

1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21
22
23
24
25
26
27
28
29
30
31
32
33

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SULPHUR

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India's refining industry
Sulphur recovery project listing
Sulphur management options
Sour amine systems



CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

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ISSUE 368
JANUARY-FEBRUARY 2017

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Dew point measurement

Better control in copper smelters and sulphuric acid plants.



Amine system failures

Strategies to tackle the most common issues.

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SULPHUR

www.sulphurmagazine.com

NUMBER 368

JANUARY | FEBRUARY 2017

CONTENTS

10 India's refining industry

India's refining capacity has nearly tripled in 15 years, and the country has emerged as a major refining hub, with a knock-on effect on sulphur output.

12 Sulphur 2016 index

A full listing of all news items and articles published in Sulphur last year

14 Sulphur recovery project listing

Sulphur's annual survey of recent current and future sulphur recovery unit construction projects maps the developing shape of brimstone production from fuel and gas processing plants worldwide.

16 Sulphur 2016 conference

A review of the 31st Sulphur International Conference and Exhibition, held at the Hilton London Metropole, London, UK, in November 2016.

18 Acid dew point measurement and control

Acid dew point measurement of sulphur condensables is desirable for controlling critical process parameters as well as plant safety in processes where SO₃ gas is present. C. Chothani of Breen Energy Solutions discusses the application of acid dew point measurement technology to a copper smelter in Spain as well as the development of a new sensor for use in sulphuric acid plants.

19 Sulphur management for an unconventional gas-condensate field

M. Jacques of Technip and A. König-Adolph of ENGIE E&P review the main outcomes of a conceptual study for a gas-condensate field development located in Germany, processing raw gas with a low content of H₂S, moderate H₂S to CO₂ ratio and low H₂S to organic sulphur compounds ratio.

22 Why amine systems fail in sour service

Amine systems failures have cost the industry billions of dollars over the last 30 years. In this article, P. le Grange, M. Sheilan and B. Spooner of Amine Experts draw attention to the highest probability threats to a facility's operability and reliability.

27 Simplification of the modified Claus process

P. D. Clark et al of ASRL describe modifications to the existing Claus sulphur recovery process which aim to simplify the overall design and increase the efficiency of the system. The key modification is to incorporate a catalytic section in the combustion chamber to enhance hydrocarbon and CS₂ conversion.

REGULARS

- 3 Editorial
- 4 Price Trends
- 5 Market Outlook
- 6 Sulphur Industry News
- 7 Sulphuric Acid News
- 9 People/Calendar

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

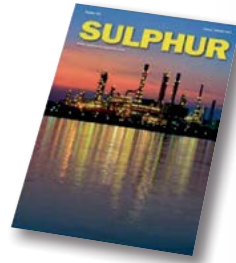
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Has OPEC rescued oil prices?



As this issue of Sulphur was going to print, OPEC managed to finally agree a deal to cut oil production and try to bring the oil market back into balance. While many have been sceptical about previous OPEC deals, which have often been widely ignored in practice by members of the cartel, one of the key features of the current deal is that it also involves eleven large non-OPEC producers, including Russia, Mexico and Kazakhstan. Indeed, the 600,000 barrel per day output reduction by non-OPEC producers, which is due to come into effect from January 1st 2017, will see Russia responsible for 300,000 bbl/d of that cut. It comes shortly after OPEC's November 30th meeting which agreed to cut OPEC output by around 1.2 million bbl/d to 32.5 million bbl/d, also from January 1st. The stumbling block had been rivalry between Saudi Arabia and Iran, but Saudi Arabia has now agreed to reduce output by 500-700,000 bbl/d to 10 million bbl/d (500,000 bbl/d on average figures, but 700,000 bbl/d compared to November production), while Iran, which had been ramping production back up after the easing of international sanctions, will freeze output at its current level. Nigeria and Libya, both of which have suffered from sanctions and oil infrastructure sabotage, are exempt from the cut. Overall, OPEC is predicting that – assuming everyone sticks to their quotas (which is by no means a given), increasing demand will finally catch up with oil production in the second half of 2017, and the glut of stocks around the world will start to draw down as prices climb.

Oil prices rose briefly to \$60/bbl in the wake of the announcement, but have since slid back below \$55/bbl because of news of much higher than expected stocks in the US. US shale/tight oil production, which the Saudis had evidently hoped to put out of business by keeping the taps on back in 2014, has proved in the long run to be much more resilient than the economies of OPEC and Russia, all of which have relied heavily on high oil prices to fund investments and public services, as well as, in some case, foreign conflicts. Russia has been cushioned somewhat by the fall of the rouble against the dollar, but Saudi Arabia has had to dig deep into its huge foreign reserves to maintain public spending, and some OPEC nations like Venezuela have virtually collapsed under the strain of low oil prices. US producers meanwhile, while they have finally been forced this year to cut production by about 1 million bbl/d to 8.6 million bbl/d, according to EIA figures,

have merely been made leaner and fitter by the run of low prices, and found more efficient ways of working. OPEC's strategy now seems to be to let the oil price rise a little – to around \$60/bbl/, but not sufficient to let too much US production back on-stream. This was the price level that International Energy Agency said earlier in 2016 would trigger a recovery in US oil production. However, if US producers are able to reduce production prices still further, it remains to be seen if OPEC's gamble will pay off.

Sulphur markets of course operate at two removes from oil, via sulphur recovery at refineries, and the two year decline in oil prices has not, as yet, led to any major impact on sulphur. The falls in US oil output have come from relatively low sulphur sources – 90% of US shale oil has a sulphur content of less than 0.3% – while the break-even cost of Canadian oil sands production has so far remained below the price of oil, especially for newer steam-based rather than mined production, so that while some new expansions and upgraders have been postponed or cancelled, production has generally continued. Likewise while refining margins have slowly been squeezed by the fall, especially in the US where refiners had effectively had access to 'landlocked' oil at below international market prices, operating rates have stayed relatively constant – even increasing in some cases where lack of oil storage has prompted more conversion to finished products for which storage capacity remains available. So while removal of 500-700,000 bbl/d of sour Saudi crude from the market, averaging around 2% sulphur content, in theory represents 450-600,000 tonnes per year of sulphur that isn't reaching refiners, oil stocks remain high around the world and refiners are still having to recover more sulphur due to tightening regulations on sulphur content of fuels. The greatest impact of a recovery in oil prices – assuming it can be sustained – is likely to be in the longer term, as previously postponed or cancelled projects begin to look viable again. ■

Richard Hands, Editor



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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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Price trends



MARKET INSIGHT

Meena Chauhan, Research Manager, Integer Research (in partnership with ICIS) assesses price trends and the market outlook for sulphur.

SULPHUR

Firmer pricing, increased uncertainty

Global sulphur markets showed no signs of waning through November, with firmer prices posted in the Middle East across the board for both November and December. The backdrop of weak demand in some regions has led to questions over how sustainable this run up in prices is likely to be going into the new year. The main driver appears to be bullish sentiment and comfortable inventories for producers in major exporting regions. Project delays in Q4 2016 for new sulphur supply has also served to buoy the short term outlook for pricing. Improvement in demand in the processed phosphates markets has been absent, with downstream pricing remaining under pressure due to the lack of Indian demand.

In 2016, we have seen a major supply shift in the global sulphur market, as the Middle East has been the leading producing region of sulphur, with North America now second. There is increasing focus on price dynamics in the Middle East region as a result. Monthly contract price postings for December saw increases on November pricing, further pushing up expectations for delivered prices in key importing markets. Adnoc announced a \$6/t increase to \$88/t f.o.b. Ruwais for the Indian market for December. In Qatar, Tasweeq posted its price at \$92/t f.o.b. Ras Laffan, up by

\$10/t on November. The price was in line with the tender price award and the firmer delivered prices in China. Over in Saudi Arabia, Aramco Trading also increased its price by \$9/t to \$92/t f.o.b. Jubail.

On the supply side, the outlook for 2017 remains firm, with new projects in Qatar and Kazakhstan set to increase the availability of sulphur in the export market. However, due to project delays, the influx of fresh tonnage may not emerge until Q2-Q3 2017, leading to the potential for a stable Q1 2017. However, demand is likely to be the major bearish factor as recent reports that Moroccan phosphate output will be lower than usual in the November to January period due to maintenance turnarounds. On the one hand, this would mean reduced sulphur consumption but may provide some support to the phosphates market. However, looking ahead to 2017, the start-up of the next phase of the Ma'aden phosphates expansion could put further downward pressure on an already finely balanced market.

Spot prices in China moved back up to three digits, up at \$105/t c.f.r. Offers above this level have been heard, but no business heard at the start of December. With the boost in Middle East benchmarks, prices may temporarily firm in China, but as phosphates in China have also been a challenge, we see the potential for a price correction in the short term. At the same time, the raw material trade statistics reflect

continued strong imports of sulphur and sulphuric acid through the year to date. In January-October sulphur imports exceeded 10 million tonnes and are up by 7% year on year, also up on 2014 levels. Sulphur market share in China is dominated by Saudi Arabia, stable at around 20% to date. However, volumes from the UAE have grown by over 90%, representing 15% of China's imports this year. This compares to a year earlier, with only 8% market share.

Sulphur inventories at China's nine major ports have eroded, down to around 1.5 million tonnes at the start of December. This compares to highs of around 1.8 million tonnes in 2016. As sulphur prices have increased in recent months, we expect to see further erosion of stocks, with the potential for imports to China to also slow down.

Indian sulphur prices also firmed to three digits, up at \$99-102/t c.f.r at the end of November, in line with the Middle East and China price changes. Tender activity in recent weeks included FACT with a 25,000 tonne requirement for December arrival at Cochin, awarded at \$101/t c.f.r. Meanwhile, SPIC was reported to have awarded a sale at \$102/t c.f.r for a November cargo. Major end user IFFCO was heard comfortable for sulphur, with healthy supply of molten sulphur from domestic refineries.

In a bright spot for sulphur demand, Terrafame in Finland has reported achieving its key targets at below budgeted costs at its mining operations in Sotkamo and is ramping up its sulphur and sulphuric acid consumption. The company purchases direct acid but also elemental sulphur for its bioleaching operation. Due to the positive progress in recent month, the project will no longer be at risk of closure. The

Fig 1: Chinese sulphur port inventory, Jan 15 to Nov 16

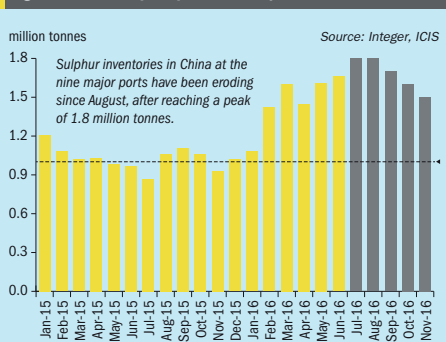
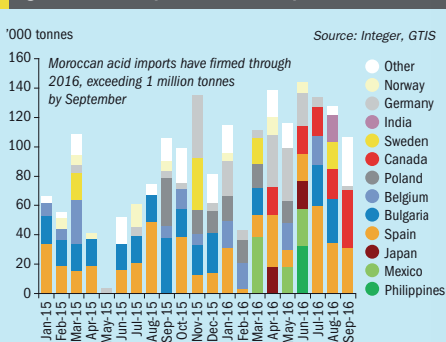


Fig 2: Moroccan imports, Jan 15 to Sep 16



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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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government is seeking a third party to partner at the operations for the new year.

Trade from Vancouver has ticked down slightly in January-October 2016 data, with ~44,000 tonnes below year ago levels. A downturn has been seen in trade to Australia this year, with a drop of around 108,000 tonnes. At the same time, volumes to China have firmed, with this leading the rankings for markets to date. This represents a reversal of trends in recent years, as Australia had been the number one market for Canadian volumes and China second. With shifting dynamics across regional supply, we expect to see further trade route shifts in the outlook.

As expected, US sulphur supply has continued to recover through into August, with a net increase year on year, bringing January-August production to just over 6 million tonnes. This has come on the back of heavier crudes as feedstock compared to a year ago. Despite the total increase, there are regional variations and gas based output reflects a decrease. The round of turnarounds at oil refineries in the final quarter of the year is still expected to slow sulphur exports and production, providing a firmer footing on pricing in the short term.

SULPHURIC ACID

Short term stability

One of the leading questions for the 2017 sulphuric acid market outlook surrounds trade and demand in Morocco. Imports of direct sulphuric acid to Morocco exceeded

1 million tonnes in September this year and are set to land at around 1.2-1.3 million tonnes by year end. This is unprecedented in volume terms and in the wide range of sources acid has been imported from. Suppliers this year have included the Philippines, Canada, Mexico and even Japan and China. European acid producers have been supported by demand this year and in more recent months Morocco has become a key outlet following the loss of the Cuba market as a major market. Going into 2017, there is no clarity on the volume of acid Morocco will likely import. Estimates in the market range from as little as 500,000 tonnes, which represent more traditional, historical levels, up to 1 million tonnes. With the variance on estimates so wide, it is hard to dismiss this as the market wildcard for the month ahead.

For now, Northeast European sulphuric acid producers remain comfortable – sitting on low inventories at smelters, healthy offtake through December from Brazil and spot prices stable ranging \$5-20/t f.o.b. Export prices have been stable through November at this level and into the start of December. There is little pointing to changes in the stable outlook, particularly with firming sulphur prices. However, if intake from Morocco drops next year, this will pose a question in trade flows from Europe.

One potential outlet will be the US market, as a heavy turnaround schedule in 2017 at smelters in Canada and the US will lead to disruptions in supply. The

impact of some of these are likely to be regional, with pockets of tightness. However, buyers may need additional spot volumes and generate increase spot trade. Particularly after the reductions seen in 2015 on the import market. In the longer term outlook, potential decreases in Canadian acid output may also lead to a transition to increased imports to the US Gulf.

Contracts for 2017 in Chile have been a sore point for many suppliers, as negotiations have led to reported settlements as low as the \$20s/t c.fr Mejillones. At the same time, spot prices have firmed into the \$30s/t c.fr in Chile in November/early December. The main issue for Chile going forward is the deterioration of import volumes of acid. Imports in 7M 2016 dropped by over 26%, leading to expectations the deficit for the year will be below 2015, at less than 2 million tonnes. With Chile less and less active in the spot market, there is much focus on trade to Brazil. With the processed phosphates markets being below par in 2016, this has also led to a contraction in acid consumption in Brazil. The country's ongoing economic issues also prevail and raise questions on the potential recovery in the outlook. Imports of acid to Brazil dropped by 18% in January-October, at less than 400,000 tonnes. This is almost even with 2014 import levels during a slow period in the market. Producers in Spain and Mexico continue to rely on Brazil as an outlet for exports, while market share for the European producers has been eroding.

Price indications

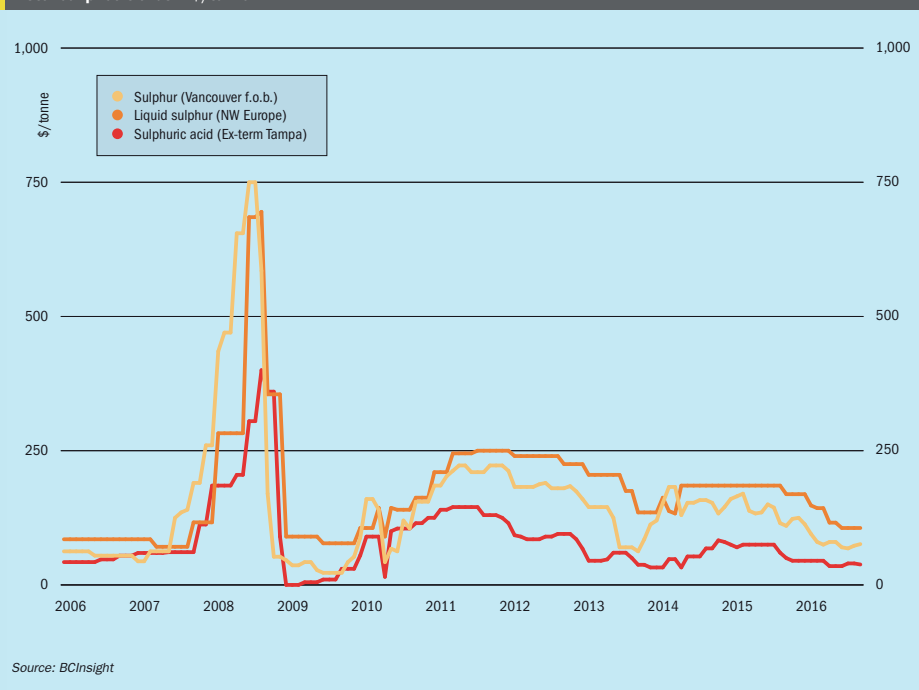
Table 1: Recent sulphur prices, major markets

Cash equivalent	June	July	August	September	October
Sulphur, bulk (\$/t)					
Vancouver f.o.b. spot	80	70	68	73	76
Adnoc monthly contract	80	80	70	77	81
China c.fr spot	78	74	77	83	97
Liquid sulphur (\$/t)					
Tampa f.o.b. contract	70	65	65	65	70
NW Europe c.fr	116	106	106	106	106
Sulphuric acid (\$/t)					
US Gulf spot	35	35	40	40	38

Source: Various

Market outlook

Historical price trends \$/tonne



Source: BCInsight

SULPHUR

- The key question in 2017 on the supply side will be when will we see fresh sulphur volumes from projects in Qatar and Kazakhstan? Both are pegged to come online in 2017, but it remains to be seen how quickly export availability will lead to any downward pressure in the market. Any unforeseen delays could lead to stable and firm pricing for the market.
- China remains a major market influencer. In the outlook it will be crucial to understand how the oil and gas sector is expanding and increasing sulphur recovery in the local market – this may lead to a downturn in import volumes.
- Low oil prices remain a factor for the long term outlook as less investment in new projects has led to an assumption of slower growth prospects for long term sulphur production and a more balanced global market. However, the recent announcement from OPEC to cut

production for the first time in 8 years may aid in oil price recovery.

- In 2016, we have seen a major supply shift in the global sulphur market, as the Middle East has been the leading producing region of sulphur, with North America now second. There is increasing focus on price dynamics in the Middle East region as a result.
- **Outlook:** Sulphur prices may be about to peak in the low \$90s/t f.o.b. Middle East due to the lack of support from the downstream markets, and sluggish prospects in the processed phosphates market for the short term. However, the balanced market on the supply side, and any potential project delays in 2017 could lead to a stable to firm outlook.

SULPHURIC ACID

- Trade volumes to Morocco will be key to balancing the European acid market in the short to medium term outlook. Any fluctuations in requirements could

lead to challenges in placing volumes or lower acid pricing.

- While imports to date have slowed to Brazil, any spot demand over the coming weeks will help support the market, with some requirements for December cargoes aiding European prices. Further ahead, any recovery in the phosphates market will help support renewed import demand in the country.
- The heavy turnaround schedule at smelters in North America next year will likely lead to some regional tightness and provide support to the spot market.
- Metals prices have rebounded recently and may provide some support to the outlook for acid if the run up continues
- **Outlook:** Stability will be the key word in the short term outlook as we look to the new year. The run up in sulphur prices may aid in keeping acid flat. Uncertainty surrounds trade flows from Europe in 2017 with the loss of Cuba as an outlet and the rise of Morocco as a leading importer in the region.

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

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UNITED ARAB EMIRATES

Foster Wheeler to boost Shah output

The Abu Dhabi National Oil Co (Adnoc) has awarded a front-end engineering design (FEED) services contract to UK-based Amec Foster Wheeler to help boost the capacity of its Al Hosn Shah Gas plant in the emirate by 50%. Al Hosn is a 60-40 joint venture between Adnoc and Occidental Petroleum.

Mansoor Al-Mehairbi, Al Hosn Gas CEO said: "The expansion project is an important part of Adnoc's integrated Gas Master Plan, which seeks to increase natural gas supplies to support various industries nationwide. By adding 50% processing capacity, the expansion will help ensure Abu Dhabi's steadily rising demand for gas is met, in a sustainable and economic manner. The project will utilise state-of-the-art technology to enhance the Shah Gas Plant's gas processing capacity, improve efficiency and performance, and increase profitability by optimising the use of existing assets to maximise the return on investment."

Expansion of Shah will increase gas processing capacity from 1.0 billion cubic feet per day to 1.5 billion cubic feet per day, while maintaining the highest degree of plant integrity and HSE requirements. The design concept scheduled to be completed by 4Q 2017, will provide the detailed design information necessary for the final investment decision to be made. Under the terms of the contract, new units will be engineered to expand the plant's processing capacity along with all associated offsites and utilities necessary to integrate the new units with existing installations, including gas gathering facilities, main gas plant, product pipelines and the sulphur granulation plant.

Shah currently produces 500 million cubic feet per day of sales gas, 4,400 t/d of natural gas liquids, 33,000 bbl/d of condensates and around 9,000 t/d of sulphur.

Adnoc's 2030 strategy also calls for it to optimise the volumes of gas used for re-injection and in field power generation, to free up valuable gas for other uses, such as electricity generation and water desalination. The expansion of Al Hosn will make Adnoc one of the world's largest producers of sulphur. Adnoc says that it has plans to maximise the value of its sulphur by working closely with key fertilizer markets, while also supporting the development of a local sulphur products industry, including enhancing its existing ammonia and urea industry with a new generation of "advanced fertilisers".

IRAN

Total first major to sign deal with Iran

French oil major Total has become the first international oil company to sign an integrated petroleum contract (IPC) with the Islamic Republic of Iran, following the easing of sanctions earlier in 2016. The deal covers Phase 11 of the development of the huge South Pars gas field, the largest in the world, with an estimated 33.8 trillion cubic metres of sour gas shared with Qatar, where it is known as the North Field, and 18 billion barrels of condensates. Total will take a majority 50.1% stake in the development, with China National Petroleum Corp. (CNPC) taking 30%, and Iran's Petropars, a subsidiary of the National Iranian Oil Co. (NIOC), holding the remaining 19.9%. The production target for Pars Phase 11 is 57 million m³/day of gas.

Total hope that the deal will also expedite negotiations on the imminent contract award for the huge South Azadegan oil and

gas field, Iran's biggest oil find since the late 1960s. The field has estimated reserves of around 42 billion barrels of oil, of which around 7 billion barrels are deemed recoverable. It was previously run under a buyback contract signed in 2009 with CNPC. The original plans were to develop the field in two phases, the first targeting an output of 320,000 bbl/d of oil and 198,000 cfd of gas, with the second phase targeting production of 600,000 bbl/d of oil and 404,000 cfd of gas. However, by the end of 2014, only 7 of the 185 wells had been drilled by CNPC, leading to the National Iranian Oil Co cancelling the contract. Iran has now signed three heads of agreements with Shell to study South Azadegan using enhance oil recovery (EOR), as well as the Kish field, which as 1.4 trillion cubic meters of sour gas reserves. Recovery rates at Azadegan are only around 6%, but Iran hopes to boost this to 25% using EOR.

It is also believed that Shell is close to signing a deal on another phase of the Pars development.

PAKISTAN

Contract to be awarded on new refinery

Trans-Asia Refinery Ltd, developing a privately owned refinery in Pakistan, is reported to be about to award the engineering, procurement and construction (EPC) contract for the project in December this year, following financial closure. The 100,000 bbl/d refinery will be a relocation of an existing facility from Italy to Karachi, in collaboration with the Emirate of Dubai, with cost put at \$750 million. The refinery is expected to be operational in 2019, according to the Pakistan Ministry of Petroleum and Natural Resources.

It forms part of a government drive to achieve self-sufficiency in refining capacity. Pakistan currently has six operational refineries, following the start-up of the Byco Petroleum Co complex at Hub, Baluchistan in 2015, with a capacity of 120,000 bbl/d. As well as the new capacity under development at Trans-Asia, Pak Arab Refinery Ltd is proceeding with the 300,000 bbl/d Khalifa Coastal Refinery at Khalifa Point, also near Hub, at an estimated cost of \$4.5 billion.

In addition to these expansions, Attock Refinery Ltd is working to enhance its capacity from its existing 43,000 bbl/d to 53,000 bbl/d at its Morgah refinery near Rawalpindi, as part of a \$250 million upgrading project. Recently Attock completed work on a 18 MW captive power plant, provided by MAN Diesel & Turbo SE, and Hyundai Engineering Co. Ltd., which replaces the previous 7.5 MW of power generation capacity. The power plant will run on heavy fuel oil and will accommodate the energy requirements of the new units to be added under the upgrading program, including a 10,400 bbl/d pre-flash unit at an existing crude distillation unit raise overall crude capacity at the site to 53,400 bbl/d; a 7,000 bbl/d naphtha isomerisation unit to lower levels of benzene and aromatics in gasoline; and a 12,500 bbl/d diesel hydrodesulfurisation (DHDs) unit to reduce the sulphur content of diesel production at the plant to below 500 ppm in compliance with Euro II specifications.

UNITED STATES

Lewis Pumps celebrates 125 years

Weir Minerals is proud to announce that 2016 marks the 125th anniversary for one of its long standing brands – Lewis Pumps. Over that time, Weir Minerals Lewis Pumps has developed an international reputation

in the design and manufacture of pumps and valves in the sulphur, sulphuric acid and phosphoric acid industries. Beginning in 1891 as Chas. S Lewis and Co., Inc., a family business that adopted alloy customisation methods to furnish custom OEM pump lines to the beer pasteurisation and bottle cleaning industries, by 1906 the company had begun manufacturing pumps in-house and later developed and manufactured its first sulphuric acid pump in 1914, beginning a specialisation which has continued to the present day. This specialisation in sulphuric acid equipment continued and, in 1975, the company manufactured its first sulphuric acid valve. The company became part of the Weir Group in 1994.

