Number 363

March | April 2016

SULPHUR

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1 47
2 48
3 49
4 50
5 51
6 52
7 53
8 54
9 55
10 56
11 57
12 58
13 59
14 60
15 61
16 62
17 63
18 64
19
20
21
22
23
24
25
26
27
28
29
30
31
32
33
34
35
36
37
38
39
40
41
42
43
44
45
46

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 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

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Cover: *View of minaret of Sheikh Zayed Mosque through the arches in courtyard of mosque. Abu Dhabi, UAE. Shahid Khan / Shutterstock.com*

CONTENTS

REGULARS

Iran after sanctions

Ambitious plans for oil and gas development.

Improved heat recovery

Utilising heat from sulphuric acid plants.

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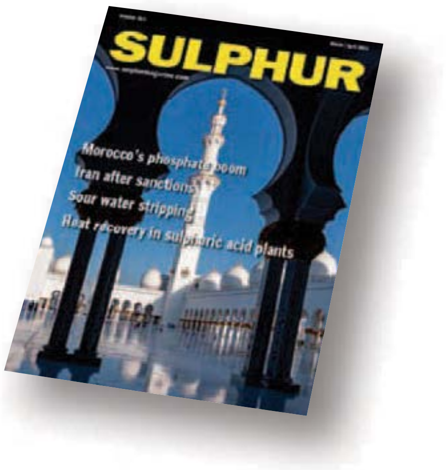
BCInsight

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 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

The sulphur surplus is here



It took its sweet time coming, but as sulphur prices drop below $100/t, and rumours abound of sales as low as $70/t, it looks like the Great Sulphur Surplus predicted since the late 2000s is finally upon us. The fact that it took so long to arrive is probably thanks in no small part to a long boom in phosphate markets, China's extraordinary spell of industrial growth, and the associated effect of that on acid leaching projects for nickel and copper, all of which have boosted demand for sulphuric acid. However, with China's industrial growth now apparently stalled and overcapacity in all key markets, phosphate prices falling and nickel and copper markets in free fall, the demand side of the equation is looking shakier than for some time, perhaps since the banking crash. Although Mosaic in the US has announced some cutbacks in phosphate production, the extent to which this will rescue phosphate markets remains very much open to question.

> **"All of this points to weakness in sulphur markets for some time to come.**

On the supply side, this coincides of course with the start-up of large sour gas projects that have been the largest incremental additions to sulphur supply for some time. The UAE's Shah has been the most notable of these, adding 3 million t/a of sulphur to Gulf supply, and it was notable that January 2016 saw UAE sulphur shipments to China overtake those from Saudi Arabia for the first time. But Chinese sour gas production has also been steadily ramping up, and more is coming from Oman, Iran, Saudi Arabia and Kazakhstan. China imported 9.4 million tonnes of sulphur in 2015, but figure for 2016 seems set to be considerably lower.

Also notable is that, in spite of the flood of oil on world markets, the supply response to oil prices falling from $120/barrel to $30/barrel has been pretty muted so far. This has been deliberate policy on the part of Saudi Arabia, though even Aramco has said it is 'freezing' its oil output, aiming to shut out US shale oil producers, amongst others. There has, finally, been a slight drop in US production, but US producers position on the cost curve seems to have been overestimated. Canadian oil sands projects are certainly taking a hit, and the decline in North Sea production has been accelerated, but no-one is betting on a recovery in oil prices for a couple of years. In the meantime, refining capacity continues to grow, and the push to lower sulphur standards in fuels is driving retrofits in established refineries to desulphurise product streams even further.

All of this points to weakness in sulphur markets for some time to come. There are still some bright spots remaining – Morocco's massive investment in downstream phosphate processing capacity, as we discuss this issue, will continue to see North Africa be a major destination for sulphur cargoes, and Saudi Arabia and China are also continuing to expand phosphate production. China's economy has slowed, but provided it can avoid a 'hard landing', and the signs are so far still encouraging on that score – an upturn will eventually come even in markets like copper, nickel and steel. There is also a continuing push for sulphur as a plant nutrient, and sulphur-containing fertilizer formulations continue to grow in popularity. But with sulphur supply forecast to grow by 7% this year, and demand only 3%, it could be some time before markets move back into balance.

Richard Hands, Editor

1 47
2 48
3 49
4 50
5 51
6 52
7 53
8 54
9 55
10 56
11 57
12 58
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14 60
15 61
16 62
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35
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37
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39
40
41
42
43
44
45
46

Price trends

MARKET INSIGHT

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

SULPHUR

Subdued markets

The sluggish start to the market in 2016 has shown no signs of improvement through February and into March. The improvement in activity hoped for in the processed phosphates market has not emerged thus far, further dampening hopes for an imminent and meaningful recovery in the sulphur market. However, some signs of life in the phosphates market in early March could bode more positive for the second quarter of the year. Phosphates prices in the Americas improved slightly on the back of fresh deals, although China remains subdued, with a focus on trade in the domestic market. Many buyers in other importing markets are also taking a wait-and-see stance, and this is likely to continue. The start of the second quarter is expected to see continued stability or a possible uptick in pricing – assuming demand picks up in the global phosphates market.

Sulphur price developments in the Middle East have led to a more stable outlook for March, with minimal price drops in monthly producer postings. Adnoc/UAE dropped its price by $17/t for liftings for India, down to $88/t fo.b. Ruwais. This followed the $2/t drop from Tasweeq in Qatar for March at $87/t f.o.b.. The posted prices were considered in line with achievable c.fr prices, based on freight of around $10/t from the Middle East to China/India. Meanwhile, Aramco Trading earlier set its March price at $90/t, a drop of $25/t on February. Market sources indicated Tasweeq would not be announcing its regular monthly tender for March shipment, due to an interruption in supply. The Middle East benchmark dipped down to $75/t f.o.b. on the low end of the range at the end of February, representing sales from Iran to China of crushed sulphur. Crushed sulphur is usually around $10/t below the price of granular tonnes. Aramco Trading was also heard securing a sale to China at the start of March. There were mixed views on the price, pegged at above $90/t f.o.b. from one source, while others indicated a price in the mid $80s f.o.b. was more likely.

