE REACTION FURNACE
However, according to the project specification, the Our stack emission should be less than 50 ppmv of SO2, we concluded that if we have chiller with formulated solvent we can meet 50 ppmv of SO2 and without the chiller we need the caustic scrubber after incineration to meet the SO2 emissions as summarized below in table 2.
TABLE 2 - CAUSTIC SCRUBBER VERSUS CONVENTIONAL STACK

<table>
<thead>
<tr>
<th>Case</th>
<th>Case 1</th>
<th>Case 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chiller</td>
<td>NO</td>
<td>YES</td>
</tr>
<tr>
<td>Incineration type</td>
<td>Caustic Scrubber</td>
<td>Stack</td>
</tr>
<tr>
<td>COS hydrolysis reactor</td>
<td>NO</td>
<td>YES</td>
</tr>
<tr>
<td>Tail Gas Amine circulation</td>
<td>Lower, with higher loading</td>
<td>Higher, lower lean loading</td>
</tr>
<tr>
<td>Recovery</td>
<td>99.99%</td>
<td>99.9%</td>
</tr>
<tr>
<td>Reliability, operability</td>
<td>High</td>
<td>Mid</td>
</tr>
<tr>
<td>Capital / operating cost</td>
<td>Lower (no chiller, less power, less circulation)</td>
<td>Higher (chiller, more power, more circulation)</td>
</tr>
</tbody>
</table>

In the quench system, we could reduce the water circulation rate from 2300 gpm to 1500 gpm due to adding a chiller to cool the water to 37°C (100°F). In the amine section of the tail gas unit, we needed to increase the circulation rate from 1800 gpm to 2500 gpm, to meet the H2S overhead of 10 ppm even though we had a chiller on the lean amine circuit we needed more circulation.

The total Chiller duty is 35 x 10^6 kcal/hr (140 MMBtu/hr) with the 2500 KW chiller system for one train.

The following figures 5, 6, and 7 represent the Rich” S-MAX” partial enrichment tail gas unit and the forced draft incineration with Caustic Scrubber.

FIGURE 5 – TAIL GAS UNIT RICH”S-MAX” & HYDROGENATION SCHEME
Figure 5 represents the tail gas treating unit which starts with indirect steam re heater with the low temperature hydrogenation catalyst, then it is followed by the quench system where the gas is cooled off and then is routed to the Rich "S-MAX" absorber. Cooling is only by air cooler since cooling water would not be effective in summer. The Rich "S-MAX" absorber receives the cooled gas from the quench tower overhead plus up to 25% amine acid gas from the amine unit.

FIGURE 6 – RICH"S-MAX” TAIL GAS UNIT REGENERATION SYSTEM WITHOUT CHILLER
Figure 7 represents the forced draft incinerator with the incinerator waste heat boiler to recover the heat for energy efficiency followed by the venturi scrubber and caustic scrubber tower, which was designed by RATE. The remaining H2S and sulphur species were directed to incinerator and converted to SO2. This scheme provides ZERO SO2 emissions. Caustic is known as the best absorbent to absorb the SO2. The gas leaving the caustic overhead is H2S free, SO2 free and sulphur free. The bottom of the caustic scrubber contains sodium bisulfite is cooled and pumped to waste water treatment.

**Hydrolysis Reactor in Tail Gas Design**

We had also to add the hydrolysis reactor after the hydrogenation reactor in the tail gas unit; because overtime the tail gas hydrogenation catalyst loses efficiency and the COS & CS2 hydrolysis will decline resulting emission increase and would have impact on the unit performance.
As we know, since the feed gas composition to the SRU is not rich in H2S, one of byproduct from the reaction furnace is COS. Even though, we have some titanium catalyst in the first SRU reactor for COS / CS2 hydrolysis, we have seen significant COS in the tail gas stream in similar applications. We also know that the hydrogenation catalyst as known as CO-MO catalyst hydrolyze the COS if the reactor temperature is high such as inline burner, where the inlet temperature is lower using steam re heater and low temperature catalyst. Even though the simulation would not show significant COS from the hydrogenation reactor outlet, the operational data and our experience indicate, that COS could be in a range 30-40 ppmv at equilibrium condition.

As the results we concluded that if we design the unit without the caustic scrubber, then we need the hydrolysis reactor to make sure all of the sulphur specious will be hydrolyzed in addition we need to have a chiller to reduce the lean amine temperature so the tail gas absorber overhead will meet 10 ppmv of H2S and we need to design the tail gas unit based on 0.1 mol/mol rich loading with resulting higher circulation rate.

The following table 3 shows comparison of the tail gas unit using chemical solvent with or without a chiller.