"It is an honour to be celebrating 125 years of developing and manufacturing our Lewis pump range in this ever-changing industry. We have witnessed many challenges and opportunities throughout our history and continue to advance to meet the demands of our markets," said Bob Elliott, divisional director EHS/VCE for Weir Minerals and former managing director of Weir Minerals Lewis Pumps.

Matrix buys Houston Interests

Matrix Service Co. has acquired privately held global solutions company Houston Interests LLC. Houston Interests, based in Tulsa, Oklahoma, provides consulting, engineering, design, construction services and systems integration. The transaction is valued at \$46 million. Houston Interests will operate as part of Matrix subsidiary Matrix PDM Engineering. The Tulsa operations of both companies will be combined into one facility, and will be led by an integrated team comprised of Houston Interests existing management, as well as current Matrix PDM management. Long-term Matrix executive Brad Rinehart will assume the role as president, while Houston Interests' founder Doug Houston will serve in a consulting role, providing strategic and technical expertise as well as ongoing client support.

Matrix PDM Engineering will now be able to offer expertise in new markets such as natural gas processing and sulphur, providing for further growth and diversification of revenue streams as well as expansion of services into international markets.

"The integration of Houston Interests into Matrix PDM Engineering supports our strategy for continued growth and diversification, and provides added bench strength and expertise to support the significant

infrastructure project opportunities we see in front of us," John Hewitt, Matrix Service Co. President and CEO, said in a statement.

FRANCE

Axens improves bid for Heurtey

Axens says that it is increasing the price of its voluntary takeover bid for Heurty Petrochem to a fixed price of €25 per share, in order to strengthen the attractiveness of the bid. The increased price represents a 37.1% premium over the Heruty share price at the time of the original bid in October, and a 44.1% premium over the volume-weighted average price over the previous three months. Axens, part of the Institut Francais du Petrole (IFP) Group, is an international provider of advanced technologies, catalysts, adsorbents and services, focused on the conversion of oil, coal, natural gas and biomass to clean fuels and production and purification of major petrochemical intermediates. Heurty Petrochem is an international oil and gas engineering group covering two market segments: process furnaces for refining, petrochemicals and the production of hydrogen; and the processing of natural gas through its subsidiary Prosernat. In this sector, Heurty operates in both EPC engineering and as a technology licensor.

AUSTRALIA

100th reference for hydrotreating catalyst

Axens has been awarded its 100th reference for its Impulse hydrotreating catalyst for ultra low sulphur diesel (ULSD) service. BP's Kwinana refinery selected the HR 1246 CoMo catalyst following results of pilot plant tests conducted by BP's Refining Technology & Engineering Groups and subsequent tendering process. Thanks to its intrinsic high activity, HR 1246 will be used with regenerated HR 626 and enable a <10 ppm sulphur diesel product.

RUSSIA

New condensate purification plant

Federal agency Glavgosekspertiza Russia has approved the construction of a process condensate purification facility at the Gazprom Neft Omsk refinery. The project is a part of a modernisation programme at the refinery which will include environmental upgrades. The new facility is expected to be commissioned in 2018, and will process

100 m³/hour to remove ammonia and sulphur-containing compounds from process condensate produced at secondary refining units. About 97% of water collected from the separation and purification process will be fed back to the plant's process cycle. The sulphides and ammonia separated in the process will be recycled into sulphur and nitrogen. The facility is being designed by Technological Engineering Holding PETON of Ufa.

KUWAIT

Contract for SRUs awarded

Black & Veatch has secured a contract for the licensing technology and related services for sulphur recovery units (SRUs) and acid gas removal units (AGRUs) to support natural gas processing on the Jurassic field, a project central to Kuwait Oil Company's (KOC) strategy to increase gas production. Kuwait has a very aggressive gas production target for 2030. Developing the Jurassic field is key to attain this goal. Black & Veatch's contribution centres on acid gas removal and sulphur recovery at Jurassic Gas Facility 1, which will produce about half of the Jurassic field's potential.

Thomas Christensen – Black & Veatch's Managing Director, Oil & Gas, Middle East and North Africa – said, "We were able to offer Kuwait Oil Company a total natural gas processing solution, delivered by a single source, bringing together multiple technologies."

The contract includes two acid gas removal trains designed for feed gas flow; and three parallel identical sulphur recovery trains, with dedicated tail gas treatment units for each sulphur train. Black & Veatch is responsible for the basic engineering design package and is bringing together various technologies to meet the performance requirements.

AFRICA

Move to end imports of high sulphur diesel

Nigeria, Benin, Togo, Ghana, and Ivory Coast have jointly agreed to ban imports of high-sulphur diesel fuels as part of an initiative organised by the UN Environment Programme (UNEP). Permitted levels of sulphur in imported diesel will fall from as high as 3,000 parts per million (ppm) in some of the countries to 50 ppm. In Europe the maximum has been 10 ppm since 2009.

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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Sulphuric Acid News

CHILE

Codelco orders two replacement acid plants

SNC-Lavalin says that it has been awarded an engineering, procurement and construction (EPC) contract by Corporación Nacional del Cobre de Chile (Codelco), one of the largest copper producers in the world, for the construction of two sulphuric acid plants at the Chuquicamata Copper Smelter Complex, located in the Antofagasta region of northern Chile. The project will license sulphuric acid technology from MECS Inc., with whom SNC-Lavalin has successfully executed projects for over 50 years. The plants will treat off-gas from the Chuquicamata smelter, each producing up to 2,048 t/d of market grade sulphuric acid. These new plants will replace those currently in operation at the facility and are part of Codelco's ongoing environmental compliance

plans. Construction is expected to begin in early 2017 with SNC-Lavalin performing basic and detailed engineering, procurement of equipment, and construction of the acid plants via their Santiago and Toronto offices.

"Following our recent contract award for the replacement of the effluent treatment plant at the Chuquicamata Copper Smelter, this new contract once again supports our strong position in Latin America", said Jose J. Suárez, President, Mining & Metallurgy. "We are very proud to be part of a local project that, once completed, will not only provide Codelco with first class acid plant facilities, but will be a key element in their environmental program for the future."

EGYPT

Evergrow expands scope of phosphate complex

Belgian phosphate producer EcoPhos says that it has signed a joint venture agreement with Egypt's Evergrow to build a dicalcium phosphate (DCP) plant as part of a 50:50 joint venture. The venture will construct a plant in Sadat City, 95 km from Cairo, to produce 110,000 t/a of DCP for animal feed. The complex will also comprise plants to produce 100,000 t/a low-cadmium fertilizer, 198,000 t/a sulfuric acid and 60,000 t/a calcium chloride. Investment costs are estimated at \$120 million.

EcoPhos will provide the technology, detailed engineering and equipment for the project. The company added that it is also planning further investment in an NPK plant, but gave no further details. The DCP output will be sold and distributed by Aliphos, EcoPhos' feed ingredients division. EcoPhos acquired Aliphos from Tessenderlo of the Netherlands in November 2013, gaining a production site in Rotterdam and sales offices in Germany, Spain and Poland.

EcoPhos is also building a new phosphate plant in Dunkirk, France, which will start production at the end of 2017, and a fourth phosphate plant with a capacity of 200,000 t/a is planned to open in Dahej, India, in late 2018, in a joint venture with Gujarat Narmada Valley Fertilizers & Chemicals. EcoPhos says that other projects are in the pipeline in Brazil, Senegal, Jordan and Morocco, and the company aims to become Europe's leading producer of feed phosphate for livestock, targeting an annual output of 1 million t/a at five plants in five countries by 2020.

Work begins on phosphate complex

A consortium of Egypt-based Construction & Design SAE ('CONSTEC'), Petrojet, and Hassan Allam Holding has reportedly begun work on the Ain Sokhna phosphate complex near Abu Tartour on Egypt's Red Sea coast. The complex, under the ownership of the state-owned El-Nasr Company for Intermediate Chemicals (NCIC), will include nine plants with a production capacity of 1 million tonnes per year of phosphate and compound fertilisers, diammonium phosphate (DAP) and pure and commercial phosphoric acid. Earlier this year Outotec was awarded the license to supply two 1,900 t/d sulphuric acid plants via its Italian partner Intecsa Industrial.

Talaat Fawzy, managing director of CONSTEC, says that civil works have begun on the complex, including site preparation and electrical supply. The project is due for completion in 2018.

UNITED STATES

Chemtrade closes acid plant

Chemtrade says that it is closing its sulphuric acid production plant at Augusta, Georgia. The plant supplied acid to DSM North America's caprolactam manufacturing site, which it was announced earlier in the year is being wound down, closing in November 2016. In an earnings conference call in August CEO Mark Davis said that the Augusta Georgia plants was "essentially a dedicated plant for this customer. Although we are exploring other merchant marketing opportunities and equipment modifications to keep our assets plant viable, we do not expect to recover the same rate of earnings historically generated by this pipeline customer."

According to local press reports, Chemtrade is closing the acid plant in 4Q 2016, with some associated costs, but hopes to re-purpose it for future operations, including water treatment products, which are also produced at the site. About 40 employees will be redeployed to other opportunities at the Augusta facility.

Chemtrade is one of North America's largest suppliers of sulfuric acid, spent acid processing services, inorganic coagulants for water treatment, liquid sulphur dioxide, sodium nitrite, sodium hydrosulphite and phosphorus pentasulphide.

DuPont supplies acid alkylation training simulators

A refinery in North America has commissioned and installed a STRATCO® operator training simulator (OTS) for two sulphuric acid alkylation units. This OTS was specifically developed and fully customised by DuPont Clean Technologies over a seven-month period, encompassing the entire refinery alkylation operation from feed preparation to product storage. DuPont says that using a simulator allows the alkylation plant operator to safely practice plant shut-downs and start-ups, hone responses to potential emergencies and improve troubleshooting skills for instrument and equipment failures in a realistic yet consequence-free environment.

Dave Dericotte, business segment leader, OTS, DuPont Clean Technologies, said: "As we develop and build the operator training simulators to specification, we have the flexibility to incorporate and adapt the elements that our customers want. At this customer's request, for example, we expanded the instructor suite of tools by adding more failure modes and adding the

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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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ability for the instructor to adjust the speed and difficulty of the failure. This helps the instructor keep each training session fresh and forces the students to learn and develop troubleshooting strategies, as opposed to simply memorising and subsequently applying rigid responses that could have disastrous consequences."

Settlement reached over sulphuric acid spill

The US Environmental Protection Agency (EPA) has reached a proposed settlement with Union Pacific Railroad over violations of the Clean Water Act (CWA) at the Herington Kansas Rail yard. As part of the settlement, the company has agreed to pay a civil penalty of \$24,000 and complete a Supplemental Environmental Project valued at \$338,100. In January 2012, two railroad tank cars at the Herington facility collided, leading to the release of 11,000 gallons of sulphuric acid. Some 1,500 gallons were contained in a ditch, but the remainder flowed through an adjoining ditch and discharged into Lime Creek.

In addition to the civil penalty, Union Pacific will be required to complete a Supplemental Environmental Project (SEP) to install earthen berms, flow barriers, and manually-operated drop gates that will seek to minimize runoff and releases from reaching Lime Creek. Union Pacific also paid \$155,300 to the state of Kansas in restitution for damage to the state's natural resources in August 2013.

CANADA

Successful trial for acid recycle plant

NioCorp Developments Ltd. Says that it has successfully completed trial operations of its calcination pilot plant, which has demonstrated a means of recycling sulphuric acid under continuous conditions. The pilot plant was constructed and operated at the SGS facility in Lakefield, Ontario. Solid material generated during the company's acid regeneration pilot plant was fed to a kiln where it was heated to decompose the solids into the gaseous precursors of sulphuric acid. The pilot plant accomplished all three of its objectives, which were to run the calcination unit operation continuously, to generate quantities of calcine product for follow-on characterisation testing, and to provide a characterisation of the gas generated during calcining.

Previous test runs have confirmed the company's scandium, niobium, and titanium production operations under continuous conditions. NioCorp is aiming to develop the Elk Creek deposit in southern Nebraska, USA, one of the largest deposits of niobium in the world, and the tests at SGS form part of the project feasibility study.

"The means of recycling the Elk Creek Project's two main reagents has now been demonstrated through the acid recovery and calcination pilot plants," said Mark A. Smith, executive chairman and CEO of NioCorp. "Recycling reagents within the plant will mean that our Elk Creek operation would be less dependent on external supplies of these materials, and is in keeping with the company's objective of developing an efficient operating plant design for the Elk Creek Feasibility Study."

SAUDI ARABIA

Saudi Arabia inaugurates phosphate complex

Saudi Arabia has inaugurated the \$35 billion Ma'aden mining and minerals processing complex at Ras Al-Khair Industrial City. The mining and minerals processing complex includes a \$5.5 billion phosphate project which is operated by Ma'aden in partnership with petrochemical giant Saudi Basic Industries (Sabic), comprising phosphoric acid and sulphuric acid capacity, an ammonia plant, di-ammonium phosphate capacity and a co-generation and desalination plant. Additional facilities at Ras Al-Khair include a \$10.8 billion aluminium facility operated by Ma'aden and US firm Alcoa.

INDIA

Coromandel seeks approval for phosphate expansion

Coromandel International Ltd is seeking approval from the Andhra Pradesh Pollution Control Board for an increase in phosphoric acid production from 700 t/d to 1,000 t/d at its site in the Visakhapatnam Port Industrial Zone. The \$33 million expansion will also upgrade phosphorus pentoxide and other ancillary facilities at the fertilizer complex at Sriharipuram. The investment aims to reduce emissions of sulphur dioxide and improving the plant performance, including \$3.8 million to be spent on environment management programmes. Once approvals are obtained from the statutory authorities, the project will be completed in 18 months. The fertilizer plant houses rock phosphate storage and

handling, a phosphoric acid and sulphuric acid plant, ammonia storage tanks and NPK fertilizer granulation and bagging plants.

ANGOLA

Minbos to acquire Angolan phosphate partner

Australian-registered phosphate developer Minbos Resources has agreed to buy its Angolan joint venture partner Petril Phosphate, in a deal worth \$20 million. Perth-based Minbos plans to acquire all Petril shares, which will result in shareholders of the two companies owning about 50% of the enlarged group each. Minbos and Petril are joint venture partners in the Cabinda phosphate project in Angola, which is currently part of a bankable feasibility study. The deal will also include Petril's two other phosphate projects in the Zaire province of Angola.

ETHIOPIA

OCF to help build fertiliser plant in Ethiopia

Morocco's Office Cherifien des Phosphates (OCF) has signed a deal with Ethiopia to build a \$3.7 billion fertiliser plant in the country. OCF said the deal would create one of the world's largest fertiliser facilities, 250 km east of the capital Addis Ababa, with an initial capacity of 2.5 million tonnes. According to OCF, the first, \$2.4 billion phase of the Dire Dawa Fertilizer Complex, scheduled for completion in 2022, "will produce fertilizer made from Ethiopian potash and ammonia gas, as well as OCF's phosphoric acid, taking full advantage of both countries' complementary natural resources." The second phase, worth an additional \$1.3 billion, will increase production to 3.8 million t/a by 2025, according to the statement.

Ethiopia's agricultural sector accounts for 45% of the country's GDP and 90% of its exports, but the country currently uses only 57 kg of nutrients per hectare of wheat and 34 kg per hectare of corn, compared to 150kg and 300kg respectively in Europe. The low use is partly because all Ethiopia's fertilizer is imported and the government, which controls much of the economy, has significant shortage of foreign currency.

The investment will also involve developing infrastructure at Djibouti harbour for handling the phosphoric acid, which will be shipped from Morocco. The nitrogen and potash will be supplied by Ethiopia.



Smelting copper in Zambia.

Mustapha El Ouafi, managing director OCP Group, said the facility will allow Ethiopia to be self-sufficient in fertiliser production.

"This joint venture will bring a range of affordable custom fertiliser products that are specifically adapted to the needs of local crops and soils, thereby ensuring a reliable and affordable supply of plant nutrients to Ethiopia's farmers," he said. "This partnership is firmly rooted in a shared vision of Morocco and Ethiopia's leadership that African natural resources should be harnessed to drive Africa's development and common prosperity."

ZAMBIA

Smelters protest over copper import duty

The Zambian Chamber of Mines says that a proposed import duty on copper concentrates will harm smelters in Africa's second-biggest copper producer. Although Zambia mines most of its own copper, the country imports concentrates from the neighbouring Democratic Republic of Congo (DRC) in order to maximise use of its smelter capacity. Zambia has capacity to smelt 3.6 million metric tons of concentrates yearly, at facilities owned by Glencore Plc, First Quantum Minerals Ltd. and Vedanta Resources Plc, amongst others, while mines in the country only produce around 2.9 million tons.

Zambian smelters also need the higher-grade and specific type of concentrates that come from mines in the DRC in order to operate optimally, the chamber said, and warned that if there was insufficient supply of concentrates, finished copper output would be affected, in addition to employment and contributions to government revenues. Zambian finance minister Felix Mutati proposed the import duty to boost government revenues in his budget speech in November. The 7.5% duty is due to take effect on January 1st 2017.

Sulphur 368 | January-February 2017

Namibian Hake Association, the Midwater Trawling Association of Namibia and the company Omualu Fishing. They argue that Namibia Marine Phosphate did not correctly lodge its environmental clearance certificate and that the process is dangerous for the environment.

RUSSIA

KMMC to cut SO₂ emissions

The Kola Mining and Metallurgy Combine (KMMC) has said it plans to reduce annual emissions of sulphur dioxide by nearly half within two years. KMMC, a daughter company of metal giant Norilsk Nickel, emits some 80,000 t/a to atmosphere, a source of contention with neighbouring Norway, but says that it plans to cut this to 44,000 t/a by 2019. The main source of emissions is said to be a briquetting workshop at the Zapolyarny site, and environmental improvements there have cut emissions by 35,000 t/a. Norilsk Nickel is also in the process of cutting SO₂ emissions at its Norilsk site by 75%, according to the company.

KAZAKHSTAN

CAM to acquire Shuak copper plant

Central Asia Metals Ltd (CAM) has signed a framework agreement to acquire an 80% effective interest in the Shuak copper exploration property in northern Kazakhstan. Shuak is a copper and gold exploration project in the Akmol Oblast region of north Kazakhstan, approximately 300 kilometres north of the capital city, Astana, and 40 kilometres north east of the regional centre, Stepnogorsk. The licence area is 197 km² and contains saprolite ores that may be amenable to copper production by leaching and SX-EW processing, as well as copper, molybdenum- and gold-bearing porphyry. Column leach testing of the saprolitic material with dilute sulphuric acid has achieved copper recoveries of over 90%. CAM says that it intends to commence field-based exploration work in Q2 2017.

Business development director Gavin Ferrar said: "We are delighted to have agreed terms to acquire a majority stake in Shuak, which we have identified as an attractive opportunity given its location and economic prospects. Given our success to date at Kounrad, our initial target would be to develop another similar SX-EW operation at Shuak."

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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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People

Israel Chemicals Ltd (ICL) has announced that its board of directors has appointed **Asher Grinbaum** as the company's interim CEO, effective from September 11th. Mr Grinbaum's appointment follows the resignation of Stefan Borgas as CEO and member of ICL's board earlier in the month. Until recently, Mr Grinbaum served as executive vice president and as the chief operating officer (COO) of the company. Mr Grinbaum will act as interim CEO until such time as a permanent CEO is appointed.

Mr Grinbaum, a resident of Israel's Negev region, has worked at ICL for over 40 years, during which time he has held a number of senior management positions. He began his employment as an engineer at ICL's Bromine Compounds unit, and since then he has held a variety of management positions, including CEO of ICL Fertilizers from 2004 to 2007, and, prior thereto, CEO of ICL Industrial Products. He holds a BA in mechanical engineering and an MBA from Ben Gurion University.



Dr Trainham receives his AIChE award.

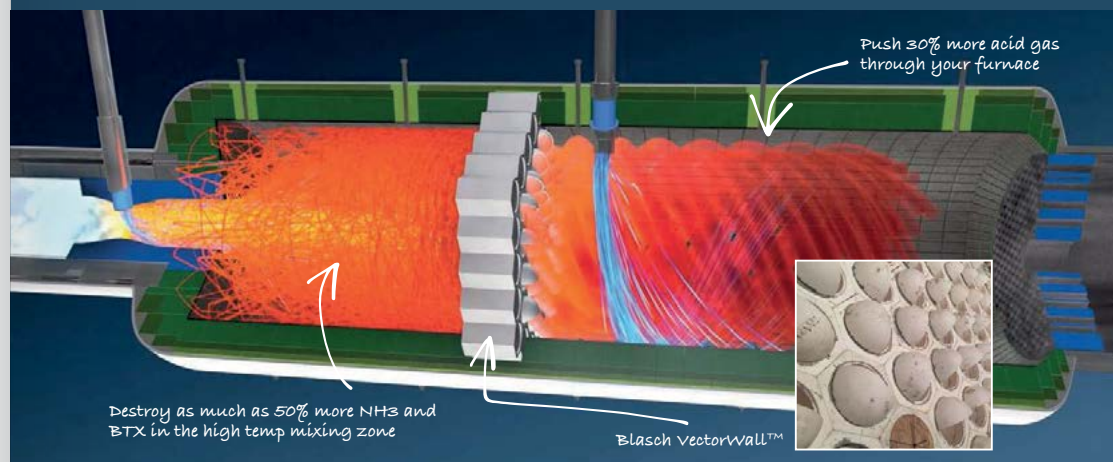
Dr James Trainham, chief technology officer for JDC Phosphates, has received the 2016 Industry Leadership Award from the American Institute of Chemical Engineers (AIChE). The Institute and Board of Directors' Award winners were recognized at the Honours Ceremony at the Insti-

tute's annual meeting in San Francisco on November 14th. AIChE's awards, honour eminent chemical engineers for career accomplishments, service to society, and service to the Institute. More than 8,000 guests attended the awards ceremony.

This award recognises Dr Trainham's impact on the marketplace and global community, the magnitude of the challenges overcome, and the innovation and technical creativity demonstrated throughout his career, including his on-going work with the development and commercialization of JDC's Improved Hard Process for producing phosphoric acid. Dr Trainham has more than 40 patents and publications to his credit and has received multiple awards and honours, including his election to the National Academy of Engineering, recognition by AIChE, with the Award for Chemical Engineering Practice (2002), his selection in 2008 as "one of the 100 chemical engineers of the modern era" and his election as a Fellow of the AIChE in 2012.

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MARCH

12-15

Phosphates 2017,
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9-21

AFPM Annual Meeting,
SAN ANTONIO, Texas, USA
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26-30

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APRIL

3-6

Sulphuric Acid Round Table,
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24-26

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MAY

9-12

SAIMM Sulphur and Sulphuric Acid
Conference 2017,

CAPE TOWN, South Africa
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South African Institute of Mining and
Metallurgy, P.O. Box 61127,
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JUNE

9-10

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OCTOBER

9-11

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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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PHOTO: ESSAR OIL

India's refining industry

India's refining capacity has nearly tripled in the past 15 years, and the country has emerged as a major refining hub, with a knock-on effect on sulphur output.

With the slowdown in China's previously breakneck growth, India has become the focus of world attention, with its GDP growth rate actually accelerating in the past couple of years from 5% in 2013 to 7.6% in 2016, and 2017 is forecast to see 7.8% growth, boosted by low oil prices which have assisted India – a major net importer of oil, and making it the fastest growing of the world's major economies. Although concerns remain about debt, corruption, creaking infrastructure and the fate of the government's reform programme, growth in sectors like IT and media have been major bright spots.

As its economy expands, so India's demand for energy and liquid fuels continues to increase. The country is the third largest energy consumer in the world. Power demand has grown by 11-15% year on year since the turn of the millennium, although the IEA estimated that 19% of the population still lacked access to electricity in 2013. Power generation is mainly via India's huge coal industry, but some 28% of primary energy consumption in 2015 was via petroleum and other liquid fuels, according to BP's Statistical Review of World Energy, and car use continues to expand rapidly, competing with the country's extensive but over-stretched rail network. Demand for energy is

projected to more than double by 2040 as the economy grows to more than five times its current size, according to Indian government projections.

Upstream

India's oil reserves are relatively modest compared to its huge coal reserves, at around 5.7 billion barrels, mostly in the west of the country, although offshore reserves have grown in recent years. This figure is relatively modest compared to major producers, but still around half that of Mexico, say. Domestic oil production is not well developed, however. Indian oil consumption grew 8% in 2015 to reach 4.16 million barrels per day (195 million t/a), taking Indian consumption just slightly ahead of Japan's, and hence making the country the third largest consumer of oil in the world, after the US and China. Figures to Q3 2016 show a rise of 9% year on year, and the government is projecting a growth rate averaging 3.6% out to 2040, taking demand to 460 million tonnes. Set against this, however, domestic oil production has remained essentially unchanged. In 2015 its tood at 876,000 bbl/d, a level it has plateaued at since about 2009. The remaining 3.3 million bbl/d of oil demand must therefore be imported, mostly from

the Middle East, particularly Saudi Arabia, and this figure continues to increase.

Domestic oil production is mainly in the hands of the large state-owned Oil and Natural Gas Corporation (ONGC), which has over 60% of domestic oil production. While there has been some foreign and private investment in exploration and production, India's complicated and bureaucratic regulatory environment has been a significant disincentive, and the country's oil patch, especially offshore, remains relatively under-explored.