A bearish factor for the short term outlook is climbing sulphur inventories in China. At the start of March, inventories at the major ports totalled 1.6 million tonnes – a level not seen in over a year. The 1 million tonne mark had become the new norm through 2015, but stocks have been climbing through 2016 so far. Major end users in China appear to be comfortable for the short term, a key factor in keeping them on the sidelines of the market. Spot prices in China were stable in the $90s/tonne c.fr in early March for granular tonnage, while crushed lump was priced around the $80s/t c.fr. Sulphur imports to China dropped 7% year on year in the month of January and the UAE was the top supplier, reflecting a 29% increase on a year earlier. Saudi Arabia, usually the number one supplier for China, ranked fourth after the UAE, Iran and South Korea. Total imports to the country were just over 1 million tonnes in January. In the longer term outlook, with the developments in sour gas projects in China and a slower rate of growth forecast in sulphur consumption compared with the last decade, the country's sulphur import requirement may soften.

Spot prices in India will be tested in FACT's purchase tender for 15,000 – 25,000 tonnes for early April arrival. Spot prices in India have been easing as availability in the market has been healthy. The price range in early March was $95-105/tonne c.fr. The weaker sentiment may see a turning point should there be any improvement in the phosphate market, in the aftermath of the announcement from the Indian government on fertilizer subsidies. The final subsidies on DAP are expected to be announced in mid-March.

Brazilian buyer Vale was heard securing a spot cargo at $90/tonne c.fr for early April shipment. Brazilian sulphur imports saw a dip in January, down 18% on a year earlier. While the US remained the number one supplier, shipments from the UAE saw a surge, at close to 80,000 tonnes. In 2015 the UAE shipped just 160,000 tonnes through the year. The increase in trade to China and Brazil from the UAE is linked to the start up of the Shah gas project. The project has the capacity to produce 3 million tonnes per year of sulphur, which would mean a potential 5 million tonnes of exports from the UAE in 2016.

Sulphur stocks in Western Canada remain in the region of 11 million tonnes. During this lower period of pricing, questions have been raised around whether Canadian producers would look to pour to block and at what price point. Suppliers continue maintain this is not the strategy in the short term – particularly owing to the cost of re-melt and bringing supply back to the market. However, with the prospect of the market moving into a significant oversupply it remains to be seen if prices would remain attractive for producers with high costs of logistics and production. The outlook for growth in sulphur production in Canada has slowed however due to the sustained period of low oil prices. Various projects have been delayed or shelved. This slow down, in combination

Fig 1: **Japanese sulphuric acid exports 2015**

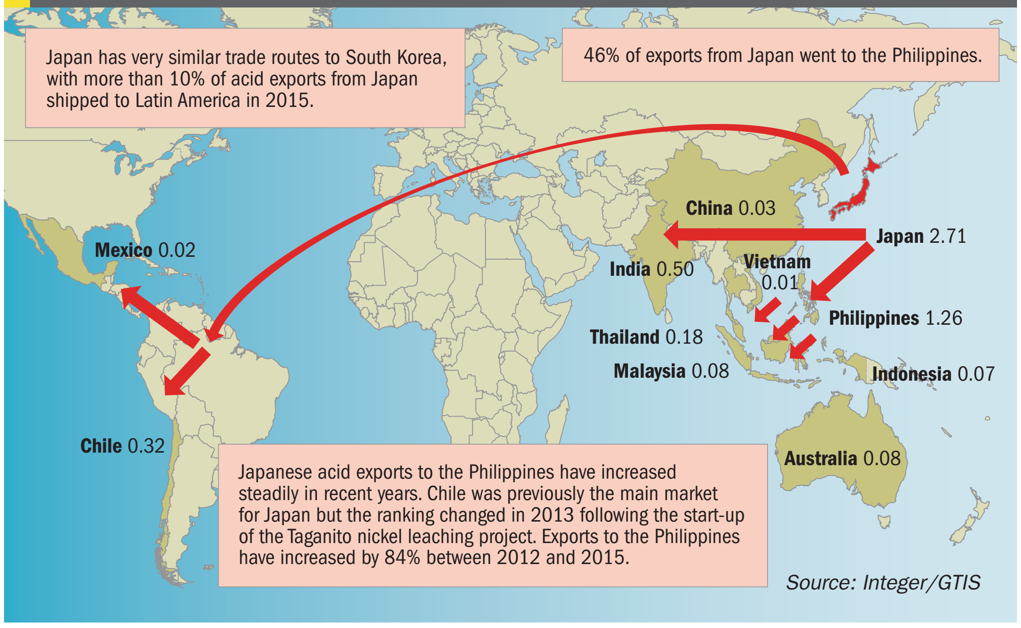

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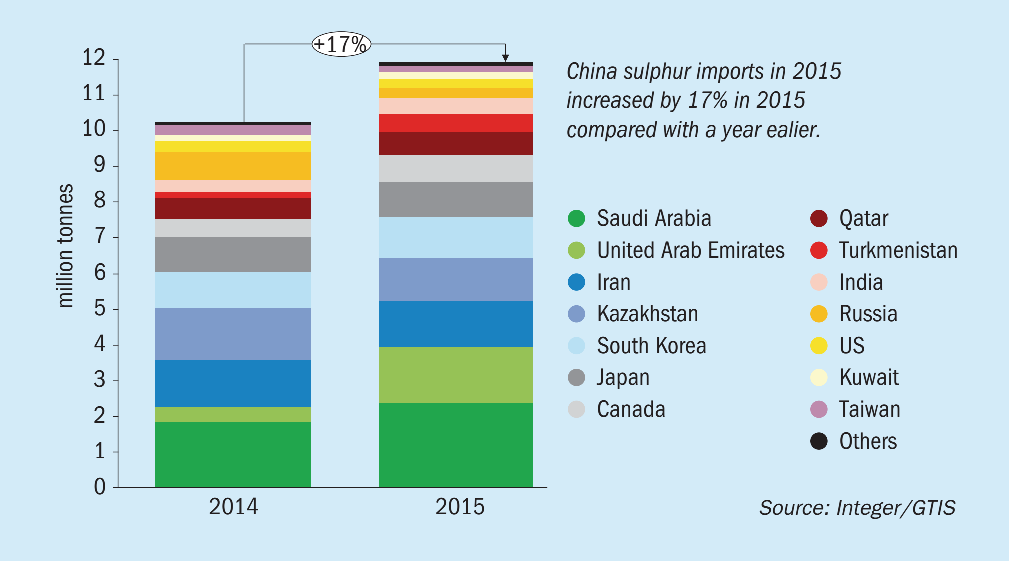
Fig 2: **Chinese sulphur imports 2014-15**