**TABLE 3 – COMPARISON BETWEEN CHILLER AND NO CHILLER**

<table>
<thead>
<tr>
<th>Item</th>
<th>Case 1</th>
<th>Case 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Amine unit</td>
<td>3000 gpm</td>
<td>3000 gpm</td>
</tr>
<tr>
<td>TGU + partial Enrichment – without chiller 0.25 loading, 80 ppm</td>
<td>1800 gpm</td>
<td>----</td>
</tr>
<tr>
<td>TGU + partial Enrichment – with chiller, 0.1 loading, 10 ppm H2S</td>
<td>----</td>
<td>2500 gpm</td>
</tr>
<tr>
<td>Tail gas unit- reactor</td>
<td>COMO reactor</td>
<td>COMO + Hydrolysis reactor</td>
</tr>
<tr>
<td>Tail Gas unit - quench</td>
<td>No chiller</td>
<td>Added chiller, circulation reduced</td>
</tr>
<tr>
<td>Incinerator</td>
<td>Forced draft</td>
<td>Forced draft</td>
</tr>
<tr>
<td>Venturi / caustic scrubber</td>
<td>82 ppm to zero ppm of SO2</td>
<td>None</td>
</tr>
<tr>
<td>Incinerator stack</td>
<td>Less than 50 ppm of SO2</td>
<td></td>
</tr>
</tbody>
</table>

In summary, we have developed a scheme that to partial enrich the acid gas for a reliable operation and with the flexibility for various feed compositions without investing for full enrichment unit, as noted above, the full enrichment unit would increase the capital cost significantly. We take advantages of the available free amine in the tail gas unit by increasing the loading from 0.1 to 0.3. The tail gas flow rate from the quench overhead is significantly larger than the slipstream from the SRU unit; therefore, bringing a 25% of the amine acid gas to the tail gas absorber is tolerable.

Based on our details calculations, we concluded that our unique scheme would meet all the requirements for case 1 through 6 of the design basis but we will still have problems with the flame stability for case 7 and 8 where we need to boost the temperature even further even though natural gas is not available at the facility but we could use fuel gas.
Figures 8, 9, and 10 represent Rich “S-MAX” partial enrichment with the chiller, and the hydrolysis reactor plus the forced draft incineration with the conventional stack.

**FIGURE 8 – RICH “S-MAX” PARTIAL ENRICHMENT WITH CHILLER AND HYDROLYSIS REACTOR**
FIGURE 9 - RICH"S-MAX" TAIL GAS UNIT REGENERATION SYSTEM WITH CHILLER
FIGURE 10 – FORCED DRAFT INCINERATION WITH STACK

RATE special scheme & Proprietary Burner lean gas application

In this scheme only for cases with very lean H2S, we developed the scheme to burn the fuel gas (NOT NATURAL GAS) with excess air first in the high intensity burner with zero to small portion of the acid gas in the burner. Most or all of the acid gas goes to the first zone of the reaction furnace not to the burner for the lean acid gas cases.

The concept is in this configuration the fuel gas and excess air burns stoichiometric (while conventional SRU is sub-stoichiometric) so there is no soot formation and the composition of the fuel gas is not important because we have excess air while we process the lean gas. As the acid gas changes to higher H2S concentration, we reduce the fuel and air and more acid gas goes to the burner. In this configuration, our reaction furnace will have with more residence time than the conventional design twice or more even though there is no ammonia but for destruction of BTEX. The remaining scheme would be the same, spike of fuel gas with excess air as needed.
Adding all these features together, would allow us to have stable flame temperature for all cases of operation.

**TAIL GAS TREATING SYSTEM** - The tail gas unit is designed based on low temperature hydrogenation catalyst so a steam reheater is used to heat the tail gas instead of an inline/RGG burner.

The quench system could be one stage column using ammonia bottle injection or it could be design with 2-sections using caustic the first section de-superheats the gas and scrub any SO2 may breakthrough from hydrogenation reactor, and the second section cools the gas and condensate the water.

The tail gas absorber is also designed for partial H2S enrichment for up to a 25% slipstream of the SRU feed without any additional circulation rate.

**INCINERATION** – it is a forced draft incineration with the heat recovery.

**Option 1** - As we noted above we are using formulated amine solvent for the tail gas amine unit. If we will have a chiller, then the quench water, and the lean amine could be cooled where we can meet 10 ppmv of H2S from the tail gas absorber, plus the unconverted COS from the hydrogenation reactor, plus a COS hydrolysis reactor, we would be able to meet the 50 ppmv of SO2 in the stack without caustic scrubber. In this case, having a chiller is a **MUST**.