There has been investment in more frontier developments and marginal fields to help offset production declines from mature basins. Major discoveries in the Barmer basin in Rajasthan and in the offshore Krishna-Godavari basin hold some potential to diversify the country's production. Meanwhile, dependence on imported oil has led Indian energy companies to diversify their supply sources by taking stakes in producing assets elsewhere. The Indian Oil Corp has bought into oil and gas fields in the US, Canada and Venezuela, for example. Other companies have stakes in Sudan, Russia's Sakhalin project, and ONGC has taken stakes in Azerbaijan and Russia's Vankor field.

Nevertheless, the International Energy Agency (IEA) projects that by 2040 India's

dependence on oil imports will have risen by 90% to 7.2 million bbl/d, putting it second only to China. Some companies have raised concerns about the sustainability of this level of imports, and suggested that India needs to boost its domestic oil production. A recent PricewaterhouseCoopers report says that: "the challenge before Indian companies is to take effective measures for enhancing the exploration and production of petroleum resources. Simultaneously, the infrastructure for refining, distribution and marketing, import, export and conservation of petroleum products must be improved."

The government seems to be aware of this and is taking policy steps designed to encourage more production. It has allowed 100% Foreign Direct Investment (FDI) in many segments of the sector, including natural gas, petroleum products, and refineries, among others, and has allowed state-owned oil firms to evolve their own crude oil import policies, including the freedom to choose the source and price of their crude oil imports, allowing them to fully compete in the market. There are initiatives like the Hydrocarbon Exploration Licensing Policy (HELP), and the government has signalled that it is content to move to a revenue sharing rather than production sharing model for new licensing rounds. The Ministry of Petroleum and Natural Gas has also announced a new 'marginal fields policy', which aims to bring into production 69 marginal oil and gas fields with 89 million tonnes or Rs 75,000 crore (US\$ 11.18 billion) worth of reserves, by offering various incentives to oil and gas explorers such as exemption from payment of oil cess and customs duty on machinery and equipment.

Refining

While oil production has not been a success story for India, refining, conversely, has boomed. India became a net exporter of petroleum products in 2001 and the country's oil imports have largely been as a result of the expansion in the refining sector. Unlike the upstream sector, downstream has seen reforms on pricing and extensive private investment, and a concerted government effort to develop the country's refining sector, particularly under the 11th Five Year Plan (2007-2012). Customs duties on oil imports were eliminated in 2001, reducing the cost of importing foreign oil, and while in the past the government had regulated the fuel price for most petroleum products, the mounting cost of fuel subsidies gradually encouraged

liberalisation, and the government has progressively lift retail price caps on most oil products. Domestic fuel price reforms led to the official deregulation of gasoline prices in 2010 and diesel in 2014.

At the start of 2016, India had 4.6 million t/a of nameplate refining capacity, as shown in Table 1, making it the second largest refiner in Asia, behind only China, and the fourth largest in the world. This figure has nearly quadrupled from a level of 1.2 million bbl/d in 1999. Three large privately-owned refineries provide 37% of capacity, with a number of government-owned companies and some joint ventures the remainder. The Indian Oil Company has 1.6 million bbl/d of capacity, or around 35%, including its subsidiary the Chennai Petroleum Corporation Ltd (CPCL). The state refineries mainly supply the domestic market, while

the larger, private refineries of Reliance and Essar are geared more towards the export market. In particular, sales of low sulphur diesel into Europe and Japan and transport fuels to the Asia-Pacific and even Middle East regions. However, refiners face an ever more competitive environment, as they must compete with new refining capacity in the Middle East and China. This puts a premium on modernising and upgrading existing refining capacity in India. In particular, many of the state-owned refineries are old and badly in need of replacement or refurbishment. The Ministry of Petroleum and Natural Gas is seeking to enhance India's crude oil refining capacity through 2040 by setting up a high-level panel, which will work towards aligning India's energy portfolio with changing trends and transition towards cleaner sources of energy generation.

Table 1: India's refineries

Company	Location	Capacity, '000 bbl/d
Public Sector		
IOC	Barauni, Bihar	120
	Bongaigaon, Assam	47
	Digboi, Assam	13
	Guwahati, Assam	20
	Haldia, West Bengal	151
	Koyali, Gujarat	275
	Mathura, Uttar Pradesh	161
	Panipat, Haryana	301
	Paradip, Odisha	301
HPCL	Mahul, Mumbai	131
	Visakhapatnam, Andhra Pradesh	167
BPC	Mahul, Mumbai	241
	Kochi, Kerala	191
CPCL (IOC)	Manali, Chennai	211
	Nagapattinam, Tamil Nadu	20
Numaligarh	Numaligarh, Assam	60
Mangalore Ref. & Pet.	Mangalore, Karnataka	301
ONGC	Tatipaka, Andhra Pradesh	13
Joint-Ventures		
Bharat-Oman Refinery Ltd.	Bina, Madhya Pradesh	120
HPCL-Mittal.	Bathinda, Punjab	181
Private Sector		
Reliance	Jamnagar	663
	Jamnagar Special Economic Zone	542
Essar Oil	Vadinar, Gujarat	402
Total		4,632
Source: EIA		

Demand

Domestic demand for fuels is dominated by diesel, which represented 40% of all petroleum product demand in 2015, mainly for commercial vehicles, as well as the industrial, power and agricultural sectors. Diesel demand flattened for a year after the lifting of subsidies and decontrol of prices, but has since begun to take off again, rising 10% in 2016. Gasoline demand is also increasing quickly, however, as private cars proliferate. Total fuel consumption is expected to grow around 5-6% in the 2016-17 financial year, but consumption of gasoline is expected to grow around 9-10% over the medium term. The only major oil product sector in which India continues to import is liquefied petroleum gas (LPG). Many rural areas of India use LPG and kerosene, in addition to wood and straw, as cooking fuels. The government is encouraging a shift from kerosene to LPG, as it is cleaner and less expensive fuel, and LPG demand increased by 20% from 2013-15.

While India has been able to use its excess of refining capacity to export, demand for refined products has been increasing rapidly, and is projected to reach 3.0 million t/a in the 2016-17 financial year. Rapid growth in Indian domestic demand means that this could reach 4 million bbl/d in 2020, and it is forecast that – were refining capacity not to rise – the country would become a net importer of refined products some time in the next decade. For that reason, a major programme of refinery expansions is under way.

Capacity increases

India's refining capacity continues to rise, most recently with the start-up of the 300,000 bbl/d IOC refinery at Paradip in Odisha State. The government had actually originally planned that the country's refining capacity should rise to 6.3 million bbl/d by 2017 as part of the current five year plan, but several refinery projects have faced delays in the past few years as a result of financial issues, bad weather, and regulatory hurdles. Nevertheless, the expansions continue. The IOC is aiming to expand its smallest refinery, at Nagapattinam, which is co-owned with Iran, to 300,000 bbl/d in two stages, at a projected cost of \$5.5 billion. The first phase of the project will raise the refinery's output to between 120,000-180,000 bbl/d. IOC is planning to spend \$7.5 billion by 2022 to raise its refining capacity by about 30%



IOC's Mathura refinery.

to 2.1 million bbl/d, including expanding its Panipat refinery in northern India to 400-500,000 bbl/d. It also plans to spend \$600 million on upgrading its new Paradip refinery to new fuel standards via the addition of an isomerisation unit, a diesel hydrotreater and a hydrogen unit.

Hindustan Petroleum says that it will raise its own 300,000 t/a of refining capacity to 1.2 million bbl/d by 2030. Current investments, totalling \$4 billion, including increasing capacity at its Mumbai refinery from 130,000 bbl/d to 190,000 bbl/d by 2019, and increasing capacity at its Vishakapatnam refinery from 166,000 bbl/d to 300,000 bbl/d by 2020. Numaligarh is aiming to treble the capacity of its Numaligarh Refinery in Assam to 180,000 bbl/d, and Bharat Petroleum recently modernised its Kochi refinery, taking capacity from 190,000 bbl/d to 270,000 bbl/d.

On the private side, Reliance is in the process of completing its 'J3' expansion project at Jamnagar. This aims to gasify the petroleum coke output of the refineries, using the resultant syngas for power and synthetic natural gas (SNG) production, and to feed paraxylene and MEG/polyethylene production. Hydrogen will also be exported back to the refineries to help with clean fuel production.

The largest new refinery initiative is a joint venture at Maharashtra announced in December 2016 between three of the state-owned refiners; the Indian Oil Corporation, Hindustan Petroleum Corporation (HPCL) and Bharat Petroleum (BPCL), with a planned capacity of 1.2 million bbl/d, making it the largest refinery complex in the world. IOC will take a reported 40% stake in the new development, with HPCL and BPCL 20% each. The remaining 20% will be held by state constructor Engineers India Ltd. The total cost is put at \$30 billion, including a large associated petrochemical complex. The first phase will include 800,000

bbl/d of refining capacity, together with an aromatic complex, naphtha cracker unit and a polymer complex. The on-stream date for Phase 1 is scheduled for around 2022.

Sulphur output

One of the factors for Indian refiners to juggle with is tightening regulations on sulphur in fuels. India currently has a so-called 'Bharat Stage IV' (equivalent to Euro IV) fuel standard in selected cities of 50 ppm sulphur, which covers about half of all fuel sold in India, but outside these cities the standard is 150ppm sulphur for gasoline and 350 ppm sulphur for diesel. However, the government plans to adopt the equivalent of Euro IV fuel standards on a nationwide basis by April 2017, which will mandate a fuel sulphur level of 50 ppm, and the aim, announced this year, is to move to a Euro V/VI standard on transportation fuels by 2020 (10 ppm sulphur). Indian companies have proposed several expansions to existing facilities and new refineries by 2020 in order to meet these standards. Of course, the export-oriented private refineries which sell in to European and Japanese markets must already meet Euro-V/VI standards for their fuels.

The rapid growth in India's refining sector has seen the country's sulphur production grow from 500,000 t/a in 2005 to 2.2 million t/a in 2015. More is expected from the new refinery expansions, with around 250,000 t/a coming from the Reliance expansion at Jamnagar. Additional sulphur output could reach 1 million t/a by 2020, taking India's total to 3.2 million t/a. Some sulphur is exported from the west coast, but India remains a net sulphur importer, to the tune of around 600,000 t/a. However, new demand from fertilizer and other applications is expected to mean that the new refinery sulphur will not turn India into a net exporter, as demand is also forecast to rise over that period. ■

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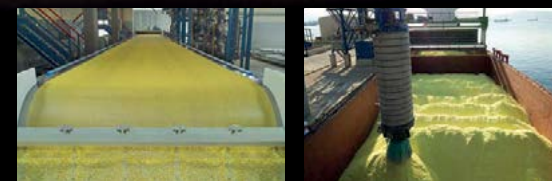
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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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SULPHUR index 2016

A complete listing of all articles and news items that appeared in Sulphur magazine during 2016.



Iran and the Gulf, Mar/Apr p22.

Article	Issue	Pg
ASRL reviews		
Storage of elemental sulphur: will it be necessary? How best to do it?	Sep/Oct	37
The chemistry and technology of sulphur degassing	May/Jun	29
Catalysts		
Sulphuric acid catalysts guidelines	Sep/Oct	56
Sulphur recovery and associated technologies		
Designing for ease of future operations	Jan/Feb	48
Extending the life of sulphur recovery units	Jul/Aug	50
How to measure temperature in the Claus furnace	Nov/Dec	38
Improved acid gas removal	May/Jun	34
Pyrometer verification and calibration methods	Nov/Dec	40
SRE targets continue to rise	Sep/Oct	42
Sulphur-assisted carbon capture and utilisation	Jul/Aug	27
Sulphur pits for durable service	Jul/Aug	44
Sulphur recovery, energy efficiency and carbon management	May/Jun	50
Sulphur recovery from waste metallurgical gases	May/Jun	47
The seven deadly sins of sour water stripping	Mar/Apr	36
Waste heat boiler design, operation and reliability	Nov/Dec	48
Conference/meeting reports		
40th Clearwater Convention	Jul/Aug	32
MESPO 2016	Nov/Dec	28
TSI 2016	May/Jun	26
TSI preview	Mar/Apr	32
SOGAT 2016	May/Jun	22
Sulphur 2016 preview	Sep/Oct	30
Health, Safety and Environment		
Controlling refinery emissions of SO ₂	Jan/Feb	24
Managing emissions in transient conditions	Sep/Oct	62
Phosphates		
Morocco's phosphate boom	Mar/Apr	18

Article	Issue	Pg
Product forming and handling		
US taps into new sulphur sources	Jul/Aug	22
Special supplements		
Sulphur forming project listing 2016	May/Jun	28
Sulphur recovery project listing 2016	Mar/Apr	26
Sulphuric acid project listing 2016	Mar/Apr	34
Sulphur industry/markets		
Acid gas reinjection	May/Jun	24
Alternative energy: the inconvenient reality for sulphur	Nov/Dec	30
Ammonium sulphate production	Nov/Dec	22
Changing sulphur market trends	Jan/Feb	18
Do oil prices affect sulphur production?	Jul/Aug	18
Iran and the Gulf	Mar/Apr	22
New applications for sulphur in polymers	Sep/Oct	26
New sulphur production in China	Sep/Oct	28
Refinery sulphur recovery – an Indian perspective	Jul/Aug	30
The global market for sulphur	Nov/Dec	18
The Shah project	Jan/Feb	32
Sulphuric acid markets		
Copper and sulphuric acid	Jul/Aug	25
Industrial uses of sulphuric acid	May/Jun	16
South Korea's sulphuric acid industry	Sep/Oct	34
Sulphur and sulphuric acid in North America	Nov/Dec	26
Sulphur and sulphuric acid in Southern Africa	Sep/Oct	22
Sulphuric acid technology		
Combining technologies for enhanced acid production	May/Jun	38
Dundee's world class sulphuric acid plant	Sep/Oct	76
Improved heat recovery in sulphuric acid plants	Mar/Apr	50
Improved WSA plant layout for smelter applications	Mar/Apr	58
Monitoring sulphuric acid concentration	Sep/Oct	54
Rapid WHB retubing at IPL Mount Isa site	Nov/Dec	44
Sulphur burning optimisation	Jan/Feb	38
Weighing up the options for acid cooling	Jul/Aug	34

Sulphur industry news

Country	Sulphur industry news	Issue	Pg
Algeria	Contracts awarded for three new refineries	May/Jun	10
Austria	OPEC says no incentive to invest in SO ₂ scrubbing	Jan/Feb	11
Azerbaijan	KT to license SRU to SOCAR SOCAR awards refinery contract	Mar/Apr Jan/Feb	12 12
Canada	Blackbird to process gas from Elmworth CNRL fined for H ₂ S releases Fires shut down oil sands production Last modules shipped for oil sands project New sour gas processing plant Production begins at West Ells	May/Jun Jul/Aug Jul/Aug Nov/Dec Nov/Dec Mar/Apr	11 10 10 12 12 12
China	RioTinto allowed to increase SO ₂ emissions SemCAMS receives environmental approval Trans Mountain pipeline given go-ahead	Jan/Feb Sep/Oct Jul/Aug	12 15 10
China	K+S buys Chinese magnesium sulphate producer Production begins at Luojiazhai Start-up for new hydrotreater	Sep/Oct Jan/Feb Jul/Aug	14 12 12
Colombia	Ecopetrol studies revival of old refinery units	Mar/Apr	12
Germany	Conference urges phased sulphur restrictions Shell and UFT collaborate on sulphur enhanced urea	Jul/Aug May/Jun	10 10
India	Chennai Petroleum upgrade complete by November IOC to expand two refineries Modi inaugurates Paradip refinery Paradip told to cut sulphur emissions Refinery upgrade for HPC-Mittal	Jul/Aug Nov/Dec Mar/Apr Sep/Oct Sep/Oct	12 11 12 14 14
Indonesia	Avens in refinery licenses for Pertamina	Nov/Dec	11
Iran	Drilling to begin at Pars Phase 14 Iran nears deals on smaller oil and gas fields	Jul/Aug Nov/Dec	11 10
Iraq	Mishraq ablaze after recapture from Islamic State	Nov/Dec	10
Kazakhstan	Kashagan operators facing \$740 million fine Kashagan restarts oil production Kazakhstan looks for decision on Tengiz expansion	Jan/Feb Nov/Dec Jul/Aug	11 10 12
Kuwait	SRU contracts awarded	Nov/Dec	10
Malaysia	Enersul to supply S forming units to RAPID Petronas to install DynaWave scrubbers Refinery to install sulphur dioxide scrubbers	Mar/Apr May/Jun Nov/Dec	11 10 12
Mexico	Claus catalyst recovery project First sulphur exports from Pajartos Fluor wins Hidalgo refinery upgrade Government plans \$23 billion refinery upgrades Malaysian firm wins sour gas pipeline contract	Mar/Apr Sep/Oct Jan/Feb Jan/Feb Sep/Oct	12 14 12 12 12
Middle East	Collaboration on heavy fuel upgrading project	May/Jun	10
Morocco	Shell licenses Thiogro technology to OCP	Nov/Dec	11
Netherlands	ExxonMobil expands Rotterdam hydrocracker Fluor completes acquisition of Stork	Jul/Aug Mar/Apr	11 13
Oman	Sulphur facilities part of new port development Work begins on second refinery	Jan/Feb Sep/Oct	11 12
Qatar	Delay for Barzan project	Nov/Dec	10
Russia	License awarded for new Caspian development Lukoil starts up refinery upgrade	Sep/Oct Jul/Aug	15 11



US taps into new sulphur sources, Jul/Aug p22.

Country	Sulphur industry news	Issue	Pg
Saudi Arabia	Contracts signed for Fadhili gas project New control system for desulphurisation plant Rabigh sulphur plant on-stream later this year Start-up for Hasbah Sulphur loading facility contracts awarded	Sep/Oct Jan/Feb Jul/Aug May/Jun Jan/Feb	14 11 12 11 11
UAE	Al Hosn looking at Shah expansion Oxy considering expansion of Shah project Shell bows out of Bab	May/Jun Mar/Apr Mar/Apr	11 12 12
UK	Comments invited in steels for sour service IMO to impose sulphur cap from 2020 Sour gas processing plant cancelled	Mar/Apr Nov/Dec Jul/Aug	13 11 12
US	August start-up for new hydroprocessor Axens to expand its sulphur abatement portfolio Gloomy outlook for US oil and gas companies Honeywell spins off AS business IHS predicts return of risk premium to oil market Ineos buys Calabrian Holdings Marathon faces opposition to SO ₂ emissions New ProTreat version launched Refiners face credit crunch Refinery investigated over SO ₂ release Refinery suffers SRU outages Shell and Aramco agree to split Motiva JV Sulphur recovery market to reach \$2.1 billion TSI launches new sulphur fertilizer website Upgraded sulphur scavenger Worker killed by sulphur pile collapse	Jul/Aug Jan/Feb Mar/Apr Jul/Aug Jul/Aug Sep/Oct Mar/Apr May/Jun May/Jun Mar/Apr Sep/Oct Jul/Aug Jul/Aug Nov/Dec Sep/Oct May/Jun May/Jun Jul/Aug Nov/Dec Sep/Oct Jan/Feb	11 12 11 11 11 12 11 11 11 11 12 11 11 11 11 11 11 11 11 11 12
World	Oil prices have fallen "too far" Slow market impacts refining capacity expansion	Mar/Apr Nov/Dec	11 11

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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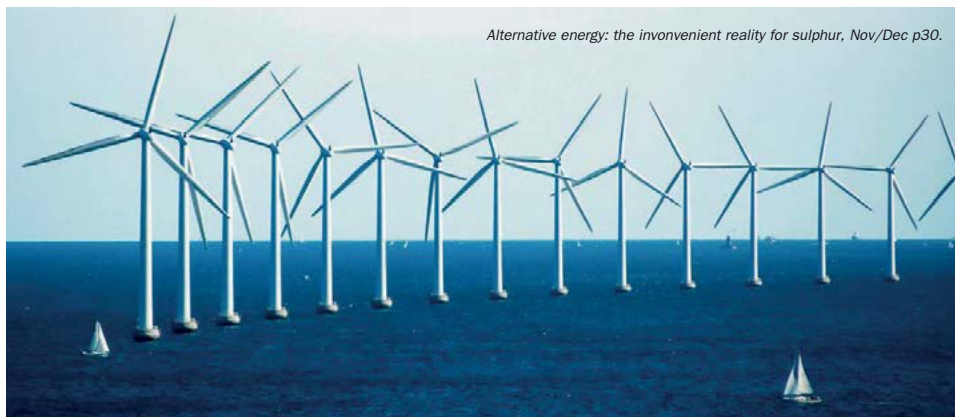
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Sulphuric acid news

Country	Sulphuric acid news	Issue	Pg
Algeria	Indorama to invest in large scale phosphate production	Sep/Oct	16
Australia	BHP looking at faster expansion of Olympic Dam Freight train derailment causes acid spill Funding secured for cobalt project Ifco payment upheld New mine to start up this year	May/Jun Jan/Feb Sep/Oct Jan/Feb Jan/Feb	13 14 19 14 14
Brazil	Anglo American sells phosphate business to China Mosaic interested in Vale fertilizers? South32 may bid for Anglo phosphate business	May/Jun Jul/Aug Jan/Feb	13 15 15
Canada	Acid sales agreement signed Ariane gets go-ahead for Lac a Paul	May/Jun Jan/Feb	13 14
China	Start-up at copper smelter	Jul/Aug	15
Cuba	Start-up for third acid plant	Sep/Oct	16
Dem Rep Congo	Acid plant nears completion	Mar/Apr	15
Egypt	Chinese to invest in phosphoric acid plant EcoPhos announces further international expansion Outotec to supply two acid plants to El Nasr Two more phosphoric acid plants Veolia to supply concentrator to phosphate plant	Sep/Oct Nov/Dec Jul/Aug Sep/Oct Jul/Aug	18 14 14 18 14
Finland	Acid from pulp production Electrochemical waste water treatment system New acid plant for paper mill New acid plant for Harjavalta smelter Outotec awarded acid plant contract from Boliden	Mar/Apr Jul/Aug Jul/Aug Jan/Feb May/Jun	15 14 14 14 12
Germany	Order for acid pumps	Nov/Dec	15
Guinea Bissau	GB Minerals completes test work for phosphate plant	Jan/Feb	15
India	FACT looking to more acid production Food grade phosphoric acid plant New copper smelter for Gujarat OCP in Indian NPK joint venture	Nov/Dec Jul/Aug Jul/Aug Nov/Dec	14 14 14 14
Indonesia	CB&I to license acid alkylation unit New partners sought for Weda Bay nickel project	Sep/Oct Sep/Oct	16 16
Kazakhstan	Copper production ramping up at leach site	May/Jun	12
Lithuania	Lifosa looks towards food grade phosphoric acid	Jul/Aug	15
Macedonia	Outotec supplies SX/EW technology	Nov/Dec	14
Madagascar	Sherritt revises Ambatovy production downwards	Nov/Dec	15

Country	Sulphuric acid news	Issue	Pg
Malawi	License granted for hydrochloric acid process	Sep/Oct	18
Morocco	King Mohammed inaugurates new phosphate plant	Mar/Apr	14
Myanmar	Amnesty seeks relocation of acid plant	Sep/Oct	19
Namibia	Gecko awaits environmental report Official inauguration of new acid plant Weatherly trades acid price cut for Tsumeb option	Mar/Apr May/Jun Jan/Feb	15 12 14
New Zealand	Chatham seeks to recover expenses	Mar/Apr	15
Peru	Another attempt to save La Oroya Exports begin from Las Bambas Study for Bayovar phosphate project Uranium leaching proposal	Jul/Aug Mar/Apr Mar/Apr Mar/Apr	15 14 14 14
Poland	Technip to provide roaster for copper smelter	Mar/Apr	14
Saudi Arabia	Ma'aden to borrow \$1 billion	Jan/Feb	15
Senegal	Another offtake agreement for Baobab Mining begins at Baobab New partnerships in phosphate sector	Sep/Oct May/Jun Mar/Apr	18 13 15
South Africa	Foskor operating at record low levels	Mar/Apr	14
South America	Acid plant revamp	Jul/Aug	15
Switzerland	Phosphorus recycling becomes obligatory	Jan/Feb	15
Togo	Strikes hit phosphate industry	Sep/Oct	16
Tunisia	Plans for new acid, TSP plant Production up, but still at reduced levels Protests halt CPG phosphate production	Jan/Feb May/Jun Sep/Oct	15 12 19
Uganda	Financial closure for phosphate project	Mar/Apr	15
US	Copper SX/EW project posts update Elemental sulphur from leaching Jacobs buys Bayer acid technology Kemper County plant producing acid New sulphuric acid alkylation process Non-acid alkylation route Phosphoric acid leak at New Wales Veolia buys Chemours Sulphur Products division	May/Jun Sep/Oct Sep/Oct Nov/Dec Jan/Feb Nov/Dec Nov/Dec Jul/Aug	12 18 18 14 15 14 14 15
Vietnam	Letter of intent for rare earth leaching project	May/Jun	13
Zambia	Pressure to re-start nickel mine	Mar/Apr	15
Zimbabwe	Acid plant "remains on the table"	Mar/Apr	14

Alternative energy: the inconvenient reality for sulphur, Nov/Dec p30.



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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
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JANUARY-FEBRUARY 2017

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Sulphur recovery plant

Sulphur's annual survey of recent current and future sulphur recovery unit construction projects maps the developing shape of brimstone production from fuel and gas processing plants worldwide.