with the depletion of gas based production in Canada, has significantly changed the outlook for sulphur production. Vancouver prices ranged $80-90/tonne f.o.b. through February with further softening anticipated in light of the weaker tone in the rest of the world. The US Gulf range was indicated at a similar level based on shipments to Brazil.

SULPHURIC ACID

Negative territory

Global sulphuric acid prices have dropped below $0/tonne f.o.b. in key exporting regions on the back of weaker sentiment. Northwest European export deals in the spot market concluded through February reflected a range of $-5-5/tonne f.o.b.. Upcoming turnarounds include Aurubis' Pirdop smelter in Bulgaria and Boliden's Kokkola and Harjavalta smelters in Finland. This will likely help balance the European market through Q2 and Q3. However, a key factor influencing forward pricing will be the availability of sulphur based acid from Mexico for the spot market. Cargoes from Mexico to Morocco have been fixed at a significant negative netback. Freight from Mexico to Jorf Lasfar is estimated at around $70/tonne while the price of the cargoes were heard around $20-30/tonne c.fr. However, exports from Mexico may decrease as domestic consumption increases through 2016. In January, Mexico also exported acid to Brazil, around 34,000 tonnes were sold on a spot basis.

In the outlook, another major consideration for European acid exporters is the planned sulphuric acid plant in Cuba for the Moa nickel leaching project. The construction of the 2,000 t/d acid plant will eradicate the need for Sherrit to import acid. This will leave acid producers looking to other outlets to place these diverted volumes.

Acid exports from South Korea and Japan represented close to 5 million tonnes collectively in 2015. Japanese exports saw a 4% drop year on year, due to maintenance turnarounds at smelters. The outlook for exports for the year ahead remains stable. The bearish sentiment has led to prices eroding down to $-10 to 0/tonne f.o.b. in February. Increased supply from the Philippines was cited as a key issue, as well as the weaker sentiment in downstream markets. Quarterly contracts were settled at decreases of $1-3/tonne for East Asia meanwhile and down by up to $10/tonne for Southeast Asia.

Over in Latin America, the market remains long in Chile, with volumes carried over from 2015, and the Brazilian market is also fully stocked for the short term. While some Brazilian buyers are expected to return to the market for Spring planting, there is no sign that Chilean copper producers will be seeking significant additional acid tonnes beyond agreed contracts. Acid imports in Chile were stable in 2015, dropping slightly by 1%. However, the outlook for Chilean import demand remains weak, with a gradual downward trend anticipated as demand declines. Acid imports in Brazil increased in January year on year by 8% to 43,000 tonnes.

In the US Gulf, prices have drifted to $35-40/tonne c.fr, due to the weak sentiment.

Price indications

Table 1: **Recent sulphur prices, major markets**

Cash equivalent	**September**	**October**	**November**	**December**	**January**
Sulphur, bulk ($/t)					
Vancouver f.o.b. spot	115	105-115	120-125	125	113
Adnoc monthly contract	115	125	130	130	122
China c.fr spot	118	110-138	125-145	125	108
Liquid sulphur ($/t)					
Tampa f.o.b. contract	137	110	110	110	95
NW Europe c.fr	185	153-185	153-185	148	148
Sulphuric acid ($/t)					
US Gulf spot	60	45-55	40-50	45	45