**Option 2** - As we noted above we are using formulated amine solvent for the tail gas amine unit. If we don’t have a chiller, then the H2S from the absorber OVHD would be more than 10 ppmv, plus the unconverted COS from the hydrogenation reactor goes to the incinerator without COS hydrolysis reactor and the caustic scrubber has to be provided to meet the emission of 50 ppmv of SO2. In this case, the caustic scrubber is a **MUST**.

Please note the followings:
- In order to eliminate the caustic scrubber, we needed to have chiller to the quench system and lean amine circuit in the tail gas unit
- In order to eliminate the caustic scrubber, the rich / lean loading had to be kept very low to meet the absorber overhead below 10 ppmv of H2S

**LIQUID SULPHUR DEGASSING**

Liquid sulphur degassing inside of the pit and outside of the pit was evaluated to achieve 10 ppmw of H2S in liquid sulphur. External degassing outside of the sulphur pit was selected to reduce the sulphur pit sizing.

**COST ESTIMATE COMPARISON**

After we completed the technical evaluation and we selected the tail gas partial enrichment we did the cost estimate for 2 cases, case 1 without the chiller and with caustic scrubber versus case 2 with the chiller, hydrolysis reactor and the regular stack and the results are provided below in table 4.
TABLE 4 – CAPITAL AND OPERATING COST COMPARISON

<table>
<thead>
<tr>
<th>Case</th>
<th>Case 1 (Rich S-Max) without chiller, without hydrolysis reactor, with caustic scrubber, rich loading 0.25 mol/mol</th>
<th>Case 2 (Rich S-Max) with chiller, hydrolysis reactor and without caustic scrubber Rich loading of 0.1 mol/mol</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Capital Cost</td>
<td></td>
</tr>
<tr>
<td></td>
<td>100</td>
<td>115</td>
</tr>
<tr>
<td></td>
<td>Utility, based on 5 years operation</td>
<td></td>
</tr>
<tr>
<td></td>
<td>100</td>
<td>125</td>
</tr>
<tr>
<td></td>
<td>Catalysts and Chemicals based on 5 years operation</td>
<td></td>
</tr>
<tr>
<td></td>
<td>100</td>
<td>116</td>
</tr>
</tbody>
</table>

CONCLUSIONS

The drivers for the development of Sour Gas Field Development (SGFD) are based on:

a) Increasing energy costs
b) Growing demand for the natural gas
c) Higher demand to treat sour gas fields
d) 40% of the world natural gas reserves are in form of H2S and CO2 and undesired components
e) Economics

According to the design base discussed in this paper, the sour gas field development was designed considering special proprietary schemes and features to handle a wide range of operating cases and to maintain the stable operation while meeting the required performance. We concluded that the formulated chemical solvent MDEA based are more effective than hybrid solvents. In addition the tail gas absorber could be designed with a higher rich amine loading to achieve better results.

RATE offered the following technologies and features for this project.

- AGRU – Acid Gas Removal – formulated MDEA based Chemical Solvent
- Dehydration unit – TEG solvent was used with gas stripping configuration to achieve water content, of 7lb/ MMscf maximum
- Sulphur Recovery Unit
  1. Acid gas and air preheater for more stability
2. 3-stage reactor to achieve 96% recovery
3. RATE Proprietary Special 2-zone reaction furnace & high intensity burner
4. Ti catalyst in the first bed
5. High H₂S content – 2-zone RATE special design of reaction furnace & Rich "S-MAX" Partial enrichment
6. Low H₂S content – 2-zone RATE special reaction furnace & Partial enrichment and RATE special scheme with fuel gas supplement with excess air

- Tail gas Treating Unit and Incineration
  - Low Temperature catalyst TGU, with steam reheater instead of inline burner
  - Single Stage Quench system
  - Formulated TGU solvent, selective based MDEA solvent
  - TGU absorber with partially enriched design higher rich loading
  - Start up Eductor to minimize flaring during the start up
  - RATE guaranteed less than 30 ppmv of SO₂ emission with (ZERO EMISSION)
    - With caustic scrubber, no chiller, no COS hydrolysis reactor (Option 1)
    - No caustic scrubber, More amine circulation rate, COS hydrolysis reactor, chiller (Option 2)

As was described we selected formulated based MDEA chemical solvent for the amine unit versus hybrid solvent.

In regard to SRU / TGU design, we also chose a selective MDEA formulated solvent with option 1 where there was no chiller and no hydrolysis reactor with less solvent circulation with higher rich loading and we used caustic scrubber after incineration to meet the SO₂ emission of zero.

This option it seems more robust due to feed variation if the H2S in the absorber overhead exceeds the caustic scrubber would absorb the SO₂ and the SO₂ emission will always be met. In addition according to the cost estimate the operating and the capital cost is lower than the option 2.