Operating company	Operating site	Process type	Total new capacity t/d	Licensor(s)	Lead contractor	Project type	Start date
ARGENTINA							
AXION Energy	Campana	Claus, TGT	2 x 30	Prosernat	Prosernat	New	2017
AZERBAIJAN							
SOCAR	Heydar Aliyev	Claus	n.a.	KT Kinetics Tech.	n.a.	New	n.a.
SOCAR	Garadagh	H ₂ S, CO ₂ , amine	n.a.	Amec Foster Wheeler	Amec Foster Wheeler	New	2018
BAHRAIN							
Bapco	Sitra	Claus, NH ₃ , amine, SWS, AGRU	3 x 250	WorleyParsons	n.a.	New	2017
BELGIUM							
ExxonMobil	Antwerp Refinery	O ₂ enrich, amine TGT	325	WorleyParsons	Amec Foster Wheeler	Revamp	2017
BRAZIL							
Petrobras	Premium I	SuperClaus	2 x 240	Jacobs	n.a.	New	2017
Petrobras	Premium II	SuperClaus	240	Jacobs	n.a.	New	2017
Petrobras	Maranhao Premium 1	2 x Claus, NH ₃ H ₂ , amine TGT	238	Amec Foster Wheeler	n.a.	New	2017
CAMEROON							
SoNaRa	Limbe	SRU, SWS	17	Amec Foster Wheeler	KT Kinetics Tech.	New	2017
CHILE							
ENAP	Aconcagua	EuroClaus	45	Jacobs	n.a.	Revamp	2016
CHINA							
CNPC/Chevron	Chuadongbei	Claus, SCOT	2 x 687	WorleyParsons	n.a.	New	2016
Guo Tai	Ordos	O ₂ , H ₂ , TGT, D'GAASS	10	Fluor	n.a.	New	2016
Inner Mongolia	Ordos	SRU	51	Jacobs	n.a.	New	2017
Manshi							
Jiutai Energy	Linyi, Shangdong	EuroClaus	32	Jacobs	n.a.	New	2017
Liaoyang Ref	Liaoyang	NH ₃ , H ₂ , TGT, D'GAASS	140	Fluor	n.a.	New	2016
Sinopec	Fujian	SuperClaus	513	Jacobs	n.a.	New	2018
Shaanxi Yancheng	Yulin, Shaanxi	EuroClaus	41	Jacobs	n.a.	New	2018
COLOMBIA							
EcoPetro	Barrancabermeja	Claus, NH ₃ , amine	2 x 130	WorleyParsons	n.a.	New	On hold
CROATIA							
INA	Rijeka	Claus	n.a.	WorleyParsons	n.a.	New	2018
EGYPT							
MIDOR	Alexandria	Claus	410	n.a.	n.a.	New	2018
FRANCE							
Total	Donges	Claus	n.a.	n.a.	n.a.	Revamp	2017
Total	Normandy	SuperClaus	96	Jacobs	n.a.	Revamp	2019
GABON							
Samsung C&T	Refinery of Gabon	Claus	50	Prosernat	Samsung C&T	New	2016
Samsung C&T	Cap Lopez	TGT	50	Prosernat	Samsung C&T	New	2016

Operating company	Operating site	Process type	Total new capacity t/d	Licensor(s)	Lead contractor	Project type	Start date
GERMANY							
OMV	Berghausen Refinery	O ₂ enrich	2 x 24	WorleyParsons	n.a.	Revamp	2017
GREECE							
Hellenic Petr.	Thessaloniki Refinery	Claus, O ₂	80	Siirtec Nigi	n.a.	Revamp	2016
INDIA							
Essar Oil	Vadinar	Claus, SCOT	675	Jacobs	n.a.	New	2019
HMEL	Bathinda Refinery	Claus, TGT	2 x 750	Prosernat	n.a.	New	2017
Reliance	Jamnagar	O ₂ , NH ₃ , amine TGT	4 x 1,300	WorleyParsons	n.a.	New	2017
INDONESIA							
PT Medco E&P	East Aceh	EuroClaus	48	Jacobs	n.a.	New	2019
Pertamina	Balongan	Claus, NH ₃ , H ₂ , amine TGT	1,100	Amec Foster Wheeler	n.a.	New	n.a.
IRAQ							
Turkish Pet Int	Mansuriyah	Claus, amine	230	WorleyParsons	n.a.	New	2017
ISRAEL							
Bazan	Haifa Refinery	O ₂ enrich	3 x 140	WorleyParsons	n.a.	Revamp	2018
KAZAKHSTAN							
Agip KCO	Kashagan	Claus, TGT	2 x 1,900	WorleyParsons	Black & Veatch, Petrofac	New	2017
Pavlodar Oil Chem	Pavlodar Refinery	Claus, TGT	180 + 260	Siirtec Nigi	Rominserv, Technip	New	2017
KUWAIT							
Chevron	Wafra	Claus, amine	2 x 218	WorleyParsons	n.a.	New	2018
KNPC	Al Zour Refinery	Claus	1,500	Amec Foster Wheeler	n.a.	New	2019
KOC	JPF	Claus, TGT	2 x 100	Siirtec Nigi	Schlumberger	New	2018
KOC	JPF	SmartSulf	2 x 100	Prosernat	Spteco	New	2018
MALAYSIA							
Petronas	Johor	SuperClaus	3 x 470	Jacobs	n.a.	New	2019
MEXICO							
PEMEX	Duba	SRU	n.a.	Amec Foster Wheeler	n.a.	New	n.a.
PEMEX	Cadareyta	SmartSulf, NH ₃	132	WorleyParsons	n.a.	New	2017
NIGERIA							
Dangote Oil	Lekki Refinery	SuperClaus	2 x 115	Jacobs	n.a.	New	2017
OMAN							
PDO	Yibal Khuff Sudair	Claus, TGT	250	WorleyParsons	n.a.	New	2016
PERU							
Repsol	La Pampilla	2 x Claus, NH ₃ , O ₂ , H ₂ , amine TGT	83	Amec Foster Wheeler	SAINC	New	2016
POLAND							
Grupa Lotos	Gdansk Refinery	O ₂ enrich	2 x 72	WorleyParson	Tecnimont	Revamp	2018
QATAR							
Qatar Petroleum	Mesaieed	Sour gas, AGE, Claus, TGT	310	Worley Parsons, Prosernat	Petrofac, Black&Veatch	Revamp	2016

KEY

BTX = BTX destruction
 NH₃ = Ammonia destruction
 SWS = Sour water strip
 Fuel = Fuel gas supplemental burning

H₂ = Hydrogenation
 TGT = Tail gas treatment unit
 O₂ = Oxygen enrichment
 SRU = Sulphur recovery unit
 n.a. = Information not available

Operating company	Operating site	Process type	Total new capacity t/d	Licensor	Lead contractor	Project type	Start date
RUSSIA							
Angarsk Petchem	Angarsk	Claus, NH ₃ , amine	2 x 57	WorleyParsons	n.a.	New	2017
Bashneft	Ufa	Amine, SWS	n.a.	Amec Foster Wheeler	n.a.	New	2018
Bashneft	Ufa	SmartSulf	115	Prosemat	n.a.	New	2018
Gazpromneft	Moscow	LPG treat, amine	n.a.	Amec Foster Wheeler	Amec Foster Wheeler	New	2020
Rosneft	Novokubishevsk	Claus, NH ₃ , TGT	2 x 192	WorleyParsons	n.a.	New	2017
Lukoil	Volgograd	NH ₃ , H ₂ , amine TGT, D'GAASS	2 x 76	Fluor	n.a.	New	2017
Lukoil	Perm Neftegas	Claus	14	Prosemat	n.a.	New	2016
Mariisky	Mari El Republic	SRU, TGT, amine	n.a.	Shell	Amec Foster Wheeler	New	n.a.
Orsknefteorg	Orsk	EuroClaus	2 x 99	Jacobs	n.a.	New	2017
Varino Refinery	Varino	SmartSulf	15	WorleyParsons	n.a.	New	2017
TAIF-NK	Nizhnkamsk	Claus, TGT	2 x 390	Prosemat	n.a.	New	2016
SAUDI ARABIA							
PetroRabigh	Rabigh	EuroClaus	292	Jacobs	n.a.	New	2019
SERBIA							
NIS	Pancevo Refinery	Claus, NH ₃ , amine	170	WorleyParsons	n.a.	Revamp	2017
SPAIN							
Cepsa	Algeciras	Claus, SCOT	280	Jacobs	n.a.	New	2019
Petronor	Muskiz	EuroClaus	86	Jacobs	n.a.	Revamp	2018
THAILAND							
Thai Oil	Sriracha Refinery	Claus, NH ₃ , Flexsorb	2 x 837	WorleyParsons	n.a.	New	2021
TURKEY							
STRAS	Aliaga/Izmir	SRU, TGT, amine, SWS	463	KT Kinetics Tech.	Amec Foster Wheeler	New	2017
Turkish Petroleum	Mansuriya	Claus, amine	230	WorleyParsons	n.a.	New	2017
Tupras	Izmir	Degassing	73	Jacobs	n.a.	New	2017
TURKMENISTAN							
Turkmenbashi Oil	Turkmenbashi City	SuperClaus	25	Jacobs	Hyundai	New	2019
UNITED ARAB EMIRATES							
IPIC	Fujairah	SRU, SWS, amine TGT	330	Amec Foster Wheeler	n.a.	New	2018
UNITED STATES							
Chevron	Richmond, VA	O ₂ enrich	580	WorleyParsons	n.a.	Revamp	2018
Hydrogen Energy California	Kern County, CA	O ₂ enrich, NH ₃ , H ₂ , amine TGT, D'GAASS	100	Fluor	n.a.	New	n.a.
Jupiter Sulphur	Nillings, MN	D'GAASS	135	Fluor	n.a.	New	2016
Sinclair Oil	Sinclair, WY	Claus	n.a.	Amec Foster Wheeler	Amec Foster Wheeler	Revamp	2016
UZBEKISTAN							
Lukoil	Bukhara, Karasul	SuperClaus, TGT	2 x 405	Jacobs	n.a.	New	2018
Mubarek	Mubarek Gas Plant	Claus, amine	1,000	WorleyParsons	n.a.	New	2018
VENEZUELA							
PDVSA	Ei Palito	SRU, amine TGT, SWS	250	Shell	Amec Foster Wheeler	New	2018
PDVSA	Monagas	Amine reg, SWS	54	Amec Foster Wheeler	Amec Foster Wheeler	New	2016
PDVSA	Puerto La Cruz	Claus, NH ₃ , amine	2 x 225	WorleyParsons	n.a.	New	2017
VIETNAM							
Bin Son Refinery	Dung Quat	Claus, SCOT	2 x 105	Jacobs	n.a.	New	2019

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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
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SULPHUR

2016

A review of the 32nd Sulphur Conference and Exhibition, held at the Hilton Metropole in London, UK in November.

For the first time in many years CRU's Sulphur conference returned to its home base of London, after the deteriorating political situation in Turkey necessitated a move from the originally planned Istanbul. CRU's Mike Gallagher, manager of the fertilizer business unit welcomed delegates to the conference, which this year had been preceded by a number of workshops hosted by Optimised Gas Treating, SNC Lavalin, MECS/DuPont and Haldor Topsoe.

Market papers

Lahan Mahadeva of CRU gave a general economic outlook. The bright spots, he said, were Vietnam and Indonesia, where growth is around 6% year on year, while India is moving back towards its trend growth of 6%. Russia has recovered to 1.5-2% growth on a slightly stronger oil price, and Brazil has recovered from its recent crisis towards 1.25% growth next year. In the US, unemployment is at low levels and there has been real wage growth, although business confidence remains low. Overall growth is set to be 1.5% this year and 2.3% next, but the debt overhang, strong dollar and weakness in investment and construction continues to hold the economy back. China has achieved 6.7% on a government stimulus package, but this cannot continue, and growth is forecast to slip to 6.2% next year. As the economy rebalances from years of overinvestment, there is a worry of a repeat of Japan's 'hard landing' in the 1990s. Commodity prices seem to be bottoming out, and are forecast to begin growing in 2017, with oil trending towards \$70/bbl.

On the oil and gas front, James McCullagh of Energy Aspects said that the current market weakness is almost entirely



Pigeons on the Millennium Bridge, St. Paul's cathedral in the background, London.

down to the boom in US shale oil production which has added 6 million bbl/d over the past 5-6 years – something that no one really saw coming. OPEC meanwhile is still largely failing to agree production cuts, and Libyan and Nigerian production is returning. Demand is improving from India but stocks are high and only slowly being drawn down, and rebalancing the market will take time unless OPEC can agree serious output cuts. More production is ahead from the restart of Kashagan although there is lower sour output from Venezuela, Mexico and Colombia. Refineries meanwhile continue to have to reach lower sul-

phur levels. Margins are low for refiners, and while there are more capacity additions in India and especially China, there are likely to be more Atlantic Basin closures and capacity cuts, especially those overburdened with heavy fuel oil and high sulphur fuel oil production. Some refineries in Europe are converting to 'bio-refineries', including Venice and Total in France.

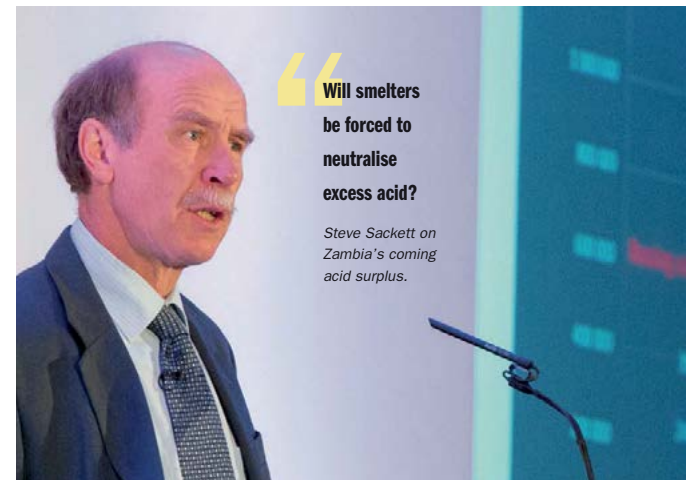
Sulphur and acid markets

Peter Harrison of CRU looked at sulphur markets, which remain volatile, he said, ranging from nearly \$200/t in early 2015

to below \$70/t in 3Q 2016. Global sulphur trade continues to change, with imports growing in most importing regions and huge growth in exports from the UAE as the Shah project begins full production. Saudi Arabia has also increased exports in 2016 but the new Ma'aden plant at Ras al Khair will see this fall in 2017. Canada and Kazakhstan have seen lower exports and less stock drawdown, and sour gas production in Canada is still declining. Over the next few years, the largest supply increases will be from the CIS, Middle East and Asia, although not as fast as was once anticipated, as some refinery expansions have been pushed back due to lower oil prices. Furthermore, a lot of supply growth is in remote regions, such as Norilsk in Siberia or the Caspian Sea, from Kashagan, Tengiz and Astrakhan, where there may well be stock build. Demand growth to 2021 will rise from 61.4 million t/a to 71.7 million t/a, mainly from phosphoric acid production in Morocco, Saudi Arabia, and to a lesser extent China and India. All of this leads to a peak in sulphur surplus in 2018 of around 1.6 million t/a, and stock building may exceed the surplus by 2019, implying that 2016-17 may see the lowest prices, with some improvement thereafter.

On the sulphuric acid side, CRU's Brendan Daly said that many major markets had seen demand weakness in 2016, including Chile, the US and Brazil, but other markets like India, Morocco, China and SE Asia been able to take up the slack. Morocco in particular has the flexibility, via its own acid production, to switch between sulphur imports and acid imports according to prevailing prices. Looking further forward, smelter acid supply is rising by around 8 million t/a over the period 2015-20, more than half of that in China. Demand growth to 2020 is forecast to be around 1.6 million tonnes for phosphate production and 10 million t/a for industrial uses, taking total acid demand to 293 million t/a. More acid for leaching will be required in the DRC, Zambia and Mexico, but use in Chile and the US will decrease – overall about 2 million t/a extra will go to metal leaching. On a regional basis, the US market is relatively in balance, but there is a shift in end use towards ammonium sulphate, chemical production and acid alkylation. Europe meanwhile faces a future without Cuban demand to export to.

Steve Sackett presented an update on his look at Southern African sulphuric acid markets from 2015. The regional sur-



Will smelters be forced to neutralise excess acid?

Steve Sackett on Zambia's coming acid surplus.

plus he predicted had not materialised, he said, in part due to regional drought which had less to less demand for fertilizer and electricity outages as hydroelectric dams were at low levels, and in part due to low copper prices impacting on producers, coupled with breakdowns and project delays. The same amount of acid was produced in 2016 as 2014 in spite of capacity increases. Namibia's coming surplus has now been pushed out to 2019. Meanwhile, the excess acid capacity from smelters in Zambia is still coming, possibly displacing sulphur burning acid across the border in DRC – although the situation is complicated by power generation that comes from sulphur burning. Will smelters be forced to neutralise excess acid?

Other markets

Chris Lawson of CRU gave the phosphate outlook paper, commenting that it had been a painful year for phosphate producers, with a 'tidal wave' of supply emanating from the Middle East and North Africa. DAP prices have dropped from \$500/t f.o.b. Tampa in 2014-15 to \$330/t. There has been some producer response, such as a fall in Chinese exports, but Chinese production hasn't fallen much while consumption there remains relatively flat and likely to fall as the government moves to cut over-application of fertilizers. Stocks are building in China and some rationalisation among the smaller producers seems likely. But just that morning, he said, larger

phosphate producers in China had agreed to cut production – by 30% for those producing more than 2 million t/a, by 20% for those producing more than 1 million t/a, and by 10% for those less than 1 million t/a. It remained to be seen how widely this was adhered to.

Robin Behr of Societe Generale looked at base metal markets. Base metals prices peaked in February 2011, and bottomed out in January 2016. Since then zinc has moved up 50% and nickel 35%. Lead has seen a 10% rise, but copper prices remained flat. There are still some fears of a Chinese 'hard landing', and the credit boost has not boosted Chinese investment. Copper still faces strong supply growth carried over from the previous boom years. But nickel is moving from oversupply into deficit thanks to the Indonesian ore export ban and Philippines moves on mining companies. There has been a low rate of metal discoveries and exploration in recent years, probably leading to market deficits in the 5-10 year time horizon. Even copper faces a supply gap from around 2019, and the industry faces increasing difficulties due to environmental and community challenges, scarcity of water supplies and declining ore grades. A price of \$7,000/t is needed to incentivise new production, and current prices are only around \$5,000/t.

Finally, Robin Macdonald of Wood Mackenzie considered caprolactam markets. Caprolactam is used in film, fibre, filament and engineering plastics, with demand

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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growth at around 2.5% per year, dropping to 2.2% recently as China slows. China will represent 50% of the market by 2020, and represents all incremental growth over the past 25 years, as plastics manufacture has moved to Asia. Sulphur consumption by nylon manufacture has fallen with time on a tonne for tonne basis due to improvements in process technology, and sulphur consumption is highest in legacy plants in Europe and North America, so the switch to China has not brought as much new sulphur demand, and indeed as caprolactam plants have closed in Europe and America, so sulphur demand has even fallen slightly. China has hugely overbuilt capacity, with 13 projects over the 2016-18 period which will add 3 million t/a to a 5 million t/a market, but most are integrated with wither fibre or benzene producers and many larger are trying to force other suppliers out of business within China in a zero sum game, as most of this capacity is not internationally competitive. The net result is that China will become self-sufficient but possibly not impact too much on the rest of the world, aside from some inevitable rationalisation in e.g. Europe.

An unusual topic for a Sulphur conference was covered by Chris Hankin of Imperial College who is also the Director of the UK government's Research Into Trustworthy Industrial Control Systems (RITICS) programme. In spite of the spread of digitisation and convergence between domestic/office IT systems and industrial control systems (ICS), he said, major differences still remain. ICS tend to be operating continuously in real time with time critical functions and high throughput, and involve complex interactions with physical processes. There are also legacy issues – ICS systems can be in place for 15-20 years of operation, and access to components can be difficult. Nevertheless, cyberattacks on industrial control systems are increasing. The US recorded 295 attacks in 2015, mainly 'spear phishing', and Dr Hankin looked at strategies for protection of key control systems from cyber attack.

Peter Clark looked at the future of sulphur in a world where global oil and gas production was decreasing – perhaps as little as 10-20 years ahead of us – where the sulphur might come from. The ability of Frasch mining to take up the slack is highly

dubious, he said – at its peak sulphur mining had never produced more than 15 million t/a and many of the larger deposits were exhausted – certainly not enough to replace 60 million t/a or more of fossil fuel derived sulphur. More on this subject can be found in his article in the previous issue of Sulphur, November/December 2016, on pages 30-37.

Christian Ryming of Bery Maritime gave an update on the dry bulk shipping market. The story is still one of overcapacity. While sulphur represents only 0.5% of the dry bulk market, most is comprised of oil and coal, mainly heading to China and the rest of Asia, with China representing 75% of growth in tonne-miles since 2002. But the decline in Chinese industrial expansion has taken year on year growth of 5.5% in the dry bulk market to 2.6% over the next couple of years. Meanwhile there is 55 million deadweight tonnes of overcapacity, or up to 90 million dwt if operators stopped their current policy of 'slow steaming'. New ship orders have slowed tremendously, and supply growth is down to 1% over the next two years, with the market moving back towards balance in 2016 and thereafter possibly freight rates rising slightly. However, Christian cautioned; "ship owners are greedy and have short memories."

Sulphur recovery units

Beginning the technical sessions, a trio of papers considered ways of improving sulphur plant efficiency. Elmo Nasato looked at improving plant operating conditions to boost performance, including the often overlooked utility streams. Angie Slavens next focused on the benefits of energy recovery technologies which allow waste heat from the exothermic Claus reaction to provide high or low pressure steam, using a case study based on a large sour gas plant. Frederic Tonnaire of Aspen Technology considered ways in which simulation can help with optimising operations, by incorporating Sulsim by Sulphur Experts into the Aspen HYSYS program, allowing simulation of an entire plant from hydro-treaters through acid gas recovery, sulphur recovery and tail gas treating unit.

On a similar theme, KT Kinetics Technology presented their AA-SRU supervisory analytics digital platform, which aims to allow operators to detect the need for

maintenance intervention to avoid upsets and minimise operating costs.

Other SRU papers included Philip Le Grange of Sulphur Experts, who considered why amine systems fail in sour service – this paper can be found in full in this issue on pages 42-51. Simon Weiland of Optimised Gas Treating discussed the influence of reactor configuration and operation on the effectiveness of ammonia destruction in a Claus furnace. A different use for ammonia, meanwhile, was presented by Peter Lu of Jiangnan Environmental Technology, looking at using it to treat SRU tail gas in a coal gasification-based plant, producing ammonium sulphate. Miike Shields and Sheldon McKee of AMGAS showcased some advances in their H₂S scavenging chemicals for treating sour crude oils and process waters, and Mark Duisters of Jacobs Comprimo Sulphur Solutions explained the benefits of modular construction of SRUs. The session ended with a Sulphur Recovery Troubleshooting Clinic moderated by Elmo Nasato.

Sulphur handling

Moving downstream from the SRU, the session on sulphur forming began with a paper by Franco Chakkalakal and Mike Allenspach of Pentair Thermal Management in conjunction with Kent Kalar of Topside Solutions, looking at re-melting of sulphur in a sulphur pipeline. They recommended fibreoptic temperature measurement be incorporated early in the pipeline design, allowing a predictive model to be constructed from data collected which can assist during the difficult task of re-melting.

Mark Gilbreath of Devco gave an update on the new sulphur melter at Mosaic's New Wales plant, which we discussed in detail in our July/August 2016 issue, and he was preceded by Raiz Basheeruddin of Furnace Fabrica, who discussed OCP's 6,000 t/d sulphur melt plant which was completed in 2015 at Jorf Lasfar.

With the prospect of more sulphur going to long-term storage over the next few years, Paul Davis of ASRL reviewed the results of ASRL's multi-year study into long term storage of sulphur, both above ground and below. The key in the latter case is to make sure the sulphur stays above the water table, but above ground storage is better. Using an acid neutralising material such as limestone as a protective cover on sufficient thickness to prevent the temperature fluctuations that cause stress crack-

Lifetime Achievement Award

Peter Clark, retiring from his positions as Director of Research at Alberta Sulphur Research and Professor of Chemistry at the University of Calgary, was given a lifetime achievement award for his 32 years of work in the sulphur industry, most of it with ASRL.

At a subsequent evening event Paul Davis of ASRL and Dave Sikorski of HEC Canada took Peter on a humorous journey through his life, from his time in Hull University in the UK and his support of Wolverhampton Wanderers through marriage, children, relocation to Canada and his time with ASRL, punctuated with some suitably embarrassing video clips!



ing of the sulphur is also recommended.

Sulphur dust is a well known fire and explosion hazard, but Gerry d'Aquin was also concerned with what he called volatile sulphur particulate – tiny jagged sub-micron particles of sulphur which are potentially similar to coal, silica or asbestos dust in their effect on lungs.

Sulphuric acid plants

The sulphuric acid sessions began with a look at some recent projects. Dundee Precious Metals' new Tsumeb metallurgical acid plant was featured in our previous issue. Also showcased was the Kansanshi acid plant in Zambia, here presented by Bodrick Mumba of FQM, and the 2,200 t/d Toros Tarim acid plant in Turkey, in a joint paper between licensor MECS and designer Desmet Ballestra. A case study was also presented by the Indian Farmers Fertilizer Cooperative (IFFCO) concerning improvements to their acid plant at Paradeep, including a new sulphur filter pre-coating system and steam flushing of the secondary sulphur filters, as well as a new clean sulphur storage tank, all of which lowered ash content of the sulphur from 40ppm to 10ppm.