Source: CRU

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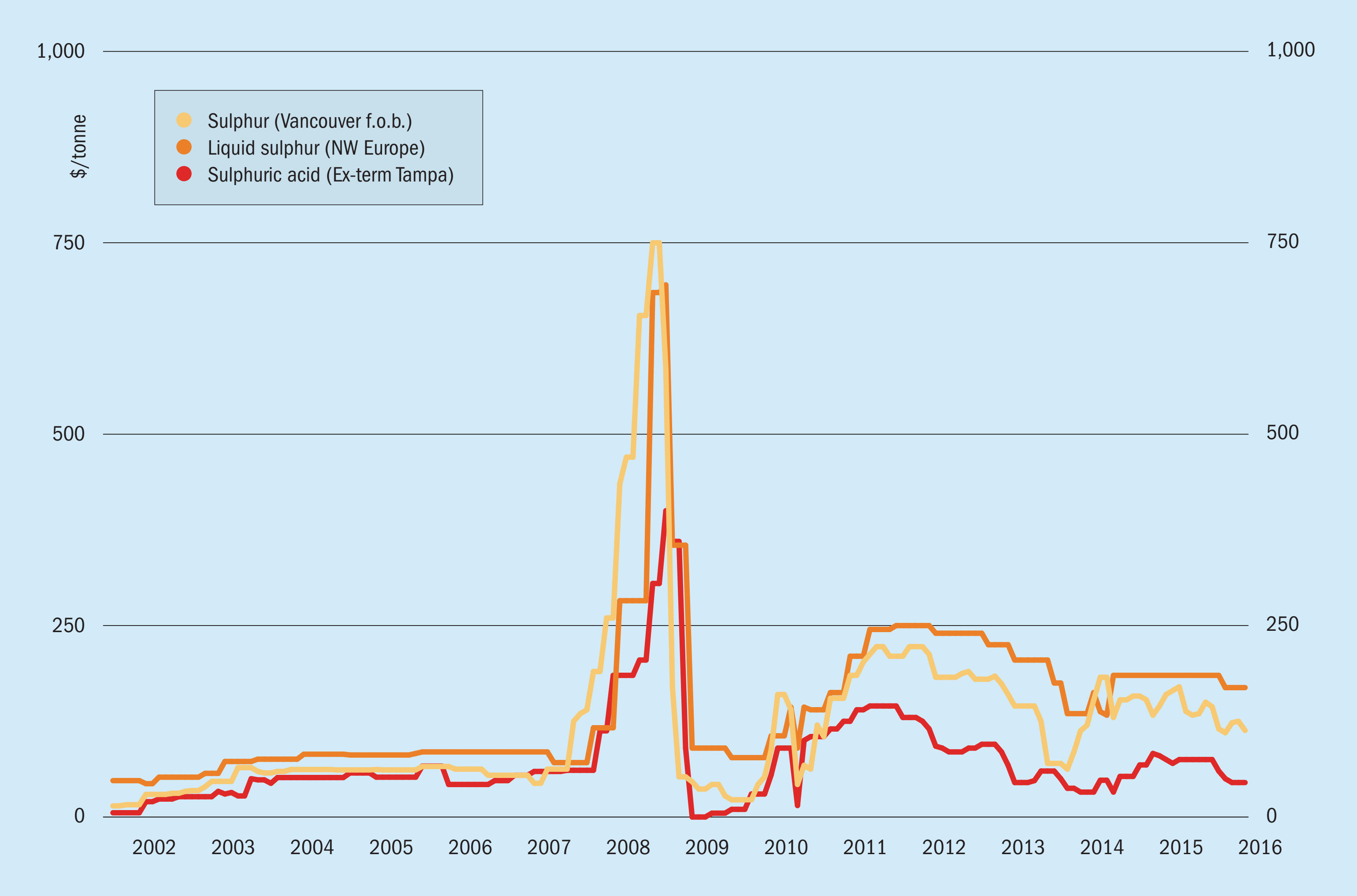
Market outlook

Historical price trends $/tonne



Source: BCInsight

SULPHUR

- Prices are expected to remain stable in the short term due to the weaker sentiment for downstream markets. Further softening is possible – unless the phosphates market continues to show positive signs of recovery to support sulphur pricing through the second quarter.
- The level at which Mosaic utilises its melter in 2016 will be pivotal to the North American market outlook, with the potential for increased availability from Canadian producers, if the railed volumes to the US decline.
- Sulphur availability from new projects in the UAE, Qatar and Iran will remain an important focus for the year ahead and will have a strong bearing on pricing.
- Developments in the oil price will be crucial for sulphur production in the US in the short term as well as in the long term for future growth of oil sands based production in Canada.
- Sulphur demand in Cuba will increase with the start up of the new Sheritt sulphuric acid plant, providing another outlet for producers.
- **Outlook:** The sulphur market balance will be critical to pricing in the outlook. As new capacity comes online, this could create continued pressure on market pricing and remain at lower levels compared to 2015. However, the continued weakness could be limited by an improvement in the phosphates market but also in the broader commodity markets for the industrial sector.

SULPHURIC ACID

- Chile remains long and is not expected to have any significant requirement for spot tonnes in the short to medium term due to ample contract and domestic tonnage.
- Turkey will be an important market for European suppliers – with the Toros sulphur burner and nickel projects shifting dynamics in the local market and changing import requirements.
- Morocco is expected to import in the region of 800,000 tonnes of acid in 2016, on a par with 2015 levels.
- The maintenance turnaround at nickel projects in February-May are expected to lead to some acid cargoes being diverted to India over this period.
- Rourkela steel inaugurated its new sulphuric acid plant, with a production capacity of 125/tonne per day.
- NorFalco was heard moving forward with its plan for a new terminal in the US, at the Savannah Port. The terminal would have the capacity for full size tankers and is due for completion in the second half of 2016.
- **Outlook:** Acid prices are likely to stabilise in the West in the upcoming quarters due to smelter turnarounds and contract commitments. The wildcard factor is Mexico and the extent to which exports will continue through 2016 and at what price level. Meanwhile in Asia, availability from smelters may put pressure on the regional market, as well as the slowdown in demand in Latin America.

Sulphur Industry News

UNITED STATES

Gloomy outlook for US oil and gas companies

The depressed oil price environment is leading to a gloomy outlook for North American exploration and production companies, and significant capex cuts are needed in order for companies to align spending more closely with cash flow, according to a new analysis from IHS. Their peer group analysis of North American E&Ps assessed the impact of lower oil and gas prices on 2016 cash flow estimates and indicates that to maintain spending-to-cash-flow in the historical range of approximately 130% percent, spending would need to be cut by a further $24 billion, or 30%, from the most recent estimates, a cut of almost 50% from 2015 levels.

Paul O'Donnell, principal analyst at IHS Energy and author of the analysis said: "given that most companies made preliminary 2016 spending plans when the price outlook was comparatively higher, we expect to see further spending cuts announced throughout the fourth-quarter 2015 earnings cycle that reflect the current price environment."

The 2016 low-case price scenario assumes $40/bbl oil and $2.50/tcf of gas, and projects that North American E&Ps will spend 188% of cash flow, as compared with 133% under the base-case scenario, which assumes a $50 /bbl and $2.75/tcf. Under this scenario, for the group to show real spending discipline and live within cash flow, annual spending would have to be reduced by at least 64% compared with 2015. "These spending cuts will be particularly troublesome for the highly leveraged companies," O'Donnell said. "These E&Ps are torn between slashing spending further to avoid additional weakening of their balance sheets, and the need to maintain sufficient production and cash flow to meet financial obligations."