Sulphuric acid technology

A trio of papers considered the reliability of operation of acid plants. An Guo of Wylton International presented a spray repair tech-

nology allowing repair of sulphur furnaces or waste heat boilers which can be applied during brief periods of downtime without stopping production for the 1-2 months that it can sometimes take to make good equipment failures. Chetan Chothani of Breen Energy Systems highlighted the use of an acid dewpoint monitor downstream of the electrostatic precipitator (ESP) in a flash smelter, which allows the operator to optimise air injection so as to maintain the balance between adequate SO₃ for ESP performance while minimising downstream corrosion. Stephan Jakob of Schaaf Machinery & Services discussed replacement of a main gas blower which avoided previous problems with vibration and bearing failures caused by dust as well as increasing performance across the whole operating range. NORAM Engineering and Construction presented their approach to sulphuric acid plant modernisation projects, both in strategy and technology terms.

Heat recovery is a critical part of any acid plant. Ahma Yazdi and Axel Schulze of Hugo Peterson presented their company's Multi-Bundle Radial Flow Heat Exchanger, designed to be more compact and have a lower pressure drop than a conventional radial heat exchanger. Outotec, together with Spanish smelter operator Altantic Copper, meanwhile, gave details about the new heat recovery system Outotec has installed at the Huelva smelter, which

has increased the smelter's efficiency of operation. In order to achieve up to 100% energy recovery from an acid plant, Chemetics has developed a family of process systems, which were described by Rene Dijkstra, including case studies of the CES-Alpha process to produce low pressure steam and the CES-DSW process to produce desalinated water. thyssenkrupp Industrial Solutions also showcased their own sulphuric acid heat recovery system for recovering low grade heat.

Catalyst developments included a discussion by DuPont MECS as to how using differential scanning calorimetry (DSC) had assisted with development of a new caesium catalyst formulation. Marten Granroth and Marie Vognsen of Haldor Topsoe considered operating strategies to optimise acid plant operation, including a dust protection catalyst for lower pressure drop and longer campaign length, a dust protection catalyst for lower power consumption, reducing start-up time through the use of a caesium catalyst, and a higher activity catalyst for lower scrubber operating cost. Finally, Jacob Hjerrold Zeuthen of Haldor Topsoe introduced a new catalyst – Sulphur Monolith Catalyst (SMC), which catalyses H₂S oxidation in Claus tail gas, achieving almost complete conversion (H₂S slip is typically below 3ppm) with negligible oxidation to SO₃. The SO₂ can then be removed by wet scrubbing or other cleanup technologies.

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
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Acid dew point measurement and control

Acid dew point measurement of sulphur condensables is desirable for controlling critical process parameters as well as plant safety in processes where SO₃ gas is present. In this article **C. Chothani** of Breen Energy Solutions discusses the application of acid dew point measurement technology to a copper smelter in Spain as well as the need for a new sensor for use in sulphuric acid plants.

Acid dew point measurement in a copper smelter

Copper is a ubiquitous metal used in wide ranging industries and is a critical part of our daily lives. The process of copper extraction from its ores consists of grinding, roasting, smelting and ultimately electrolysis.

The process for extracting metals from sulphurous ores was developed by Outokumpu in Finland and first applied at the Harjavalta plant in 1949 for smelting

copper ore¹. In this process, the sulphur ore is injected along with air into the furnace and the smelting reactions take place, mostly in-flight, before the particles settle in a bath at the bottom of the furnace where the copper matte forms the lower layer and the slag forms the upper layer. Sulphur dioxide (SO₂) is a byproduct of this reaction, which is generally captured further downstream and sent to a sulphuric acid plant for the production of strong acid.

In this process, some of the SO₂ further oxidises to SO₃ to form a weak sulphuric

(H₂SO₄) acid in the gas stream. SO₃ and weak acid production is generally a nuisance as it must be removed and treated, which is costly. However, the SO₃ is necessary to maintain the efficiency of the hot electro-static precipitator (HEP) to capture fine particulate. Therefore controlling SO₃ to a minimum required level while minimising the formation of additional weak acid, is a necessary and valuable part of process improvement in a flash smelting furnace.

Atlantic Copper in Huelva, Spain has made minimisation of weak acid production an essential part of the company's continuous improvement policy. It was determined that, at Atlantic Copper, it was necessary to reduce the amount of weak acid generated for two main reasons: to reduce the costs of corrosion in the gas system processing equipment and to minimise the operating costs of the gypsum plant².

As part of its continuous improvement process, Atlantic Copper identified the various factors that impact weak acid production. These include: concentrate blend and distribution and air distribution at the burner, fuel burners and draft control, control and optimisation of the sulphation air and gas temperature. Measurement and control of sulphuric acid dew point in the gas stream at the exit of the waste heat boiler by manipulating the sulphation air was determined as a direct and critical closed loop control process in the minimisation of weak acid production. This article will discuss the implementation of this process control technique using a new and improved acid dew point measurement technology.

The process

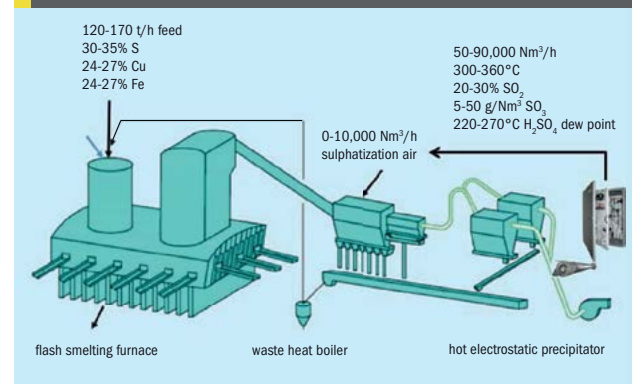
Consider the block diagram in Fig. 1 of the Huelva flash smelting furnace and the waste heat boiler.

Flue gas from the flash smelting furnace first goes through a waste heat boiler where heat is recovered from the flue gas before it passes through the hot electrostatic precipitator where particulate matter is removed from the flue gas. The acid dew point is measured after the HEP and maintained at a level that ensures efficient PM capture while minimising weak acid production.

Acid dew point measurement

Acid dew point measurement (ADM) technology has been available in the industry for over 50 years. The technique employed for ADM is very simple, namely a glass sensor

Fig 1: Flash smelting furnace block diagram



CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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Fig 2: Breen Energy Solutions kinetic dew point measurement algorithm

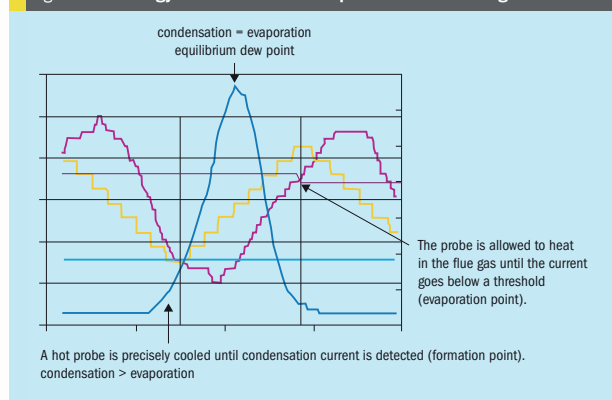
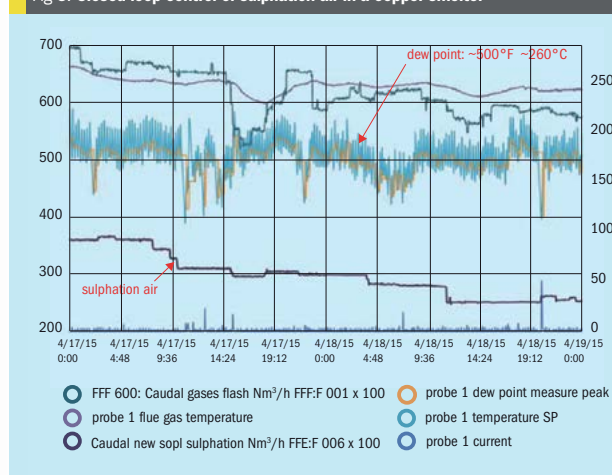


Fig 3: Closed loop control of sulphation air in a copper smelter



with two embedded electrodes is introduced into the gas stream containing acid vapour. The temperature of this surface is then controlled so as to allow acid to condense on the surface. When a liquid substrate is formed on the sensor, a current is measured across the electrodes and the temperature of the sensor is simultaneously recorded as the acid dew point. Prior technologies have been based on maintaining an "equilibrium" substrate on the sensor surface by using the current level as a surrogate and manipulating the surface temperature to maintain this equilibrium. This is sometimes referred to as the equilibrium process or algorithm

for the measurement of acid dew point. While this algorithm works well in relatively clean flue gases, when sulphates or other contaminants are present in the flue gas, particularly those that can foul the surface, alter the substrate chemistry or the electrical characteristics of the substrate, the long-term validity of the measurement is suspect and in most cases contributes to process drift or loss of measurement.

Breen Energy introduced a new ADM technology in the fossil-fired power industry about 12 years ago that was based on a similar current-temperature glass sensor, but used a kinetic algorithm.

In the kinetic algorithm (Fig. 2), the sensor temperature starts at a value above the dew point and is then cooled at a controlled rate, until something condenses. The temperature is then allowed to heat back up, during which time the measured current initially rises and then eventually falls, until it goes back to zero indicating the liquid substrate is now evaporated off the sensor surface. At this time, heated air is blown across the surface to remove any accumulated debris such as sulphates and the cycle starts all over again. Within this cycle, the temperature at which acid first condenses is measured as a Formation temperature, the temperature at which the current peaks is measured as the dew point and the temperature at which the acid fully evaporates is measured as evaporation temperature. The absolute value of the current is not relevant, which makes the algorithm immune to changes in surface chemistry and conductivity. And the material is fully evaporated and then cleaned with air, which prevents long-term fouling of the sensor surface.

Data and results

The dew point as measured by the Breen AbSensor device was reported as 260°C and was in line with process expectations (Fig. 3). It should be noted that prior dew point instruments used in this location for this application, based on the equilibrium algorithm, had reported a dew point of 175°C, which is much lower than process conditions would dictate. The control loop was placed in closed-loop control mode to maintain dew point by adjusting sulphation air at the WHB with success. A permanent installation of the system has been made and has been successfully operational for close to a year (Fig. 4).

Acid dew point measurement in sulphuric acid production

In sulphuric acid production, sulphur dioxide (SO₂) is oxidised over a solid vanadium catalyst to generate sulphur trioxide. The sulphur trioxide is then absorbed in water contained in 98% sulphuric acid to form new sulphuric acid. The SO₂ is either produced by burning elemental sulphur or is an off-gas from other processing plants such as a copper smelter. The gas laden with SO₃ in the process is kept moisture free and the only time moisture gets introduced in the gas is when there is a process upset.

Fig 4: System installation at Atlantic Copper, Huelva, Spain



This is a highly undesirable condition and can be potentially hazardous.

Moisture in process gas laden with SO₃ is reflected in the acid gas dew point and is generally indicative of one of the following conditions, among others:

- drying tower malfunction;
- moisture in feed;
- waste heat boiler tube leaks;
- economiser tube leaks;
- cleaning system malfunctions.

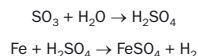
Acid dew point measurement, in this process, to a large extent, is mostly for moisture leak detection. The presence of moisture and therefore increased acid vapour, as evidenced by increased process gas dew point, in the gas stream is detrimental to equipment health and can cause extreme corrosion.

A secondary impact of moisture in the flue gas can be the formation of hydrogen gas creating an explosion hazard³.

Fig 6: AbSensor-ADM-SA installation



Fig 5: AbSensor-ADM-SA for sulphuric acid manufacturing plants



The hydrogen thus formed as shown in the equations above, can create an explosion hazard in the presence of oxygen and an ignition source.

Acid dew point measurement challenges in sulphuric acid plants

There are several challenges that must be overcome in order to successfully develop an instrument that can survive in a sulphuric acid plant and provide the necessary measurement. Chief among them are:

- completely sealed design to isolate process gases from the system internals;
- must withstand a process pressure of up to 20 psig (typical 6-10 psig);
- heated tip to allow operation in process temperatures close to the dew point;
- safety features to isolate the system in case of sensor failure/breach;
- material of construction designed to handle high concentrations of sulphuric acid;
- process mounting system designed to seal and allow online system removal for maintenance.

The Breen AbSensor-ADM-SA (Fig. 5) was specifically developed for this challenging environment and is currently undergoing

round 2 of beta testing at a sulphuric acid manufacturing plant (Fig. 6).

Beta testing round 1, Sept 2016

The following goals were established for the first round of beta testing:

- Can the system handle the challenging process conditions (positive pressure, temperature, acid loading)?
- Can the system measure a consistent dew point without fouling?
- Can the system respond to a simulated moisture ingress event?
- Can the system be installed and uninstalled while the plant is in operation?

The system performed well under these conditions, measuring a dew point of 240 ± 10°F. Additionally, when soot blowing air, laden with moisture, was passed across the measurement surface, the sensor responded with a current spike thereby simulating a response to a moisture leak event.

The system has since been removed for examination and implementation of new design elements to allow it to operate in moisture leak failure detection mode. In this mode, the system would typically operate at higher temperatures where acid dew point is expected in case of a moisture leak. Once a day, the system would lower its operating temperature to ensure system health. The second round of beta testing started in early November, 2016, and is expected to continue through the first half of 2017.

Acknowledgement

Many thanks to Atlantic Copper, and specifically Mr. Francisco Jimenez and Mr. Guillermo Rios Ransanz. Many thanks also to the informal Hydrogen Safety Workgroup comprising various vendors and consultants from the industry.

References

1. Outokumpo flash smelting process; Outokumpo.
2. Jiménez F., Ramos M., Pérez I. (Atlantic Copper): "A review of recent improvements to control weak acid production at Huelva smelter"; IFSC 13th (Outotec).
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Sulphur management for an unconventional gas-condensate field

Nowadays, emission regulations dictate deep removal for all sulphur components in natural gas processing plants. **M. Jacques** of Technip and **A. König-Adolph** of ENGIE E&P review the main outcomes of a conceptual study conducted by Technip for ENGIE E&P Deutschland GmbH for a gas-condensate field development located in Germany, processing raw gas with a low content of H₂S (<0.2 vol-%), moderate H₂S to CO₂ ratio (2.5-3 volume basis) and low H₂S to organic sulphur compounds ratio (5 volume basis).

Fig 1: Impact level of project orientation during the course of a project

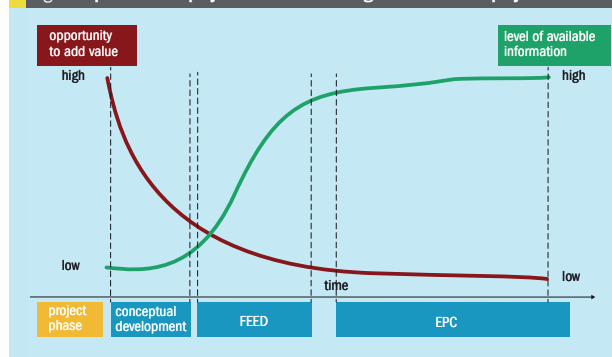
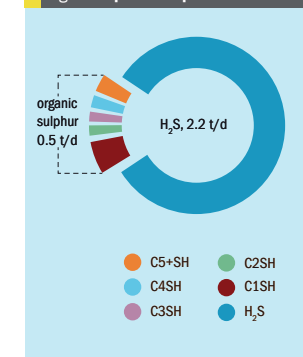


Fig 2: Sulphur compounds



Concept definition/pre-FEED is known to be crucial for successful project development (Fig. 1). The best time to influence the cost and profitability of projects is during the early phase. Shortening the concept phase may result in significant cost and schedule overruns when the project goes to detail engineering phase.

Pre-FEED adds a further level of detail to the shortlisted options identified in the basic phase to achieve more precise definition, a better cost estimate and, ultimately, selection of a single option for FEED.

If these two important steps are well managed, the project development can enter FEED with a single, well defined option. The development's major design issues have been decided; risks and uncertainties are

understood; the cost estimate, budgets and schedules are clear and complete.

As part of the pre-FEED activities, technologies selection and integration are important because:

- Early identification of the potential issues allows proactive management of the interfaces and faster optimisation of the overall project.
- Early integration of the technologies is essential for smooth management of the interfaces and compliance to the project constraints and is the key to avoid re-work at a later stage.

Technology selection and integration provides the basis for clear instructions to licensors for the development of their PDP,

which is paramount for an efficient way forward. The following aspects, in particular, should be clearly defined:

- firm up the basis of design;
- ensure homogeneity in design (design margins, turn-down, material selection);
- compliance with site-related constraints (layout, construction, environment regulations).

These are critical to avoid re-work at the engineering phase, and to increase the net present value (NPV) of the project by making sure the right line-up is selected.

The pre-FEED study presented in this article is a perfect example of the importance of allowing sufficient time during process selection to reach these objectives.

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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Table 1: Sulphur related product specifications

Product	Specification	Maximum content
Sales gas	H ₂ S + COS content	5 mg/Nm ³
	Total sulphur	6 mg/Nm ³
Propane or butane	H ₂ S	1 mg/kg
	Elemental sulphur	1.5 mg/kg
	COS + elemental sulphur	5 mg/kg
	Total sulphur	20 mg/kg

Table 3: Sulphur recovery system requirement

Specification	Minimum	Maximum
Sulphur recovery efficiency (1)	99.5%	N/A
H ₂ S in stack gas	N/A	10 mg/Nm ³
COS + CS ₂ (as S) in stack gas	N/A	3 mg/Nm ³

(1) Sulphur recovery efficiency is calculated over the whole treatment chain (including SRU and TGTU) as the fraction of sulphur in the feed that is recovered in the sulphur stream routed to the collection pits. When the applied technique does not include a recovery of sulphur (e.g. seawater scrubber), it refers to the sulphur removal efficiency, as the % of sulphur removed by the whole treatment chain.

Fig 3: Märkisch Buchholz CPF basic scheme

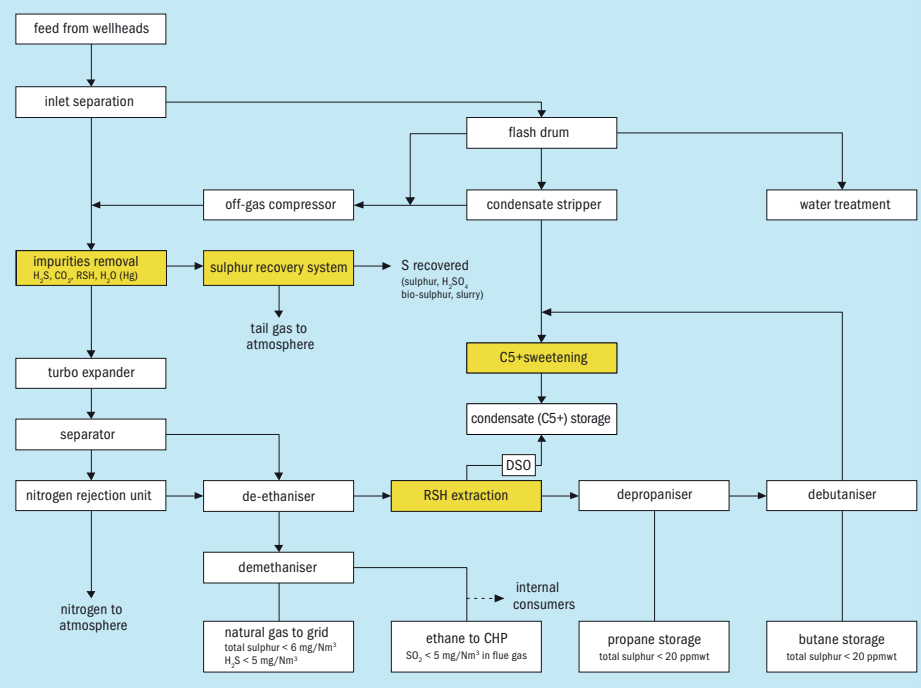


Table 2: Emission limits for combustion installation for generation of electricity and steam

Specification	Maximum
SOx (as SO ₂)	5 mg/Nm ³

Project background

ENGIE E&P Deutschland GmbH is developing the Märkisch Buchholz gas-condensate field. The development focuses on the reasonable and nearly complete material and energetic utilisation of the content of the reservoir.

Märkisch Buchholz gas-condensate field is peculiar with a nitrogen content of about 90 vol-% and the presence of other impurities: carbon dioxide (CO₂), hydrogen sulphide (H₂S), mercaptans (RSH) and not proven mercury (Hg). Another particularity of the Märkisch Buchholz gas-condensate field is the low H₂S to RSH ratio of 5 on a volume basis, whilst the total sulphur content in the feed is about 2.7 t/d.

Table 4: Pre-selection – desulphurisation options matrix

Options	A1	A2	A3	B1	B2	B3	C	D1	D2	E
High pressure gas desulphurisation	X	X	X							
Chemical solvent based acid gas removal										
Hybrid solvent based acid gas removal				X	X	X				
Silica gel (RSH, C6+)								X	X	
H ₂ S scavenger										X
H ₂ S scavenger NG product			X			X				
Molecular sieves (CO ₂ , H ₂ S, RSH, H ₂ O)							X			
Molecular sieves (CO ₂ , H ₂ S, H ₂ O)								X	X	
Molecular sieves (H ₂ S, RSH, H ₂ O)	X	X		X	X					
Molecular sieves (CO ₂ , RSH, H ₂ O)										X
Molecular sieves (H ₂ O)			X			X				
LPG DeRSH			X			X				
Butane alkali wash	X	X		X	X		X	X	X	X
Sulphur recovery										
Claus sulphur recovery	X			X						
Caustic scrubber		X								
Flue gas wet scrubber	X			X						
SO ₂ regenerative absorption							X	X		
Sulphuric acid		X	X		X	X	X	X		X
Biological or liquid redox sulphur recovery									X	
DSO Processing		X	X		X	X	X	X		X

Considering the above and the stringent product specifications and emission limits detailed in Tables 1-3, the Märkisch-Buchholz is a very interesting and challenging case, calling for the consideration of different sulphur management options compared to the ones generally applied in a gas processing plant.

Finally, the overall sulphur recovery shall comply with the European directive 2014/738/EU and TA-Luft emission regulation given in Table 3.

As illustrated in Fig. 3, the basic scheme of the Märkisch Buchholz Central Processing Facilities (CPF) consists of inlet facilities, condensate stabilisation and sweetening, acid gas and sulphur components removal from the gas followed by cryogenic nitrogen removal. The produced hydrocarbons are then fractionated into natural gas (mainly C1), ethane, LPG (C3 and C4 cuts) and hydrocarbon condensate (C5+).

Before entering the FEED phase, ENGIE E&P Deutschland GmbH instructed Technip to carry out the pre-FEED activities in order to finalise the concept phase. The main task of the pre-FEED was to define

the most valuable integrated concept for the surface facilities and to provide the basis of design and standards to be used for the FEED phase.

Prior to commencement of the core pre-FEED work, a review and assessment of different process alternatives for the removal of acid gas and sulphur compounds was performed. This work, defined as the pre-selection study for gas desulphurisation schemes, aimed to select two relevant desulphurisation schemes. These two schemes have then been thoroughly compared considering several key criteria such as the opex, capex, NPV but also public relations, permitting, environmental impact and operational robustness, taking into account technology suppliers and/or licensors information.

Technologies mapping

In the pre-selection study, a total of five basic schemes were identified and named option A to option E. For options A, B and D, sub-cases were studied in order to assess the optimum downstream treatment:

- Option A: Chemical solvent wash
- Option B: Hybrid solvent wash
- Option C: Purification on molecular sieves
- Option D: Purification on silica gel + molecular sieves
- Option E: H₂S scavenger + molecular sieves

A detailed line-up of the technologies involved for each of the options is given in Table 4.

The pre-selection study demonstrated that schemes A3 and D2 were the most promising considering capex, opex, robustness and operability. Option D2 was also attractive thanks to the marketability of the bio-sulphur within the Märkisch Buchholz region. However, option D2 was finally discarded due to the difficulty to assess peak mercaptan concentration in the spent regeneration gas and associated impact on the microorganism activity of the biological unit. The option D2 process line-up was therefore modified in order to minimise the amount of mercaptans making its way to the biological sulphur recovery unit, leading to option D3 as shown in Fig. 5.

It should be emphasised that although liquid redox sulphur recovery processes, (that could be considered in option D2) appear attractive for such sulphur capacity from capex and opex point of view, they were discarded in this particular case. Indeed, the low hydrogen sulphide to mercaptan ratio in the feed gas would lead to a significant amount of disulphide present in the sulphur formed that could result in operational issues requiring frequent shutdowns to clean out the unit.

In option A3, illustrated in Fig. 4, the H₂S, CO₂ and a portion of the mercaptans are removed by a chemical solvent in the AGRU. The acid gas is then routed to a wet gas sulphuric acid unit where all the sulphur compounds are converted into sulphuric acid. The treated gas from the AGRU, which is water saturated, is first cooled down in order to remove the bulk of water before being dehydrated on molecular sieves. Since only a fraction of the mercaptans are

removed in the AGRU, and since the molecular sieves grade is selected to absorb only water (to prevent having to manage mercaptan content diluted in the regeneration gas), the entire LPG cut, where most of the mercaptan will concentrate, is processed in an alkali wash unit in order to remove the RSH down to the required specification. Disulphide oil (DSO) produced in the LPG alkali wash unit is then sent to the wet gas sulphuric acid unit thus enabling it to be

recovered as marketable product. Finally, in order to satisfy the H₂S specification of 5 mg/Nm³, the sales gas is polished on a H₂S scavenger solid bed.

In option D3, the gas is first fed to a mercaptan adsorption unit where RSH (C1 to C5SH) are removed down to circa 5 ppmv, water is removed down to circa 10 ppmv, and heavier hydrocarbons (C7+) down to 1 ppmv. The adsorbent (specially formulated silica gel) is regenerated using feed gas (avoiding the need of a recycle gas compressor). Sour heavy hydrocarbons recovered from cooled regeneration gas are diverted to the inlet facilities.

The H₂S and CO₂ still contained in the treated gas exiting the mercaptan adsorption unit are removed in the acid gas removal unit to meet 1 ppmv H₂S and 50 ppmv CO₂ specifications. The treated gas from the AGRU, which is water saturated, is first cooled down in order to remove the bulk of water before being dehydrated on molecular sieves. The butane cut is processed in an alkali wash unit, the produced DSO is mixed with the sweet hydrocarbon condensate recovered from debutaniser bottom. As in option A3, the

sales gas is finally polished on a H₂S scavenger solid bed.

The acid gas from the AGRU is processed in a low pressure biological unit where the H₂S is converted into bio-sulphur. The treated gas from the bio-scrubber is incinerated before being released to atmosphere.

Capex and opex comparison

A ±40% estimation of capex has been performed for the two selected cases. Fig. 6 shows a normalised comparison of the two schemes, case D3 being used as the 100% reference.

The comparison of the opex including variable cost (utilities and consumables), fixed cost (direct operation cost, non-routine work, insurance and contingencies), maintenance cost and general expense cost (safety and environmental protection and administrative costs) has also been performed and is summarised in Fig. 7, the case D3 still being the 100% reference. Both for capex and opex, option A3 has a slight advantage over option D3 that lies within the range of the estimation uncertainty.

Final Assessment

Although option A3 capex and opex are marginally lower than option D3, other criteria are in favour of this latter option:

- lower environmental impact: SO₂ released to atmosphere is half of the one of option A3;
- ease of marketing bio-sulphur compared to sulphuric acid;
- better operability, e.g. foaming risk in the AGRU is greatly reduced thanks to the nearly complete removal of C6+ and BTEX in the upstream mercaptan removal unit;
- smaller footprint;
- lower preventive maintenance effort;
- handling of solid, non-dangerous sulphur versus sulphuric acid.

Taking these aspects into account, and particularly the lower environmental impact, potential permitting issues and public relation aspects, ENGIE finally selected option D3 for further development of the Märkisch Buchholz gas-condensate field.

Fig 4: Scheme A3 simplified block flow diagram

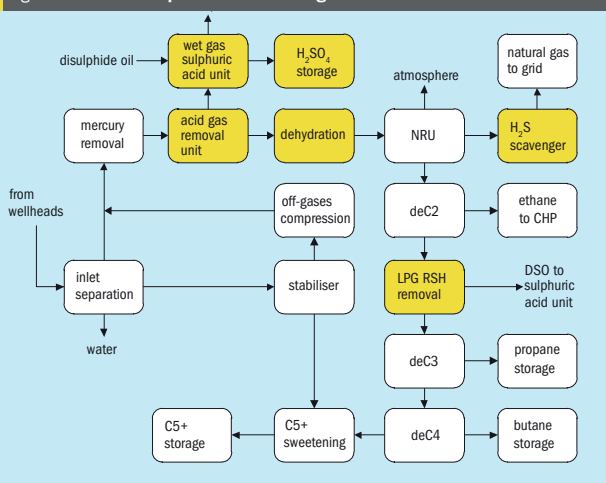


Fig 5: Scheme D3 simplified block flow diagram

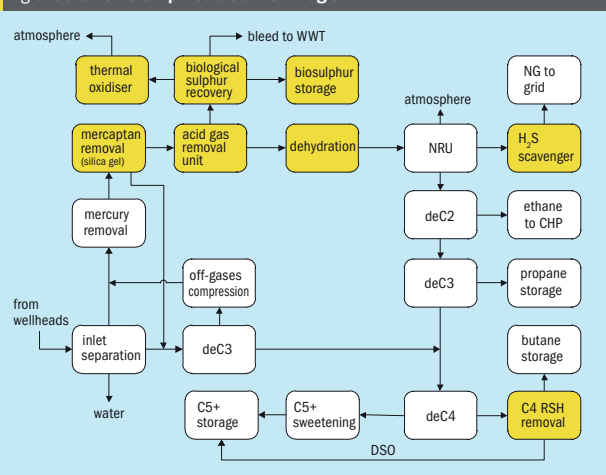


Fig 6: Case A3 & D3 relative capex

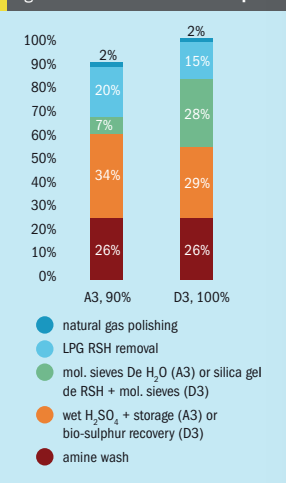
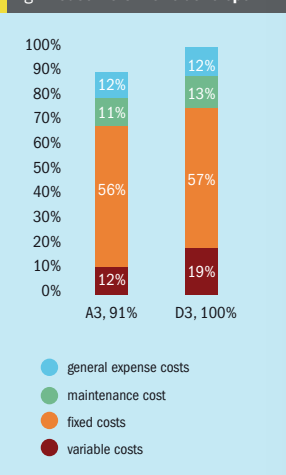


Fig 7: Case A3 & D3 relative opex



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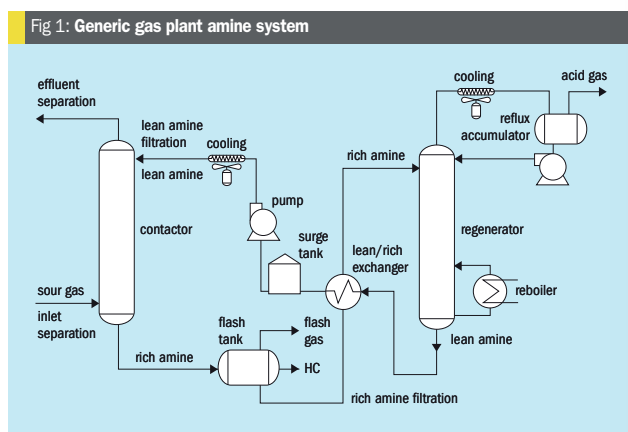
Why amine systems fail in sour service

Amine systems failures have cost the industry billions of dollars over the last 30 years. In this article, **P. le Grange**, **M. Sheilan** and **B. Spooner** of Amine Experts draw attention to the highest probability threats to a facility's operability and reliability. The most prevalent causes of failure in sour amine systems are determined and strategies to prevent these are given.

Equipment malfunction or unplanned shutdown of an amine system can have a devastating effect on a production company's profitability and an equally serious impact on the environment. The goal of this paper is to determine, and focus, the industry's attention on the highest probability threats to their facility's operability and reliability. The threats were identified by analysing hundreds of cases of amine system failure investigated by Amine Experts and others industry leaders over the last 18 years. These include detailed root-cause analysis related to corrosion, foaming, hydraulic restrictions and incorrect specification that have limited or crippled facilities the world over.

To the authors' knowledge, this is the first time that such a database has been compiled and published specifically focused on amine systems in sour plants. While the root causes of these failures are not necessarily 'new' knowledge to the industry, the frequency and likelihood of specific root causes occurring relative to other common root causes has not previously been qualified to this level of detail. For the purposes of this article a sour amine system has been defined as containing more than 1,000 ppm of H₂S in its feed gas. This incorporates refinery amine systems (including tail gas units), many gas plants and amine systems used in the steel mill coke oven sections. Amine systems that do not meet this criteria, e.g. those used in ammonia production and CO₂ capture will not be analysed in this article.

A generic amine system is shown in Fig. 1. Amine systems work by absorbing the hydrogen sulphide (H₂S) and carbon dioxide (CO₂) from hydrocarbon streams into the circulating liquid amine solvent



solution. This is done in a contactor/absorber column and allows for the hydrocarbon product to be further processed, transported or sold without the corrosive potential and dangers associated with H₂S and CO₂. The amine solution is then regenerated in the regenerator/still/stripper column by dropping pressure and the addition of heat, which releases the H₂S and CO₂ from the amine solution so that it may be reused. The H₂S and CO₂ may be further processed, reinjected or incinerated.

Amine solutions typically comprise predominantly an amine and water. The amines most commonly used are methyl-diethanolamine (MDEA), 2-(2-Aminoethoxy)ethanol (DGA™), diisopropanolamine (DIPA), diethanolamine (DEA) and monoethanolamine (MEA). MDEA is the most prevalent in the gas industry at the time of publication. Other components

often added to amine solutions include phosphoric acid (for improved regeneration of the solvent), piperazine (for increased CO₂ removal in MDEA), sulfolane (for COS and mercaptans removal) and anti-foam (for improved operational performance).

Failure types

Of the 320 cases compiled in the study, 246 come from Amine Experts' field troubleshooting reports with an additional 74 being sourced from literature or other industry experts. Only cases where production was heavily impacted and there was an in-depth on-site analysis done by the Amine Experts consultant or paper's author were included in the database. Anecdotes or papers in which it is not apparent that the majority of potential causes were considered were not included in the database.

Fig 2: Number of cases of critical failure in all sour and sweet amine systems

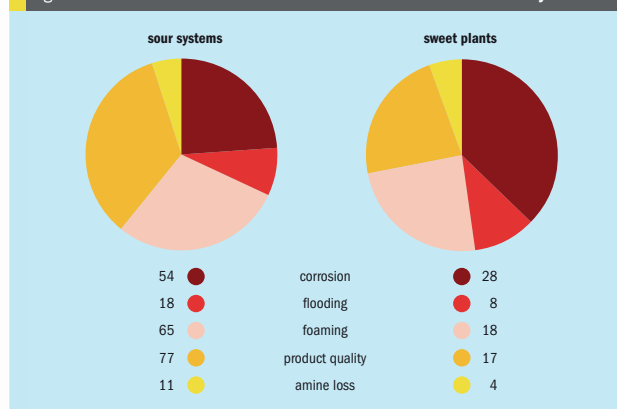
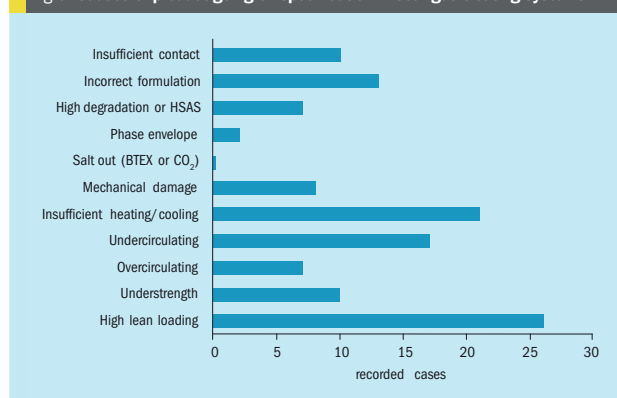


Fig 3: Causes of product going off-specification in sour gas treating systems



In the cases supplied by Amine Experts, the cost in terms of lost production ranged from \$250,000 to \$250 million per case. It is easily possible that the lost production from the failures in this database represents several billion dollars.

The types of production failures experienced fall broadly into five categories: off specification hydrocarbon product, corrosion, foaming, flooding and excessive amine loss (excluding loss caused by foaming). Note that amine losses are often also a side effect of a foaming system, however, amine replacement costs tend to be eclipsed by the cost of production losses. For this reason, amine losses caused by foaming were not considered in the amine losses category as these cases

were covered in the foaming area and double counting of cases was not considered by the authors to be desirable. Any of these problems can result in expensive losses of production. The relative frequency of their occurrence in sweet and sour application is shown in Fig. 2 (16 cases were not easily classed as sweet or sour and have hence been excluded from Fig. 2).

From Fig. 2, it is clear that the most frequent failure areas facing both sweet and sour amine treating systems are foaming, product quality and corrosion. As a result of this relative frequency, these three failure mechanisms will be the primary focus of this article. A clear difference between the failures on sweet and sour systems is the higher frequency of product quality inci-

dents faced in sour amine systems where typically a low ppm maximum H₂S specification needs to be met on the hydrocarbon product.

These cases predominantly occurred within the last 30 years, with the bulk of them taking place in the last 15 years. As a result, this database represents a view of current challenges in industry. Historically, other severe problems may have played a larger role (e.g. hydrogen stress corrosion cracking prior to introduction of post-weld heat treating as a standard in amine plants). Since the early 1980s there has also been a shift from MEA and DEA toward MDEA-based solvents. This shift has impacted on the frequency of certain root causes. Analysis of the frequency of failure of the different amine types will not be undertaken in this article and may form the topic of future work. There are approximately 60 root causes that are a factor in system failures that are generally accepted in industry. The most prevalent of these in sour systems are determined and explained in detail in the following sections.

Product off-specification

Gas leaving an amine absorber must meet certain specifications. What those specifications are depends on the destination of the gas, which is typically to sales, fuel gas or a downstream processing unit. The most common specifications operators are concerned with on a regular basis are:

- H₂S;
- CO₂ (or the overall energy value of the gas);
- total sulphur (COS, mercaptans, etc.).

There are many reasons why the amine system may not perform according to expectations and result in off-specification product, and many of the problems are interrelated. Generally, the most common reasons for gas to be off-specification are high lean loading or improper amine temperature (Fig. 3).

High lean loading: The idea behind the counter-current flow between gas and amine in an absorber tower is that the last thing the gas comes in contact with is the amine with the least amount of bonded acid ions. The reason this is important is because the partial pressure of H₂S in the gas phase will be in equilibrium with that of the H₂S in the amine. This means the more H₂S in the lean amine entering the top of the tower, by default, the more H₂S will remain in the treated gas. This effect

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

SULPHUR
ISSUE 368
JANUARY-FEBRUARY 2017

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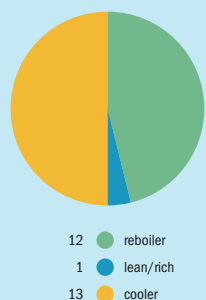
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Fig 4: Off-spec product due to exchanger limitation



is more pronounced the lower the overall pressure of the absorber, with tail gas treating units being on the extreme end of the spectrum.

In cases where a high lean loading is at the root of an off-specification product, there is typically insufficient heat medium flow being sent to the reboiler, insufficient contact with steam in the regenerator column or contamination of the amine with a strong base. Strong base contamination is often the result of over-vigorous addition of sodium or potassium ions in order to neutralise acidic components that build up in the system over time. Monitoring and control of the lean loading in an amine system is central to meeting H₂S specifications.

Insufficient heating/cooling: Heat exchange plays a central role in amine treating. Undersized or fouled exchangers will not be able to supply sufficient heat to adequately heat and cool the amine. The location of the exchanger limitation is shown in Fig. 4.

A limitation on heat supply to the system will decrease the degree of amine stripping in the regenerator, raising the lean loading. This will, in turn, result in high lean loadings and put the product off-specification. Higher lean amine temperatures from a loss of cooling duty will shift the equilibrium in the absorber to favour CO₂ removal over that of H₂S. Usually, the H₂S specification is much lower and more strictly adhered to compared to CO₂. For this reason, the industry standard is to maintain lean amine temperatures at 50°C (122°F) or lower (although there are plenty

Fig 5: Effect of absorber temperature bulge on H₂S removal from gas systems

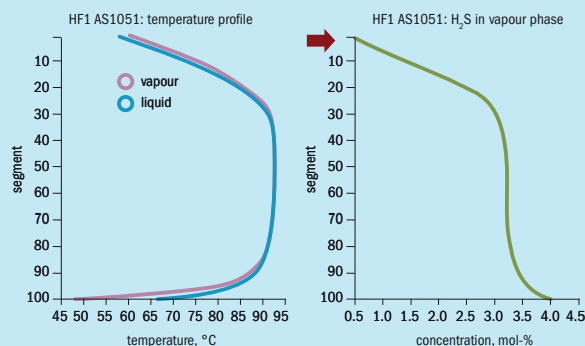
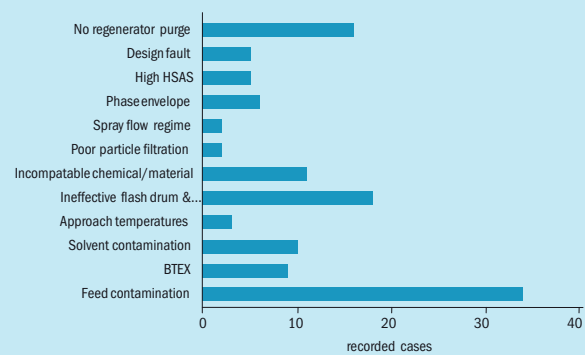


Fig 6: Root causes of foaming in sour gas treating systems



of exceptions). Higher pressure absorbers and better regenerated amines are more resistant to temperature effects.

Further, the temperature profile within the absorber can also determine the rate at which the amine will remove H₂S. Excessive temperature bulges, in effect, render large portions of the tower useless for removing H₂S from the gas, as the amine loses capacity for any sort of absorption. As shown in Fig. 5, the amine does not absorb any H₂S in the hottest sections of the absorber. The temperature bulge is usually controlled by the circulation rate and lean amine temperature.

Over-circulation is often linked to poor exchanger performance, as higher than necessary flow rates will consume heating and cooling capacity. Thus, increasing the

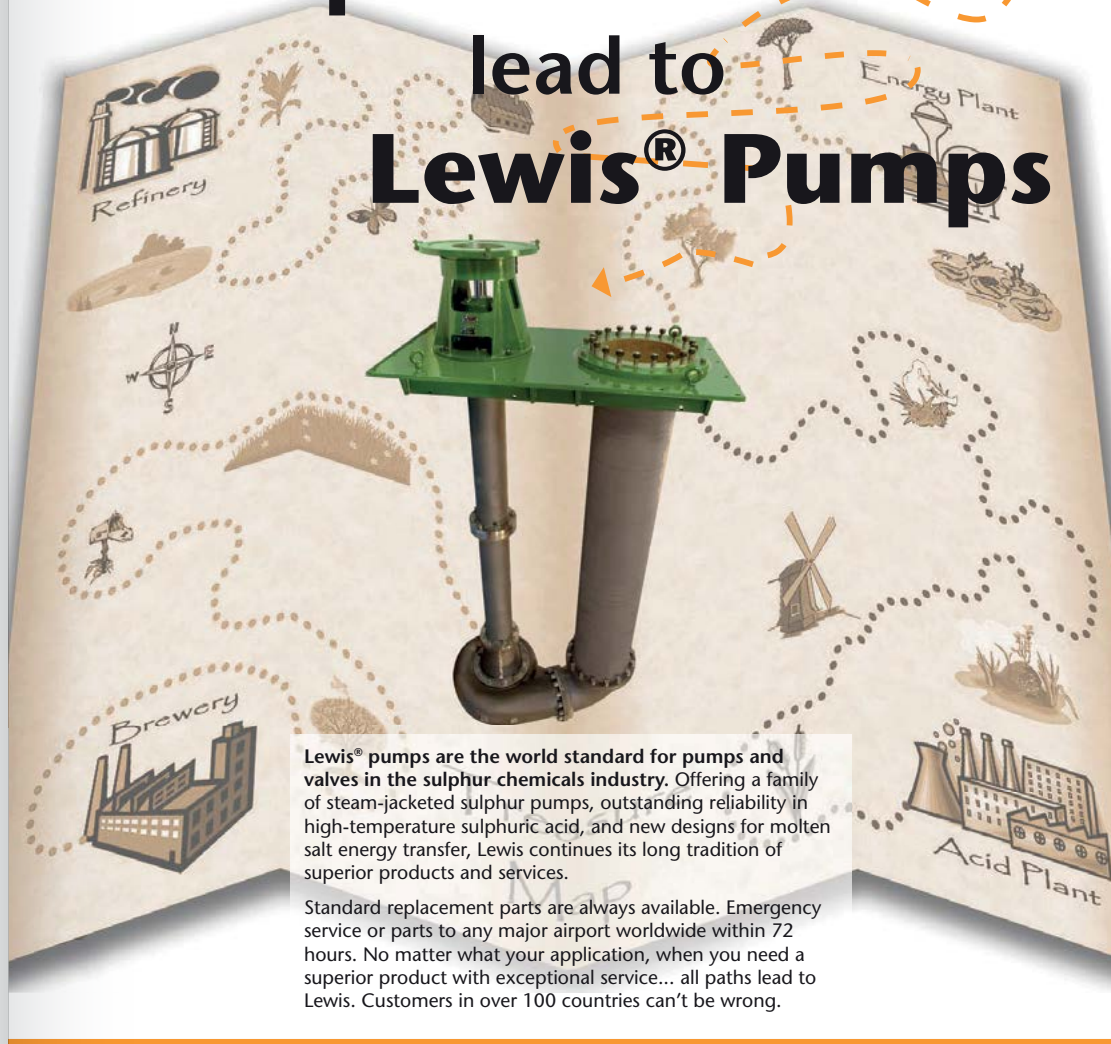
circulation raises both the lean loading and the lean amine temperature. More amine circulation does not necessarily equate to better treating.

The lean loading, system temperature and circulation rate are all intertwined with the H₂S absorption reaction equilibrium. Ensuring sufficient amine regeneration, both operationally and in design/maintenance, is key to meeting H₂S specification.

Foaming

Foaming occurs when gas is incorporated mechanically into the amine liquid phase, resulting in a froth in which gas bubbles are surrounded by a liquid film. The formation and stability of the foam tends to be more strongly dependent on surface

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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

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ISSUE 368
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Table 1: Inlet Contaminants

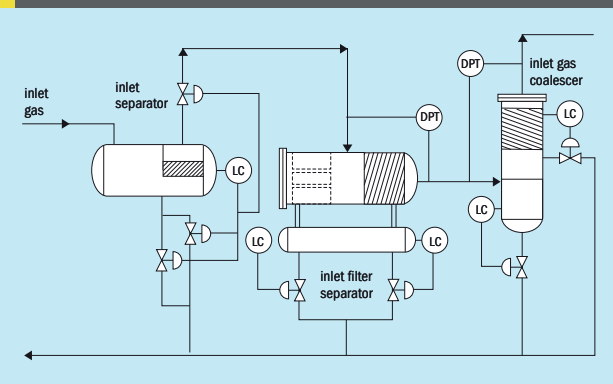
Solid	Liquid
Scale	Compressor lube oils
Desiccant fines	Hydrocarbon condensates
Corrosion products	Amines
	Glycols
Dirt (silica, sand etc.)	Pipeline chemicals
Iron sulphides	Completion fluids
Iron oxides	Condensed water
Adamantane/Diamantane	Well brines
	Organic acids

characteristics of the liquid phase than physical conditions like temperature and pressure¹.

A foaming amine plant is characterised by high and erratic differential pressure measurements in the absorber and/or regenerator columns². It may also display fluctuating levels and erratic flow valve positions and generally results in reduced treating efficiency, amine carryover to downstream systems and loss of production capacity.

The common causes of amine plant foaming have been divided into 12 categories. Of the 65 recorded cases of amine system foaming in natural gas treating plants in the database, 34 of these were found to have foam promoting contaminants entering the plant with the feed, see Fig. 6. From this finding it seems clear that proper treatment of the feed gas is a key focus area for lowering the foaming risk in gas plants.

Fig 7: Inlet separation plan for a natural gas plant with extreme inlet contamination



There are numerous potential inlet contaminants in the feed gas (Table 1) that can promote or stabilise foaming, including: compressor lubrication oils, brine water from downhole, drilling and pipeline chemical additives, soap sticks, iron sulphides/oxides, silica and sand. It is critical to remove these before they enter the amine system where they can accumulate and induce foaming incidents. In order to do this, a combination of 3-phase separators, knock out drums, particle filters and coalescing filters is typically used. For certain feed streams with unconventional contaminants, water washes and silica gel beds may also be necessary.

For a plant with a high feed ingress rate of foam causing contaminants an inlet 3-phase separator followed by a particle filter followed by a coalescing filter (see Fig. 7) is thought to be optimal.

It is critical that inlet separation equipment is attentively monitored and well maintained. A good review of these systems and how to operate them correctly is given in the paper "The seven deadly sins of filtration & separation systems in gas processing operations" by Sheilan and Engel³. Specific areas to be cognisant of:

- Differential pressure instruments should be in good working order and operating limits not exceeded to prevent particle filter and coalescing cartridge blow out.
- Lubricants containing surfactants should not be used in element installation.
- Cartridges that are present in particle filters and coalescers should be installed correctly and securely to prevent bypassing; filter cartridge elements are only as good as the sealing surface

between the element and the vessel, so care must be taken during the cleaning and installation stages to ensure very close seals (double O-ring gaskets are preferred in gas phase applications as they provide the best seal).

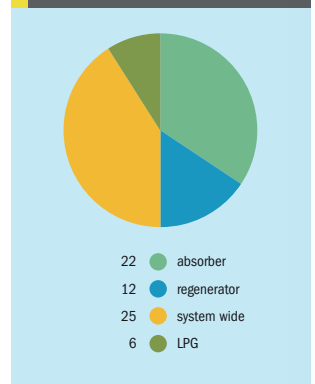
- Level control instruments and valves should be maintained in good working order for equipment to function
- It is important to select a filter cartridge where all of the components in the cartridge are compatible with amine service.
- Differential pressure over a filter should rise slowly with time, zero differential pressure for long periods of time is not indicative of healthy filter operation and should be investigated.