Refinery investigated over SO_2 release

The Delaware City Refinery, owned by PBF Energy, is being investigated by the Delaware Department of Natural Resources and Environmental Control (DNREC) over a release of sulphur dioxide on February 18th. According to DNREC, 144 lbs (65kg) of SO_2 was released during the incident.

Marathon faces opposition to SO_2 emissions

Marathon has applied to increase SO_2 output from its Detroit refinery by 22 t/a. Around 75% of this is already covered by existing permits and requires no additional approval or review, but the remainder would require a new air pollution permit, although Marathon says that other projects in the works will, over the next few years, result in a net reduction of sulfur dioxide emissions from the refinery. The emissions would be the result of Marathon's plans to update its liquefied petroleum tank storage and install equipment to meet an EPA mandate to produce lower-sulphur gasoline beginning in 2017, as part of a so-called 'Tier 3' project.

The request to extent the air pollution permit has drawn considerable opposition form local residents, and Detroit Mayor Mike Duggan has threatened a lawsuit if the proposal moves forward. The region around the refinery is currently considered by the EPA to be "in non-attainment" of federal guidelines for emisisons of SO_2 and several other air pollutants, although the refinery is a relatively small emitter compared to a nearby coal-fired power station which produces the lion's share (97%) of SO_2 emissions in spite of installed scrubbing technology. Marathon also saus that it has reduced air emissions by over 70% since 1999, and as part of the Tier 3 project will reduce SO_2 emissions by 5.2 t/a at a cost of $2 million. It has also committed to reduce flare stack emissions as part of a settlement with the EPA, and a $58.5 million project will install or modify gas recovery systems on two of its flares, and lead to the shutdown of a third flare. Those modifications will reduce sulfur dioxide emissions by 50 t/a according to the company.

WORLD

Oil prices have fallen "too far"

Forecasters appear united in their view that oil prices have fallen too far, and that a rebound to higher prices is likely witihn the next couple of years. A recent report by Bank of America Merrill Lynch concludes that "the current forward oil price structure has fallen too far to enable a medium term balance in supply and demand" and that "oil prices need to average $55-$75/bbl so that non-OPEC output can re-attain 2015 levels by 2020 and prevent a huge shortfall". The report estimates that the price drop has already lead to a demand increase of 1.7 million bbl/d in 2015, and that this will increase by 5.9-8.4 million bbl/d over the next five years, depending on the prevailing price, while oil prices at $30/bbl to 2020 could lead to an output shortfall of 4.8 million bbl/d.

The International Energy Agency (IEA) likewise has warned that it expects oil prices to start recovering in 2017, and that this will be followed by a sharp spike in price as supply will have shrunk following under-investment by struggling producers. Brent crude is currently hovering between $30-35/bbl, far below its high of $115/bbl in June 2014. The IEA expects global oil supply will grow by 4.1 million bbl/d between 2015 and 2021, as compared to an increase of 11 million bbl/d between 2009 and 2015. It also expects investment in oil exploration and production to fall by 17% in 2016 following a 24% decline last year.

MALAYSIA

Enersul to supply sulphur forming units to RAPID

Calgary-based Enersul LLP says that it has been awarded an equipment supply contract from Petrofac International for the Petronas Refinery and Petrochemical Integrated Development (RAPID) project in Pengerang, Johor Province, Malaysia. This contract is for the supply of five Enersul *GX*™ sulphur granulation units. This project will be delivered in the fourth quarter of 2016.

The RAPID project consists of a world-scale integrated site which includes refining activities and petrochemicals production. The refinery will have a capacity of 300,000 bbl/d, while the steam cracker's combined annual production is anticipated to be more than 3 million t/a of ethylene, propylene and C4-C6 olefins products. Products from the refinery and steam cracker will be the feedstock to produce premium differentiated specialty petrochemical products.

UNITED ARAB EMIRATES

Oxy considering expansion of Shah project

Occidental Petroleum, which owns 40% of the major Shah sour gas project in Abu Dhabi, via its Al Hosn joint venture with the Abu Dhabi National Oil Company (Adnoc), says that it is looking at the possibility of the expansion of Shah, which reached full production capacity last year. Speaking to the UAE's Gulf News, president and CEO Vicki Hollub said: "Abu Dhabi would like to have additional gas. We would certainly want to make that happen. The expansion would add about 25-50% production to the plant." According to Hollub, Occidental is conducting a feasibility study at the moment and will take a final decision on whether or not to proceed by 2018.

Shell bows out of Bab

Royal Dutch Shell says that it has decided not to prceed with the Bab sour gas project in Abu Dhabi. In a press statement the company said that falling oil prices have played a key part in the decision.

"Following a careful and thorough evaluation of technical challenges and costs, Shell has decided to exit the joint development of the Bab sour gas reservoirs with ADNOC in the emirate of Abu Dhabi, and to stop further joint work on the project," it said. "The evaluation concluded that for Shell, the development of the project does not fit with the company's strategy, particularly in the economic climate prevailing in the energy industry."

Shell has been engaged in a worldwide review of its operations as part of cost-cutting measures due to the gloval fall in oil prices. Last year it abanoned drilling projects in Alaska, and shelved the Carmon Creek oil sands project in Canada.

MEXICO

Claus catalyst recovery project

The Mexican Oil Institute (IMP) has developed a titanium dioxide catalyst reactivation project for Claus plants used in sulphur recovery units (SRUs) run by Petroleos Mexicanos (Pemex) Gas and Basic Petrochemicals. Dr. Roberto Garcia, project manager of the Research Directorate of the IMP, said that the focus of the research is to recover deactivated TiO_2 catalysts from SuperClaus plants, saving Pemex up to $10,000 per tonne of catalyst. The project covers three lines of research: reactivation of the titanium oxide catalyst, treatment for disposal by biological methods, and evaluation of new alternative catalyst formulas. The first strand has so far achieved recovery rates of 20%. In the second, bacteria from sulphur-containing water has been analysed and modified to achieve a higher resistance to the element. The biological treatment system proved capable of removing between 91-100% of the sulfur in the catalyst over a period of 21-35 days. Registration and deposit of two bacterial cultures in the German Microbial Culture Collection was also achieved, which yielded a patent. The third strand of research has looked at the best catalyst systems for selective oxidation of the sulphur, focusing on titanium nanotubes, catalytic systems of titanium oxide, and mesoporous stabilised materials. These systems were modified with iron, and a series of catalysts from silicon oxide were found.