In gas plant service the majority of foaming incidents occurred in the absorber or throughout the whole system (see Fig. 8). There were very few cases of 'regenerator only' foaming recorded in the natural gas industry. This is a result of the large role that inlet contamination plays in foaming incidents as the first area to come into contact with surfactants in an inlet contamination incident is the amine absorber.

Once the amine has been contaminated, foaming will continue in either the absorber or regenerator (or both) until the contaminant has been removed or diluted. In these cases, the flash tank, reflux purges and proper operation/maintenance of the particle filter and carbon bed are critical.

Photos from an inlet contamination incident are shown in Fig. 9. In this case, absorber foaming was causing severe carry-under of gas with the rich amine into

Fig 8: Location of foaming



the amine system flash drum. The excessive gas in the flash drum resulted in high flaring of product gas and frequent pressure safety valve lifting. The cause of the foaming in the absorber was found to be a contaminating aqueous phase surfactant entering with the feed gas. Over the years, the inlet separation system of the plant had been modified from its original design and a line-up allowing the bypassing of the three-phase separator on the plant feed was installed. This bypass was subsequently utilised when the flow meter on the standard line-up developed a fault and an alternative flow meter on the bypass routing was available for accounting purposes. The critical nature of this equipment was not understood by plant management and several months of daily operational grief were endured prior to obtaining management buy-in to restoring the original line up. During this period ~3% of gas processed was flared to atmosphere.

Corrosion

Corrosion is described in many ways, but primarily, it is "the destruction of a metal by the electrical or electrochemical reaction with its environment". In the case of amine units, the environment includes a number of serious corrosion promoters, such as acid gases (H₂S and CO₂), heat stable salts and their acid precursors, chelants (that can remove any protective films that may be formed), velocity and high temperature. Steel is also prone to corrosion because it is not the natural state of iron in the environment. Iron is found as iron oxide in nature, so its tendency to oxidise (corrode) is simply the iron returning to its natural state. For iron to corrode is a completely natural process.

The problem with corrosion is that there are multiple types of corrosion that can occur and many times it is a combination of several factors, rather than a single factor, that generates the corrosive environment in the plants. In no particular order and not in any order of frequency, it is possible to find the following types of corrosion in an amine unit:

- general corrosion;
- galvanic corrosion;
- crevice corrosion;
- under-deposit corrosion;
- pitting corrosion;
- erosion corrosion;
- oxygen-related corrosion;
- stress corrosion cracking;

Fig 9: Inlet contamination foaming

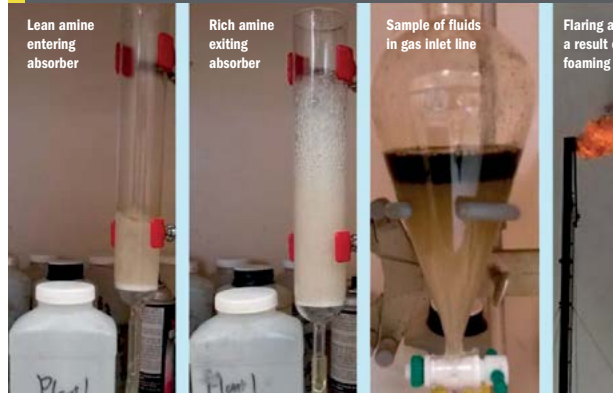


Fig 10: Hydrogen-induced cracking and blistering



- hydrogen damage;
- acid attack.

Except for ionisation, corrosion in amine units is always highly localised. The reason why overall corrosion is not seen is because amines are excellent corrosion inhibitors, and if left contaminant-free, provide excellent inhibition of all amine wetted surfaces. The key to troubleshooting corrosion incidences is to try to determine the logical mechanism, and then determine why that particular part of the plant failed to be protected from corrosion. In natural gas plants, the majority of corrosion is related to acid gas breakout and the subsequent attack of the metal surface, usually at areas of elevated temperature. Since corrosion is a chemical reaction, high temperatures will always accelerate corrosion activity because reactions occur faster and more aggressively at higher temperatures.

Historically, hydrogen embrittlement or hydrogen-induced cracking (HIC) has been a frequent cause of corrosion in the amine industry. In this process the steel becomes brittle and fractures as a result of the introduction and subsequent diffusion of atomic hydrogen into the metal. In amine service, this corrosion is manifested as blisters or cracks in the steel walls of vessels and piping. Atomic hydrogen, which is formed by a number of mechanisms in an amine system, is a small enough atom that it can diffuse through the steel wall to the environment and then re-combine with another atomic hydrogen to form the more stable molecular hydrogen gas. If, however, atomic hydrogen finds an anomaly in the steel (discontinuity or crack caused from an impurity in the manufacturing and forming process) it may stay in the crack until another atomic hydrogen enters the same location causing the unstable atoms to combine to form the stable H₂ molecule. This H₂ molecule is now too large to diffuse through the steel matrix so it is trapped within the steel walls of the vessels where it develops a gas pressure. Ultimately, if enough hydrogen diffuses into and forms hydrogen gas in this discontinuity, the gas pressure exceeds the tensile strength of the steel, leading to the cracking and blistering seen in Fig. 10.

Following numerous failures and a particularly tragic incident in 1984 in which an amine LPG treater at a USA refinery ruptured, causing an explosion and fire that killed 17 people^{6,7} a NACE survey⁴⁴ of 294 amine units was conducted. The survey focused on refineries (272 out of 294

CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

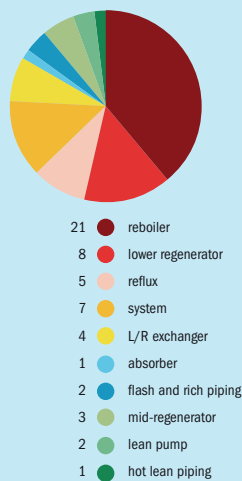
Sour amine systems

SULPHUR
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JANUARY-FEBRUARY 2017

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Fig 11: Corrosion locations in the 54 sour gas treating cases



cases), which were predominantly operating MEA and DEA. The survey found that in 98% of the cases the cracking was linked to the use of non-Post Weld Heat Treated (PWHT) steels. Most facilities built post-1990 use HIC resistant steels. Typically, the plate is tested to meet the requirements of TMO284, which is the NACE standard for the "evaluation of pipeline and pressure vessel steels for the resistance to hydrogen-induced cracking". Reputable steel manufacturers have full control of the entire process, from the steel's metallurgy to rolling, stress relieving and annealing.

Fig. 11 provides a summary of the corrosion locations where localised failures have occurred. Over 50% of corrosion incidents occurring in the reboiler and bottom section of the amine regenerator column, which is connected to the reboiler. The same tendency for corrosion to be predominantly located in the reboiler was also found in a survey of 80 United States Refineries conducted by a refining industry work group¹².

Fig. 12 provides a summary of the probable conditions that led to the corrosion failures. Poor operating conditions were seen to be the most common condition associated with corrosion. Poor operating conditions effectively describes a situation where percentage levels of the acid gases absorbed in the rich amine are stripped in

Fig 12: Root causes of corrosion in natural gas treating systems

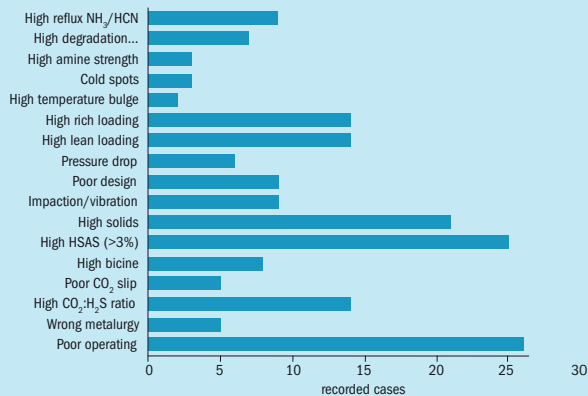


Fig 13: Reboilers corroded by excessive acid gas breakout



the reboiler rather than in the regenerator column. Typical industry guidelines vary between advising that 95 to 99% of the acid gases present in the rich amine be stripped in the regenerator rather than the reboiler.

Excessive amine stripping in the reboiler will often result in pitting type corrosion in the reboiler and lower regenerator. This is avoided by maintaining adequate heat medium flow to the reboiler. This will also generally allow the unit to operate at lower lean loadings which may in turn be beneficial in terms of meeting specification.

Fig. 13 shows examples of reboilers that have been corroded due to excessive acid gas breakout in the reboiler. In both cases, a large proportion of the stripping was occurring in the reboiler rather than the regenerator.

Another major factor in preventing corrosion in amine systems is amine system

hygiene. Good hygiene effectively requires limitation of other potentially corrosive contaminants. These contaminants include: solids, degradation products, heat stable amine salts and amino acids (including bicine).

Solids

Solids in the amine system can contribute to corrosion via erosion of the metal or passivation layers protecting it. Total suspended solids should ideally be kept below 10 ppmw based on use of a 0.45µm absolute filter to test the solution. Note that actual solids levels in the system will differ as a result of increased solubility of many solid precipitates with agitation and temperature. Adequate mechanical filtration and maintenance of the filters is indispensable in maintaining solution quality³.



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Heat stable amine salts

Heat stable amine salts (HSAS) form via reaction of the amine with a contaminant acid that typically enters the system with the feedstock. The resulting acid-amine salt that forms is not readily regenerable at normal amine system stripping conditions and is hence given the name 'heat stable'. While the acid precursors that enter the unit may be present only at trace levels the HSAS that forms will continue to build up in the system until they are removed.

Above 3 wt-% HSAS in the amine the corrosiveness of the amine rapidly increases. Figure 14 shows a section from the bottom part of a corroded regenerator which was replaced as a result of HSAS based corrosion, salts had been allowed to build up in the system to approximately 5 wt-%. This is a standard industry guideline that is consistent with the information in the database in sour applications.

Heat stable salts can be removed from an amine system by several methods:

- amine replacement;
- ion-exchange;
- electro dialysis;
- vacuum distillation;
- thermal reclamation (traditionally only MEA and DGA™).

HSAS may also be neutralised via the addition of strong base species (usually sodium or potassium containing) to form heat stable salts (HSS). There are, however, limits to the amount of HSS that a system may contain before HSSs precipitate out of the solution and plug parts of the unit. Neutralisation also has implications for the HSAS removal methods listed above and is best applied intelligently as part of a heat stable salt management strategy that takes into account the system size, geographic location and HSAS build-up rate.

Degradation products

Degradation products will build up within the amine system over time as a result of thermal exposure and the reaction of amine with CO₂ or oxygen. Some of these products may be corrosive and they should be monitored regularly and kept within vendor recommended limits. Limiting reboiler tube temperatures to less than 165°C (330°F) will largely eliminate thermal degradation in an amine system. Degradation products are normally removed via installed thermal reclaimers in MEA and



Fig 14: Severe acid attack in regenerator bottom section

DGA™ systems. For DEA, DIPA and MDEA, vacuum distillation or inventory replacement are the established options. MDEA is extremely resistant to chemical degradation and degradation rates in DIPA are lower than MEA, DEA or DGA™.

Amino acids

Amino acids develop in amine systems as a result of the ingress of oxidising components e.g. O₂, SO₂ and SX into an amine system causing oxidative degradation of the amine followed by subsequent reaction to amino acids. Bicine (one of several amino acids found in amine systems) has been linked to corrosion in several amine systems^{8,9,10}. Bicine functions as a chelant, increasing the solubility of iron in amine by a factor of just under 500¹⁰. More recently, several other amino acids have also been linked to corrosion in amine systems^{10,11}; with hydroxyethyl sarcosine (HES) being thought to be the most frequently occurring¹¹.

While several industry limits have been proposed for bicine, review of the data does not reveal any consistency in the level at which bicine will corrode a system. This is likely due to the role of other amino acids (e.g. HES) that have historically not been analysed for in amine systems.

Amino acids are zwitterions, neutral molecules with both positive and negative charge, and can thus be removed or partially removed by ion exchange and electro dialysis techniques. Neutralisation is not considered an effective method for amino acid removal. Amine replacement or vacuum distillation will also remove amino acids.

Conclusion

Off-specification product, foaming and corrosion are the three biggest challenges in reliably operating amine units and have

cost the industry billions of dollars over the last 30 years.

For H₂S removal the chemical equilibrium limitation is the central factor in meeting specification. Product not meeting its H₂S specification is most commonly caused by high lean amine loadings, predominantly due to insufficient heat in the system by design or operation.

Foaming problems in gas plants have been found to be due predominantly to inlet contamination of the feed with impurities that have surface tension affecting properties. Adequate inlet separation equipment and proper maintenance of existing equipment is critical.

Corrosion was most commonly found to occur in the reboiler and lower part of the regenerator connected to it. The dominant causes of corrosion in sour plants were found to be excessive stripping in the reboiler due to insufficient regeneration in the regenerator column and poor amine hygiene.

Review of the catastrophic failures of 225 sour amine plants showed that approximately half of the failures could have been eliminated if the following had been in place/performed:

- proper regeneration of the solvent such that lean amine loading is low and the bulk of the regeneration occurs in the regenerator;
- adequate and well maintained inlet separation equipment;
- regular solvent quality monitoring and having a strategy to maintain HSAS, degradation products, amino acids and solids within acceptable limits.

Based on these findings it is advised that plant operators recite the following mantra on a daily basis:

"Heat In, Filters On, Salts Out."

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The authors would also like to thank the authors of the external cases who have published their experiences with sufficient care and detail that others might learn from them. ■

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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

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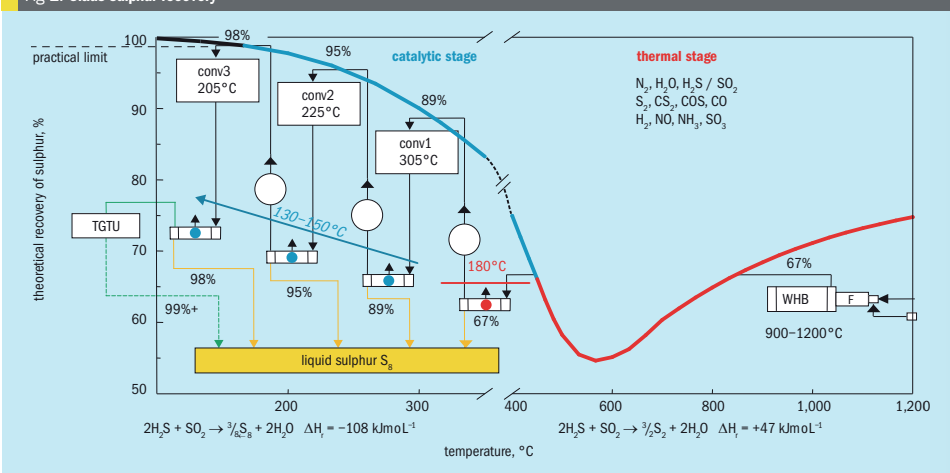
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Simplification of the modified Claus process

P. D. Clark, N. I. Dowling, D. Li, M. Huang and S. S. Bhella of ASRL describe modifications to the existing Claus sulphur recovery process which aim to simplify the overall design and increase the efficiency of the system. The key modification is to incorporate a catalytic section in the combustion chamber to enhance hydrocarbon and CS₂ conversion. A conventional tail gas treatment system could also be replaced by conversion of remaining sulphur values to ammonium sulphate fertilizer, turning the very simplified Claus process into a cogeneration system for sulphur and fertilizer.

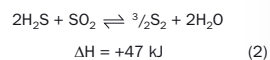
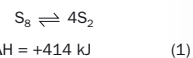
Fig 1: Claus sulphur recovery



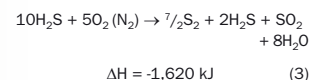
The Claus process is used in the oil and gas industry to convert H₂S to sulphur and water. It is a complex multi-stage technology which requires a combination of combustion and catalytic stages with a final tail gas process also needed to attain overall recoveries of >99.9% of sulphur. As such, it is one of the most efficient processes in industrial operation in terms of emissions. Without the modern Claus process, refineries and sour gas plants could not operate to today's expected norms of environmental stewardship.

The engineering complexity of the Claus process arises from its chemistry, a matter which is elegantly expressed by sulphur recovery as a function of temperature as predicted from equilibrium calculations (Fig. 1). This type of plot, first published by Gamson and Elkins¹, has the peculiar property of showing decreasing conversion to sulphur as temperature increases up to 500°C and then increasing conversion to sulphur at higher temperatures to around 1,500°C, the maximum temperature at which the combustion stage can be operated. This decrease and increase

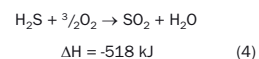
with increasing temperature results from a change in sulphur allotrope at the respective temperatures. Simply, this trend results, overall, from an endothermic transition from S₈ to S₂ as the temperature crosses the 500°C threshold which is part of the overall chemistry that produces sulphur in the combustion chamber:



The Claus process can be followed in Fig. 1 by examination of the diagram from right to left. In addition to the calculated sulphur quantity as a function of temperature, the diagram includes the basic reactors used for each stage. An acid gas, consisting of a mixture of H₂S, CO₂ and H₂O, undergoes partial oxidation in the furnace stage to produce sulphur, H₂O and SO₂ leaving some residual H₂S. The amount of air added is controlled to produce a defined H₂S/SO₂ ratio (2-4), the exact ratio being set according to the type of tail gas process used in the plant. The chemistry of the combustion stage can be summarised in the following equation for a 2:1 ratio:



This equation excludes H₂ and CO, formed as minor products, the CO arising from CO₂ and the H₂ from partial oxidation and dissociation of H₂S. Overall, the combustion stage liberates heat energy because of the exothermic formation of SO₂ and H₂O (equation 4) which outweighs the endothermic production of S₂ (equation 2).



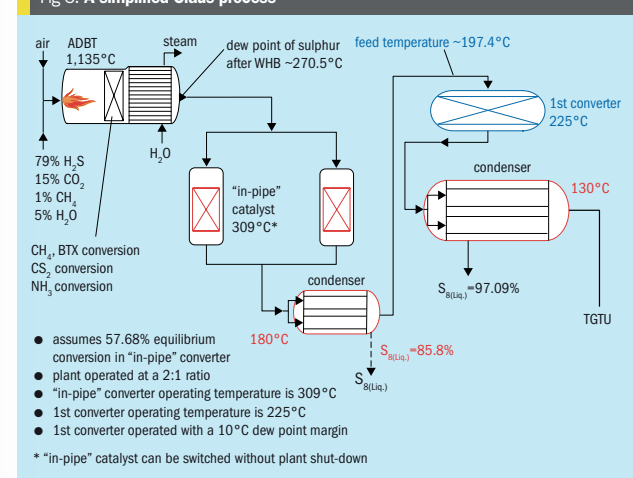
Equations (2, 3 and 4), although useful summaries, are a gross simplification of the combustion chemistry as sulphur is

Fig 2: The chemistry of the Claus process

Combustion stage (overall)	ΔH (kJ)
10H ₂ S + 5O ₂ (N ₂) → 7½S ₂ + 2H ₂ S + SO ₂ + 8H ₂ O	-1,620
Combustion stage (sulphur formation)	
H ₂ S + SO ₂ ⇌ ½S ₂ + 2H ₂ O	+47
[S ₈ ⇌ 4S ₂] ⇌	+414
Catalytic stages	
2H ₂ S + SO ₂ ⇌ ¾S ₈ + 2H ₂ O	-108
CS ₂ + 2H ₂ O → CO ₂ + 2H ₂ S	-68

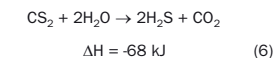
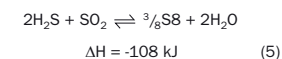
formed by several different pathways, a topic which we have described in numerous other publications^{2,3}. The energy released in the partial combustion is recovered as steam in the WHB and is used for amine regeneration and other applications. Surprisingly, despite the high temperatures in the Claus furnace (900-1,500°C), a residence time of ca. 1 s is required for kinetically favoured products formed in the flame zone to attain the true equilibrium distribution before that product composition is quenched in the WHB. An important point with respect to the modifications proposed in this paper is that the process gas mixture after the WHB is no longer at equilibrium at its new temperature (ca. 300°C) because the rates of all of the processes that would lead to the new equilibrium are relatively slow, unless catalysed. Thus, passage of the gas

Fig 3: A simplified Claus process



through the WHB essentially freezes the gas composition with sulphur production as expected for the adiabatic temperature in the combustion chamber. Comparison of the sulphur recoveries for the relevant temperatures in Figure 1 shows that more sulphur would have been formed in and after the WHB if kinetic limitations had not prevented attainment of the equilibrium for the lower temperature (ca. 300°C).

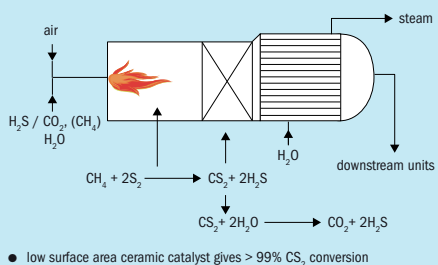
The purpose of the WHB is to recover the energy released by the inter-conversion of S₂ to S₈ and to reduce the temperature of the gas such that liquid sulphur can be recovered efficiently in a condenser (Fig. 1). One of the unfortunate aspects of the Claus process, as currently practised, is that in order to maximise the recovery of liquid sulphur in the condenser, the process gas temperature must be reduced to 180°C for production of more sulphur in the first catalytic unit which operates at ca. 320°C. Operation of the first catalytic converter at >300°C is necessary to efficiently convert CS₂ that is formed in the furnace from small amounts of hydrocarbons that contaminate the acid gas. Thus, the principal reactions in the first catalytic unit are:



In a standard Claus plant, a combustion/WHB/condenser combination plus two re-heater and catalytic converter stages, each followed by a sulphur condenser, are required to reach ca. 95% conversion to sulphur before tail gas treatment is used to increase the overall recovery above 99%, preferably >99.9%. The major chemical processes of the Claus process, prior to the tail gas unit, are summarised in Fig. 2.

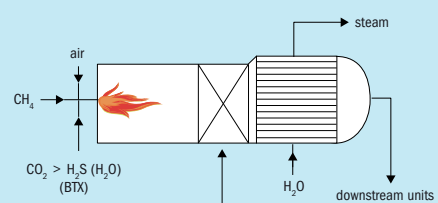
Overall, sulphur recovery in the Claus process prior to the tail gas unit is controlled by the temperature attained in the furnace stage (Fig. 1), the need to operate the catalytic stages to optimise CS₂ conversion in the first catalytic stage and to avoid formation of liquid sulphur in the catalyst pore structure. The adaptations suggested in this article overcome these limitations as well as increasing the overall thermal efficiency of the plant. The key and novel aspect of these modifications is to engender CS₂ conversion in the furnace stage which enables process adaptation to

Fig 4: Formation and destruction of CS₂ in the catalyst modified Claus furnace



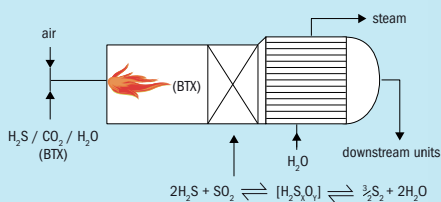
● low surface area ceramic catalyst gives > 99% CS₂ conversion

Fig 6: Processing of low H₂S content acid gas in a catalyst modified Claus furnace



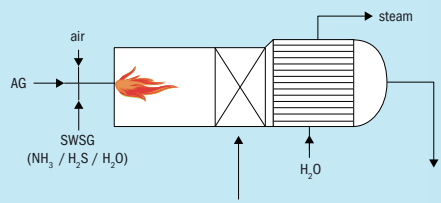
● Efficient conversion of hydrocarbons and CS₂ at 900°C
● Reduced co-firing reduces H₂O in process gas improving conversion to sulphur
[2H₂S + SO₂ ⇌ 3/2S₂ + 2H₂O]

Fig 5: Conversion of BTX and other HC in the catalyst modified Claus furnace



● The ceramic catalyst aids conversion of BTX by reaction with [H₂S₂O₄] species at the catalyst surface

Fig 7: NH₃ destruction in a catalyst modified Claus furnace



● The ceramic catalyst converts residual NH₃ (100-1,000 ppmv) remaining after the combustion

achieve equilibrium conversion to sulphur immediately downstream of the WHB, the consequence of which opens the door to higher thermal efficiency. Importantly, the proposed modifications allow use of existing equipment and are suitable to all tail gas treatment options. A further advantage accrues from improved operation of the liquid sulphur degassing unit.

The proposed simplifications

The furnace

The modifications shown in Fig. 3 represent, hopefully, a simplification to the existing Claus process improving the thermal efficiency and reducing the amount of equipment and catalyst required to achieve the same conversion to sulphur. However, it should be noted that the modifications apply only to the main sulphur recovery units, not the tail gas system. Thus, the overall recovery in any Claus plant would still be defined by the type of tail gas pro-

cess employed to treat the tail gas produced in the last catalytic unit.