AZERBAIJAN

KT to license SRU to SOCAR

The State Oil Company of Azerbaijan Republic (SOCAR) has signed an agreement with Maire Tecnimont subsidiary Kinetics Technology SpA for the reconstruction of the Heydar Aliyev oil refinery at Baku. The signing ceremony was attended by SOCAR president Rovnag Abdullayev, Gianni Bardazzi, chairman of KT-Kinetics Technology and vice president of Maire Technimont Group, and the Italian ambassador to Azerbaijan Giampaolo Cutillo.

In his welcoming speech, SOCAR President Rovnag Abdullayev said the $1 billion refinery reconstruction includes the modernisation of old refinery units and construction of new units to increase the refinery's capacity from 6 million t/a to 7.5 million t/a, including boosting the catalytic cracking unit's capacity from 2 to 2.5 million t/a and improvement to Euro 5 sulphur standards. The reconstruction also includes tha new sulphur recovery unit, which KT licensed to SOCAR.

SOCAR also signed contracts with Austrian Pörner Group, Axens and UOP relating to the refinery modernisation.

INDIA

Modi inaugurates Paradip refinery

Indian prime minister Narendra Modi led an inauguration ceremony in January for the new Indian Oil Company (IOC) refinery at Paradip in Odisha state. The $5 billion refinery, which has a capacity of 15 million t/a, including 5.6 million t/a of diesel, 3.8 million t/a of gasoline, and 2 million t/a of kerosene, has taken nearly 16 years to realise, but has pushed IOC into being India's largest refiner, overtaking Reliance Industries. IOC's other eight refineries have a combined capacity of 54.2 million t/a, while Reliance, with twin refineries at Jamnagar in Gujarat, has 62 million t/a of capacity.

Paradip, which delivered its first consignment of products in November 2015, was designed to process cheaper high sulphur heavy crude oils to Euro-V quality levels. It is one of the most modern refineries in the world, with a Nelson complexity index factor of 12.2. Associated facilities remain under development, including a polypropylene plant scheduled for completion in 2018. The refinery also has plans to set an ethylene recovery unit/monoethylene glycol (MEG) unit expected to be completed by 2021, and IOC is evaluating options of smanufacturing paraxylene, PTA and synthetic ethanol at Paradip.

COLOMBIA

Ecopetrol studies revival of old refinery units

Colombia's state-controlled oil company Ecopetrol is studying the viability of integrating its new 165,000 b/d Cartagena refinery with older, decommissioned facilities. The current Reficar refienry is adjacent to a closed 78,000 bbl/d facility, and Ecopetrol has indicated that the tie-ins and expansion could lift production above 200,000 bbl/d at Reficar. Reficar is currently runing at 110,000 bbl/d with process units still progressively coming on-stream, including petcoke and sulphur production, although cost overruns from the original $4 billion estimate up to an estimated $8 billion have led to arguments with contractor CBI and an official investigation. Ecopetrol has admitted that some of the cost overrun was justified because of damages and delays associated with flooding in 2012 and labour disputes in 2012-13.

CANADA

Production begins at West Ells

Sunshine Oilsands Ltd says that it has successfully begun first oil production at the West Ells project in Alberta. West Ells

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region covers 9,800 hectares within the Athabasca oil sands region. Sunshine's focus is on evaluating and developing its oil sands assets, targeting initial production of 10,000 bbl/d of heavy oil. The company says that it is fully committed to advancing its corporate initiatives to ensure that West Ells achieves a smooth start-up of Phase 1 facilities and achievement of nameplate capacity of 5,000 bbl/d. Phase 2 is expected to add an additional 5,000 bbl/d. The West Ells asset area has an ultimate development potential of 130,000 barrels per day, according to Sunshine.

'We are pleased that the first oil production following the first steam injection in September this year has been successfully commenced. We look forward to demonstrating reservoir potential towards achieving nameplate production from the West Ells Phase 1 project. 'said Mr. Sun Kwok Ping, Executive Chairman of Sunshine.

NETHERLANDS

Fluor completes acquisition of Stork

Fluor Corporation says that it has closed its acquisition of Stork Holding BV. Fluor announced in early December 2015 that the company had agreed to purchase 100% of Stork's shares from UK-based private equity firm Arle Capital Partners.

"I welcome our 15,000 new colleagues from Stork and we are excited to have them join the Fluor family," said David Seaton, chairman and CEO of Fluor. "Fluor's most important asset is its people, and we are fortunate to have found in Stork a company that shares the same values, pride and global heritage as we do."

Stork is a global provider of maintenance, modification and asset integrity services associated with large existing industrial facilities in the oil and gas, chemicals, petrochemicals, and power markets. Stork, along with Fluor's current Operations & Maintenance organization, will be led by Stork CEO Arnold Steenbakker and reported financially under the Global Services business segment.

UNITED KINGDOM

Comments invited on steels for sour service

BSI is developing a new standard, BS 8701 for determining the susceptibility to cracking of pipeline steels in sour service. The draft standard is now open for public comment until 31st March and comments are invited to ensure the final standard is robust. This standard updates the 1996 HSE Guidance document OTI 95 635, which features the protocol for ensuring pipelines are properly tested to avoid environmental damage from pipeline failure. Different factors have to be taken into account when testing pipes: such as environments, age of pipeline, materials used, coatings etc. In addition to this, the types of corrosion must be accounted for too, such as sweet corrosion (wet carbon dioxide) and sour corrosion (wet hydrogen sulphide).

David Fatscher Head of Market Development for Sustainability at BSI said: "This standard can help ensure that spillages from corroded pipelines is minimised, as this type of spill can be just as detrimental to the environment (be that marine or otherwise), as a tanker spill. We aim to keep all relevant guidance up to date, and by developing BS 8701 we can help manufacturers keep an eye on the hydrogen sulphide based wet pipeline scenarios."

The final standard is expected to be published in July 2016.

Sulphuric Acid News

PERU

Exports begin from Las Bambas

In defiance of the current run of low copper prices, which are currently at their lowest levels since 2009, Chinese-owned miner MMG has successfully delivered its first 10,000 tonne shipment of copper concentrate to China from the new Las Bambas mine in January, and further shipments were scheduled for February. The site, which MMG, controlled by China Minmetals Corp, bought from Glencore in 2014 for $5.8 billion, is now on course to be one of the largest copper mines in the world, behind only Escondida and Cerro Verde, producing 200,000 t/a of copper in 2016 and 400,000 t/a in 2017. New mines in Peru coming on stream this year are expected to double production to 2.8 million t/a of copper, placing the country in second place globally behind Chile, overtaking China's production.

Study for Bayovar phosphate project

Global advisory firm Advisian has signed a contract with Canadian exploration and oil production firm Americas Petrogas to study the development of phosphate resources in Peru. Advisian, a subsidiary of WorleyParsons, has agreed to conduct a preliminary economic assessment (PEA) into Petrogas' Bayovar phosphate project. The site, in the Sechura district of northern Peru, is believed to hold significant phosphate reserves. Petrogas began borehole drilling at its Bayovar 5 and 7 concessions in early December to test the potential for near surface phosphate mineralization. The PEA compiled by Advisian is expected to be released in the second quarter of 2016 and will incorporate results from recent drill holes and re-sampling of drill holes from 2011 and 2012.

Uranium leaching proposal

Plateau Uranium is looking to develop a uranium mining operation on the Macusani Plateau in Peru. The company recently issued an analysis indicating that it has access to at least 124 million lbs of identified U_3O_8 resources, and is looking to produce 6.1 million lbs/year over a 10-year mine life at an operating cost of $18.26/lb, roughly in line with Kazakh production costs. The company currently controls the only known uranium resources in Peru, and is looking at recovering 30,000 t/a of ore via open pit mining and a further 2,700 t/d via room and pillar mining at Kihitan. Processing options being examined include heap leaching and tank leaching, for 88% and 93% uranium recovery respectively. Sulphuric acid consumption is estimated at 9kg/tonne (revised down from the previous 25kg/t), for a projected acid consumption of around 750,000 t/a at peak capacity.

POLAND

Technip to provide roaster for copper smelter project

Technip has been selected to provide its Dorr Oliver *FluoSolids* roaster system for Polish mining company KGHM's Glogow I copper smelter optimisation project. The 480 t/d system will include the roaster, dry concentrate feeder and calcine cooler, and in-bed steam coils for cogeneration of electricity. These components will remove organic carbon and sulphide sulphur from copper concentrate, reduce smelter emissions and improve copper production at the site, according to Technip. Under the contract, Technip will also provide erection supervision, commissioning, and startup, as well as training assistance to KGHM. The project is scheduled for completion in 2017.

Technip Stone & Webster Process Technology president Stan Knez said: "Technip's extensive experience in roasting technology along with our proven ability to meet a demanding schedule makes us uniquely qualified for this important project."

SOUTH AFRICA

Foskor operating at record low levels

Strikes and equipment failures during 2015 caused output at South African phosphates and phosphoric acid producer Foskor's Richards Bay plant in KwaZulu-Natal to be at its lowest level since the company was established in 1976. The company reports that overall production was down to 300,000 t/a in 2015. At full capacity, Richards Bay is able produce 2.2 million t/a of sulphuric acid, 720,000 t/a of phosphoric acid and 300,000 t/a of phosphate fertiliser. However, the company says that it has a number of capital projects lined up that will increase environmental compliance, improve plant processes and boost production.

ZIMBABWE

Acid plant "remains on the table"

Zimbabwe's biggest platinum producer Zimplats has indicated that construction of an $80 million sulphuric acid plant remains a significant possibility as it re-evaluates its capital spending plans in the light of declining metal prices. Zimplats, a subsidiary of Impala Platinum, has said that it intends to proceed with its Ngezi Phase 2 expansion project, and redevelop the Bimha mine. As of the end of 2015, some $450 million had been committed to this project, which remains on schedule for completion in 2016. A further $12.2 million will be spent on the refurbishment of the Selous Metallurgical Complex base metal refinery project, of which $9.9 million has so far been committed. Bimha is due to reach full production in 2018, and last year Zimplats also announced its intention to set up of an acid plant to reduce SO_2 emissiosn from the smelter and improve the company's environmental impact and compliance. Imapala corporate relations chief Johan Theron told local press: "the previously talked about acid plant remains on the table, but has to be considered with the work we are doing in consultation with the government of Zimbabwe and Zimbabwean PGM industry to possibly expand our smelting facilities or re-commission the base metals refinery, as the design and sequencing of an acid plant is ultimately dependent on these projects."

MOROCCO

King Mohammed inaugurates new phosphate plant

Morocco's King Mohammed VI has inaugurated another part of the Jorf Lasfar phosphate complex as state phsopahte producer Office Cherifien de Phosphate (OCP) continues its ambitious construciton programme (see article elsewhere this issue). King Mohammed also officially opened a new seawater desalination plant at Jorf Lasfar. The new phosphate complex, compelted at a cost of 5.3 billion dirhams ($540 million), compreises a 1.4 million t/a sulphuric acid plant, a 450,000 t/a phosphoric acid plant, 1.0 million t/a of diammonium phosphate (DAP) capacity, a 62-MW solar power station, and up to 200,000 tonnes of fertilizer storage infrastructure.