The Claus furnace shown in Fig. 3 is essentially the same as in any plant but would be modified to have a ceramic catalyst placed in front of the WHB tube sheet. One of the purposes of the ceramic catalyst is to destroy CS₂ that has been formed in the furnace by reaction of hydrocarbons with sulphur formed by the partial oxidation of H₂S (Fig. 4). In a gas plant, the hydrocarbon impurity is usually CH₄ but in refineries it could be a mixture of larger alkanes, but including CH₄, and aromatic species such as benzene, toluene and xylenes (BTX). The advantage of having the CS₂ conversion catalyst in the furnace, as opposed to in the first converter, is that rates of reaction are many orders of magnitude higher because of the higher temperature (>1,000°C compared to ca. 320°C). If the activity of the catalysts in the furnace and conventional first converter were the same, reaction rates would be ca. 270 times faster in the furnace. Of course, activities

will not be the same, but this simple comparison illustrates that very little catalyst surface area will be required for efficient CS₂ conversion at furnace temperatures.

One of the key challenges for natural gas plants is conversion of BTX in acid gas which contains large amounts of CO₂. In these cases, the furnace adiabatic temperature is usually too low (850-1,000°C) to engender complete BTX conversion to CS₂/COS/CO₂ with the result that some B, T and X are carried to the first converter catalyst. T and X are of particular concern as they result in formation of carsul, an intractable C-S polymer, which plugs the pore structure of the standard first converter catalyst. Studies performed at ASRL showed that a ceramic catalyst operating at furnace conditions results in efficient conversion of BTX to gas phase products (CS₂, COS, CO₂) so preventing degradation of the downstream catalysts. Overall, the ceramic furnace catalyst has the dual function of destroying hydrocarbons, including BTX, and for hydrolysis of CS₂ (Fig. 5). This modification may

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 - Operational process safety experiences at the Shah field
 - Recovery processes for small to medium remote sulphur load fields
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 - Energy performance benchmarking of gas sweetening units
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CONTENTS

What's in issue 368

COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

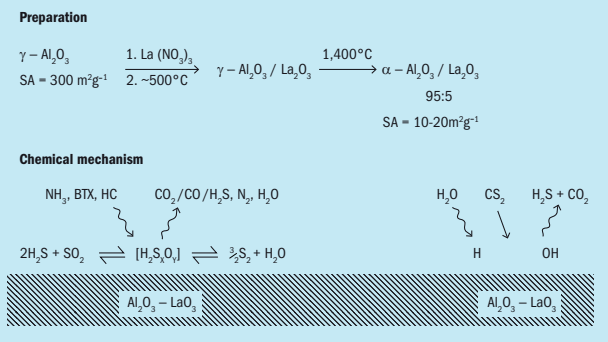
Sour amine systems

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Fig 8: The composition and function of the ceramic furnace catalyst



represent a major improvement for low quality acid gas processing as a combustion temperature higher than the adiabatic value would not be needed. At the very least, the degree of co-firing could be reduced. One advantage of reduced co-firing results from reduction of gas volume, another stems from lessening the amount of H₂O in the system, an important factor in sulphur production because of re-conversion of sulphur to H₂S and SO₂ in the reverse Claus reaction (Fig. 6).

In many refineries, NH₃ produced in HDS units presents disposal problems because usually it has no market value within the local vicinity. In addition, although the NH₃ is a valuable product used in agriculture, contamination with H₂S means that the best option in a refinery is to feed it with acid gas to the Claus furnace where it undergoes partial oxidation to N₂ and H₂O. Overall, destruction of NH₃ in the Claus furnace is complex⁴ requiring a defined addition sequence and mixing of the SWSG, containing the NH₃, with air and AG. Since mixing patterns may change as the system ages, complete NH₃ destruction may not be attained with the result of ammonium salt plugging in cooler components of the plant and plume formation at the stack top⁵. Further studies in our laboratory have shown that the same ceramic catalyst used for CS₂ and BTX conversion is also efficient for NH₃ conversion so, by placing such a catalyst in front of the tube sheet, three "birds" can be controlled with one stone, in this case, a low surface area alumina-based ceramic "stone" (Fig. 7).

A further advantage of having a ceramic catalyst layer in front of the tube sheet is that it may help to protect the tube sheet

and ferrules from physical damage, and, hence, increase the reliability of the WHB. How? One problem encountered in a Claus plant is an unexpected shutdown. In these cases, the furnace chamber must be swept with N₂ to remove flammable material to ensure a safe re-start when conditions allow. The problem with this procedure is that cold N₂ gas causes too rapid a cool down of the tube sheet resulting in cracking of the ceramic in both the tube sheet and ferrules. A CFD analysis presented at the 2015 Vail symposium⁶ showed that the bottom half of the tube sheet – ferrule system was particularly vulnerable because the "cold" and relatively dense N₂ did not mix through the entire combustion chamber causing differential cooling throughout the tube sheet. Possibly, the hot ceramic catalyst, which would be in close proximity or in contact with the tube sheet, would mitigate the rapid cool down of the tube sheet components. Of course, the ceramic catalyst would be subject to thermal stresses itself. The overall solution to this problem is to use hot N₂ gas to flush out the furnace chamber after unexpected unit trips.

What is our furnace ceramic catalyst made from and how would this material be installed in the furnace? The catalyst prepared and used in the ASRL laboratories consists of α-Al₂O₃ modified by La₂O₃. It was prepared by surface doping of γ-Al₂O₃ with a lanthanum salt followed by successive calcinations at temperatures eventually reaching 1,400°C. This process decomposes the La-salt to its oxide which becomes dispersed over the alumina surface, which undergoes transition to the stable α-form above 1,000°C. The calcined material, which has a surface area

of 10-20 m²g⁻¹, is essentially the same as the ceramic used to line the Claus furnace and so will withstand the conditions of the Claus furnace (Fig. 8). The catalytic activity results from the disordered structure imparted by the addition of La₂O₃. In most respects, the catalyst is similar to the material used for the catalyst support employed in hydrogen plants (alkaline earth modified α-Al₂O₃) where it functions to dissociate H₂O to HO and H, species which are required for conversion of CH₄ to CO and H₂. In the current application, the dissociated H₂O species serve to convert CS₂ and BTX, and, likely, aid dissociation of NH₃ and subsequent conversion of NH₂ radicals by sulphur species.

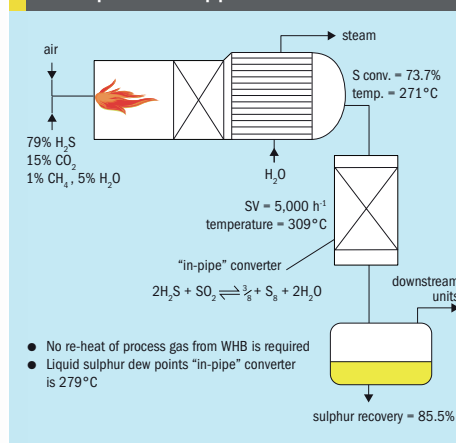
It is envisaged that the catalyst would be installed as part of a chequer wall that stands in front of the tube sheet and, so would use technology already well tested in Claus combustion. Conceivably, the catalyst could be in monolith form, which according to results obtained in the laboratory, would be operated at a space velocity of 10,000 h⁻¹. This space velocity requirement means that the catalyst would occupy 25% of the Claus combustion chamber.

Sulphur conversion in the in-pipe converter

As explained in the earlier description of the operation of the Claus furnace and WHB, the process gas downstream of the WHB contains an amount of sulphur as per the equilibrium existing at the adiabatic temperature of the furnace chamber. The WHB "freezes" the equilibrium for that temperature. This amount of sulphur is considerably less than would be present if the true equilibrium had been attained for the temperature downstream of the WHB (ca. 300°C). Thus, considerable improvement in conversion to sulphur would be obtained by passage of the process gas through a catalytic reactor immediately after the WHB and prior to condensation of sulphur in a condenser. By-passing the WHB condenser has also been considered by others^{7,8,9} although incorporation of a furnace catalyst section downstream of the flame zone was not considered in that work. As is indicated in Fig. 3, no re-heat of the gas would be required for this "in-pipe" catalyst unit as the kinetics of formation of sulphur over standard Claus catalysts is rapid at 300°C.

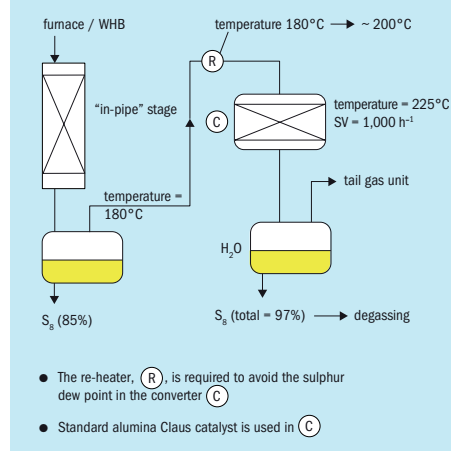
In fact, at 300°C, the space velocity required to attain equilibrium conversion to sulphur is ca. 5,000 h⁻¹, so enabling use of a reactor some 5 times smaller than

Fig 9: Equilibrium adjustment to increase conversion to sulphur in the "in pipe" converter

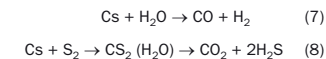


a conventional Claus catalytic reactor. In a standard plant, a lower space velocity (higher residence time) is required (ca. 1,000 h⁻¹) for the first catalytic converter as CS₂ conversion must also be achieved. In the new plant configuration suggested here, CS₂ destruction is accomplished over the furnace catalyst section so allowing use of a small "in pipe" catalyst reactor. Thus, any pressure drop concern over introduction of a catalyst section in front of

Fig 10: The "conventional" catalytic stage



the tube sheet is mitigated by reduction in size of the "in pipe" catalyst section, which in the new configuration (Fig. 9), effectively replaces the first catalytic converter of a conventional Claus plant. Equilibrium calculations (Fig. 9) show that the sulphur recovery after the combustion and "in pipe" converter stages has already reached ca. 85%, a figure considerably greater than a standard plant (ca. 65%) using a standard first catalytic converter. These calculations take into account formation of sulphur (exothermic) and avoidance of attainment of the sulphur dew point in the "in pipe" catalyst section (Fig. 9). Moreover, in laboratory studies, since it has been shown that CS₂ conversion exceeds 99% in the combustion catalyst section, the "in pipe" catalyst does not have to be designed for CS₂ conversion. In the re-modified configuration (Fig. 3), two "in pipe" reactors are shown, the purpose of the second one being back-up in the event that catalytic activity is lost in the original unit. Regarding activity of the ceramic catalyst section in front of the tube sheet, it is worth noting that this catalyst is self-cleaning with respect to an unplanned incursion of hydrocarbon as any solid carbon that might be formed in the furnace would be removed by reaction with H₂O and S₂ produced under normal operation (equations 7 and 8).



Of course, intermediates leading to CS₂ would be hydrolysed on the ceramic catalyst surface, as shown in equation (8).

The "conventional" catalytic stage

After the first condensation of sulphur, the process gas would then be sent to the first and only conventional catalytic stage (Figure 10). Because ca. 85% conversion to sulphur has already been attained in the

Fig 11: Ammonia destruction in the simplified Claus process

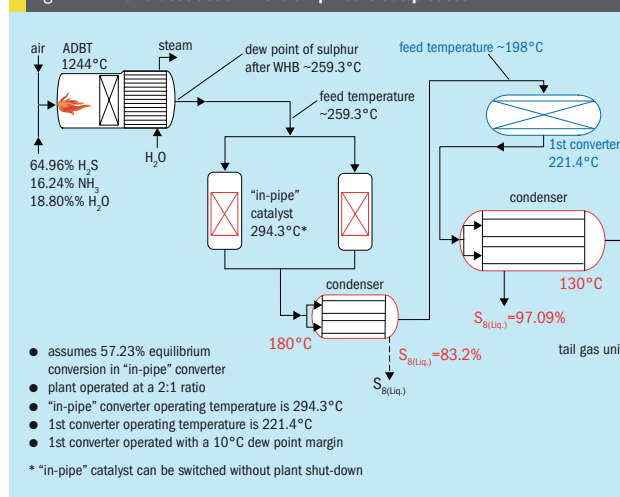
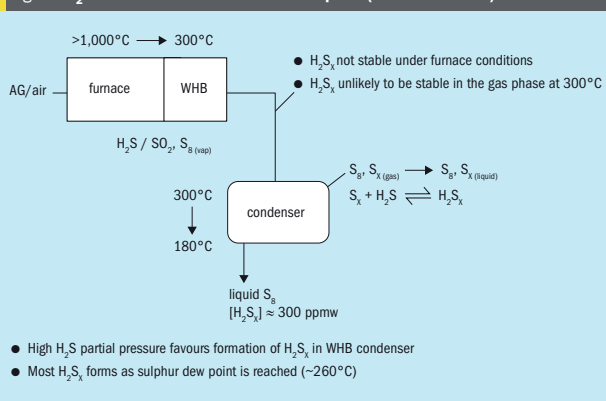


Fig 12: H₂S_x formation in a standard Claus plant (WHB condenser)



furnace and "in pipe" catalyst units, the temperature of operation of this catalyst stage is much lower than a conventional first catalytic converter (ca. 225°C compared to 300-320°C). This temperature reduction presents advantages in terms of the degree of process gas reheat required and the sulphur recovery that can be obtained in this formal catalytic converter stage. Assuming the condenser outlet temperature of the process gas is 180°C, calculations show that the gas only needs to be reheated to 200°C to avoid the sulphur dew point in the catalytic stage.

This equates to reduction in size of the re-heater and to a significant increase in the thermal efficiency of the plant. But, the most important improvement is that overall sulphur recovery after the catalytic stage is ca. 97%. This very high sulphur recovery is possible because of the lower temperature of operation of this last catalytic stage.

Compared to a standard Claus plant, the modifications suggested in this paper enable attainment of 97% conversion to sulphur with only one full size catalytic converter and two condensers, instead of the three condensers required for a stand-

ard furnace/WHB two catalytic converter plant. As already noted, only one re-heater would be needed and the degree of re-heat needed is minimal (ca. 20°C).

As may be noted from Fig. 11, very similar results are calculated for refinery acid gas – SWSG treatment in terms of total sulphur recovery after the furnace, "in pipe" reactor and formal catalytic converter stage. As mentioned previously, the total sulphur recovery for the overall plant is defined by the type of tail gas unit used in the plant, but it is concluded that the modified process (Fig. 11) would ensure very high NH₃ conversion as well as a more reliable system to deal with hydrocarbon incursions, a fact of life for a refinery Claus plant.

Liquid sulphur degassing

The liquid sulphur produced in a Claus plant contains residual H₂S which exists both as dissolved material and in a combined polymeric form, H₂S_x (Fig. 12). In a standard plant, much of the H₂S_x is formed in the condenser immediately downstream of the WHB which, in the plant configuration described in this article, is eliminated. As will be described in detail, removal of the WHB condenser adds another very significant improvement to the overall plant in terms of reduction in equipment size for the degassing system and improvement in energy efficiency for operation of that unit. In order to understand this conclusion, a

Fig 13: Formation of H₂S_x in the simplified Claus process

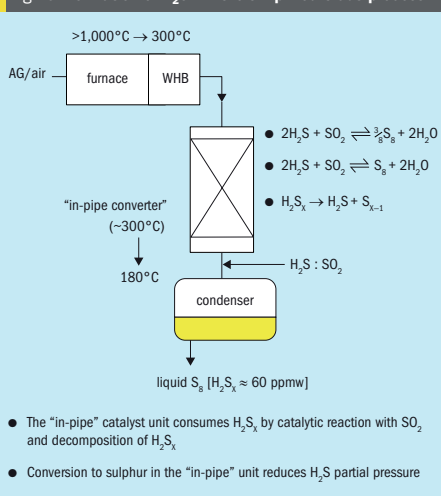
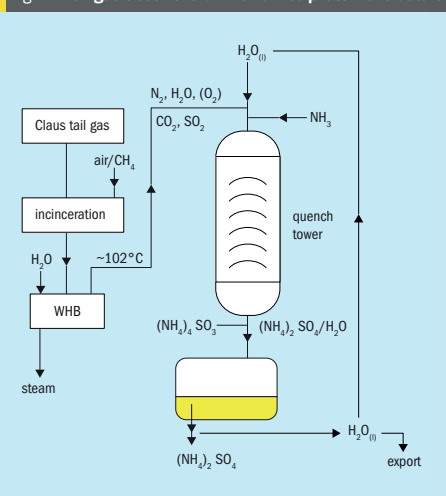


Fig 14: Tail gas treatment: ammonium sulphate manufacture



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COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

Sour amine systems

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ISSUE 368
JANUARY-FEBRUARY 2017

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Fig 15: Tail gas treatment: ammonium sulphate manufacture

Key aspects

- Sulphur recovery is >99.99%
- High thermal efficiency; steam can be used for water evaporation for $(\text{NH}_4)_2\text{SO}_4$ process
- Water and fertilizer are produced
- Known technology for flue gas treatment
- NH_3 is required (might be available in refineries)
- Economics?
- Could be implemented after furnace / in-pipe converter, depending on market for $(\text{NH}_4)_2\text{SO}_4$

brief review of H_2S_x formation and degassing chemistry is in order.

In a standard Claus plant, process gas from the WHB is passed through a condenser to recover liquid sulphur. When the process gas enters the WHB there is no H_2S_x because the temperature exceeds the value at which S-S bonds are stable. Most probably, only very small amounts of H_2S_x are present in the gas as it leaves the WHB even though the temperature (ca. 300°C) is just low enough for formation of S-S bonds. This conclusion is made because liquid sulphur is likely needed to produce and solvate H_2S_x (Fig. 12). Only when liquid sulphur begins to form in the condenser are all conditions met for formation of H_2S_x . Because of the relatively high H_2S partial pressure and the high concentration of S_x radicals present in process gas after the WHB, H_2S_x is formed rapidly as sulphur condensation occurs.

The process configuration suggested in this article will result in much less H_2S_x in the first sulphur condensation step because the process gas is first subjected to the "in pipe" catalytic stage which provides an opportunity for decomposition of H_2S_x as well as for its reaction with SO_2 at the catalyst surface (Fig. 13). In addition, the lower H_2S partial pressure in the gas leaving the "in pipe" unit means that less H_2S_x will be formed in the first condensation. Thus, the very large liquid sulphur flow from the first condenser (ca. 80% of the entire production) should contain no more than 60 ppmw H_2S_x as opposed to the normal 300-400 ppmw. Reference to a reduction in H_2S_x has also been made by others (10), where condensation of liquid sulphur after the WHB has been by-passed. This reduction should equate to improved operation of the liquid sulphur degassing unit by limiting the amount of gaseous sulphur species in the off-gas from the degasser, although any reduction in the volume of

sparge gas required would be dependent on the type of degassing process used for sulphur purification.

Experimental studies on the simplified Claus process

At time of writing, the simplified Claus process described herein has been investigated by calculation and by laboratory experimental studies. The calculations, results of which are shown in Figs 3 and 11, were performed using an in-house free energy minimisation model. Experimentally, the ceramic furnace catalyst described has been made and tested demonstrating >99% CS_2 conversion, complete BTX destruction. These results were obtained using a space velocity of 10,000 h^{-1} at the calculated adiabatic temperature for the acid gases used in the study in an experimental apparatus that duplicated a typical Claus combustion chamber. Since the ceramic catalyst was stabilised at temperatures well in excess of the operating temperature, it is expected that they would be stable over the long-term. But this conclusion can only be justified by an extended laboratory study or demonstration of the technology in the field. The latter option is preferred and could be implemented without significantly altering an existing plant as only minimal retro-fit to an existing furnace chamber is required.

The "in pipe" catalyst section is just a smaller version of a conventional catalytic reactor. Laboratory studies show that a space velocity of 5,000 h^{-1} is sufficient for attainment of the equilibrium conversion to sulphur and the following catalytic reactor is a typical Claus converter running at temperatures shown in Figs 3 and 11. Neither catalyst unit requires specialised catalyst so any existing commercial alumina product should suffice. Clearly, no studies were performed to duplicate the re-heater required after the first condenser as any available indirect reheat technology would be suitable.

Elimination of the conventional tail gas unit

In some sulphur recovery systems, as much as 50% of the capital cost of the plant is for construction of the tail gas unit. As sulphur emissions standards have tightened, operators have opted for reduction of all species to H_2S and recycle of the H_2S to the Claus furnace. This approach, although expensive and requiring multiple

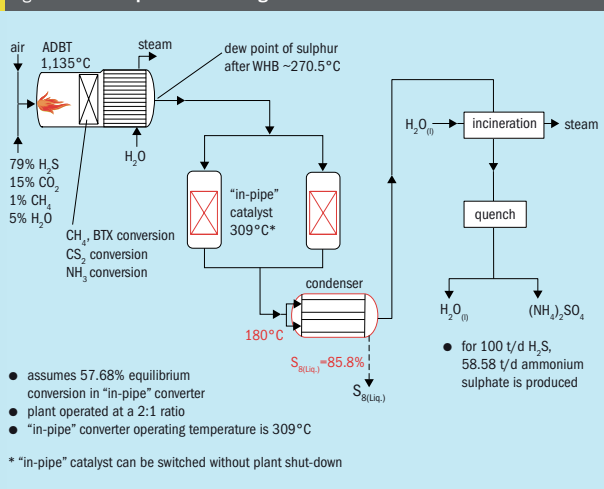
units, has been the technology of choice for over 20 years because overall sulphur recoveries of >99.9% can be achieved. More recently, tail gas oxidation and recycle of SO_2 to the furnace has emerged as a viable alternative enabling >99.9% sulphur recovery. Both methodologies, although practical, are energy intensive, mostly because of regeneration of the solvents used to capture the H_2S and SO_2 .

Here, an alternative approach (Fig. 14) is suggested where, after tail gas incineration, the SO_2 is captured as ammonium sulphate, a valuable fertilizer. There are several interesting facets of this option including increased thermal efficiency for the overall sulphur recovery plant and production of clean water as well fertilizer. Perhaps the main attribute of the scheme shown in Fig. 14 is that sulphur recovery exceeds 99.99% although it should be noted that most, if not all of the energy captured after incineration, will be required for operation of the fertilizer plant. In theory, in addition to capture of SO_2 by reaction with NH_3 , it should also be possible to capture CO_2 with further ammonia to form urea, in itself a valuable high nitrogen fertilizer. Obviously, a source of NH_3 is needed which in some refineries may be available from SWSG but for Claus plants processing sour natural gas, the NH_3 would be sourced externally.

The quench tower (Fig. 14) receives aqueous NH_3 and water to reduce the temperature and sequester the SO_2 in basic aqueous solution as a mixture of ammonium sulphite and ammonium sulphate. Although a simple stoichiometric reaction of NH_3 with SO_2 predicts formation of sulphite, the instability of this salt along with the presence of excess O_2 from incineration leads to a mixture of ammonium sulphite and sulphate. Technology for oxidative conversion of this mixture to the stable sulphate has already been developed¹¹. The major energy input into production of solid ammonium sulphate is concentration of the aqueous ammonium salt solution by evaporation of water, but again, this step is off-the-shelf technology.

The key attributes of this approach to tail gas processing (Fig. 15) include possible implementation after the in-pipe converter system so producing a further simplification of the Claus plant. Thus, in this idealised Claus plant, the only full size converter and condenser are eliminated along with the need for any re-heat whatsoever (Fig. 16). Thus, approximately 86%

Fig 16: Claus: sulphur: fertilizer cogeneration



of the inlet H_2S is converted to sulphur and 14% ends up as ammonium sulphate, equivalent to 58.58 t/d for every 100 t/d of H_2S that is processed. Of course, such a modification could be applied at any stage of a conventional Claus plant.

In a utopian view, always good for the soul, any excess water from Claus plants in desert environments would be used to grow date palms and the like using some of the fertilizer produced in the process to support that agricultural effort. Thus, a sulphur recovery plant becomes a cogeneration system producing sulphur, fertilizer and food.

Conclusions

Studies and calculations described in this article suggest that it should be possible to simplify an existing Claus plant reducing the amount of equipment required and increasing the thermal efficiency of the system. Most probably, the proposed simplifications will increase the reliability of the overall plant.

The key modification is installation of a ceramic catalyst in front of the WHB tube sheet to engender efficient hydrocarbon (BTX), CS_2 and NH_3 conversion. This modification is the key change as it allows operation of the rest of the plant in a more efficient way. The second modification is further conversion to sulphur immediately after the WHB section before condensa-

tion sulphate. This adaptation could be applied after the in-pipe converter introduced in this article or instead of the tail gas unit in a standard Claus plant.

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tion of sulphur, as practiced in a standard Claus plant. It involves operation of a small catalyst unit, referred to in this paper as the "in pipe" reactor, without re-heat of the gas since the formal WHB condenser is eliminated. This step takes advantage of "thermodynamics" using catalysis to overcome kinetic limitations.

These two modifications, in concert, allow operation of a conventional catalytic Claus stage at much lower temperature without the normal concern for CS_2 conversion in the catalyst converter train. The re-heater for this unit requires much less energy input than for a conventional plant as the process gas temperature needs to be raised by only ca. 20°C. If electric re-heat is used, all of the steam produced from sulphur recovery can be used in external operations.

Not only do the proposed modifications enable 96 – 97% sulphur recovery with much less equipment, delay of the initial sulphur condensation should produce liquid sulphur containing no more than 75 ppmw residual $\text{H}_2\text{S}/\text{H}_2\text{S}_x$. This reduction will translate to more efficient degassing operation as less sparge air or other sparge gas will be required so minimising energy consumption and degassing off-gas processing.

Finally, it is suggested that conventional tail gas processing can be replaced by conversion of all remaining sulphur compounds to SO_2 and manufacture of ammo-

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CONTENTS

What's in issue 368

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COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

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CONTENTS

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COVER FEATURE 1

India's refining industry

COVER FEATURE 2

Sulphur recovery project listing

COVER FEATURE 3

Sulphur management options

COVER FEATURE 4

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