UGANDA

Financial closure for phosphate project

China's Guangzhou DongSong Energy Group Company, which is developing Uganda's Sukulu phosphate project, says that it has signed a $240 million financial closure agreement with the Industrial and Commercial Bank of China (ICBC) to develop the project. The financial closure makes the phosphate project one of the largest privately-funded mining sector investments in Uganda. Guangzhou DongSong plans to establish a mine and a beneficiation plant with annual capacity of 2.0 million t/a of phosphate concentrate, and a 300,000 t/a phosphate fertilizer plant. The complex will also include a 400,000 t/a sulphuric acid plant, a 12 MW waste heat based power generation plant, and a 300,000 t/a steel mill.

NAMIBIA

Gecko awaits environmental report

Gecko Namibia, which has ambitious plans to develop phosphate and uranium mining in western Namibia, is still awaiting the results of an environmental study on the impact of offshore phosphate mining. The company is aiming to establish an industrial park near Swakopmund but has not yet signed a lease agreement on the 700 hectare site, which is slated to include port facilities, a desalination plant, a sulphuric acid plant, a soda ash and bicarbonate of soda plant and caustic soda and phosphoric acid plants. Gecko would import sulphur to run its acid plant, the acid from which would then be used in phosphate processing and uranium nining, while the soda ash and caustic soda would be used for local alkaline leach uranium projects operated by Langer Heinrich Uranium. However, the project has been put back by the government moratorium on seabed phosphate mining imposed at the end of 2013, initially for 18 months, and now continuing until an environmental impact study is completed.

Other companies are also eagerly awaiting the outcome of the study, including New Zealand's Chatham Rock Phosphates and Namibia Marine Phosphates, and recently Israel's Lev Leviev Namibia Phosphate indicated that it will go ahead with plans to set up a plant at Làderitz to test the feasibility of mining marine phosphates, and has signed a memorandum of understanding with Israel Chemicals Ltd (ICL) to set up a phosphate fertilizer manufacturing plant in Namibia.

DEMOCRATIC REPUBLIC OF CONGO

Acid plant nears completion

Freeport McMoRan, the majority shareholder in Tenke Fungurume Mining SA, says that construction of a second sulphuric acid plant is "substantially complete" at its Tenke mine second phase expansion, and is on-course for start-up during 2016. The expansion will raise copper production by 68,000 t/a to approximately 200,000 t/a at the site. Tenke Fungurume is 56% owned by Freeport McMoRan Copper & Gold Inc., 24% by Lundin Mining Corp. and 20% by DRC's mining company La Générale des Carriéres et des Mines (Gécamines).

FINLAND

Acid from pulp production

Finland's Metsä Group is developing what it calls a 'bioproduct mill' at Äänekoski, which will include novel processing technologies, including acid production from biological off-gases. The new process plant is being constructed at of one of the world's largest pulp mills, with an existing annual capacity of 1.3 million t/a. It will include a bark gasification plant, which will replace some 45,000 cubic metres of heavy fuel oil use per year. The gasification plant will begin production at the same time as the bioproduct mill, in the third quarter of 2017. Agreements have been signed with biogas manufacturer Eco-Energy SF Oy and biocomposite producer Aqvacomp Oy as part of the bioproduct plant; the biogas plant that EcoEnergy will build will use the sludge generated in pulp production to produce approximately 20 gigawatt hours of biogas a year, and some of the sulphur-containing gases released will be converted into sulphuric acid, to be used as a raw material by the mill instead of sulphuric acid bought from the market.

SENEGAL

New partnerships in phosphate sector

Senegal's phosphate sector is seeing new partnerships after two deals involving foreign partners. The country's own Mimran Natural Resources is now in partnership with Australian mining junior Minemakers, hoping to develop the Baobab phosphate project in Senegal, and Mimran has also very recently taken a 45% stake in the African Investment Group SA (AFRIG), in which Polish fertilizer producer Grupa Azoty took a controlling interest in 2013. AFRIG owns some of the most important phosphate and heavy mineral deposits in Senegal, and has assured phosphate rock supply to Grupa Azoty's European fertilizer manufacturing operations. It is also developing a new phosphoric acid plant which would be only the second in Senegal, after the existing plant owned by Industries Chimiques du Senegal (ICS), a subsidiary of Indonesia's Indorama Corporation.

NEW ZEALAND

Chatham seeks to recover expenses

The Office of the Ombudsman in New Zealand will investigate expenses charged by the country's Environmental Protection Authority (EPA) to Chatham Rock Phosphate Ltd. The costs were associated with a declined marine consent application filed by Chatham for offshore phosphate mining. Chatham says that it is contesting NZ$800,000 (US$530,000) in invoiced costs out of a total of NZ$2.7 million (US$1.7 million) charged by the EPA.

"The requested scope of the investigation was first to examine the EPA's costs recovery practices for its marine consent process – including the withholding of information from CRP which was relevant to whether some of the charges were authorised by law," said Chris Castle, CEO of Chatham.

ZAMBIA

Pressure to re-start nickel mine

The Zambian government is pressuring the owners of the Munali nickel mine to re-start operations or face having the mine repossessed. Operations were suspended at Munali in 2011 due to low nickel prices. The mine has been bought by UK-based Consolidated Nickel Mines, who say that it will cost around $60 million to rehabilitate and upgrade the processing facilities to produce around 3-5,000 t/a of nickel over a 7-10 year mine life, as well as 400 t/a of copper, plus some cobalt and platinum group metals. The upgrade programme includes a dense media separation plant, a solvent extraction/electrowinning plant, and a leaching and solar facility. ■

People

Walter Precourt will become senior vice president, phosphates operations at The Mosaic Company from 1 June. He is currently senior vice president, potash operations, and will move from Saskatchewan to Central Florida to take up his new role. Precourt joined Mosaic in 2009 as vice president for environment, health and safety (EHS) before becoming a senior vice president in 2012.

"Walt has provided remarkable leadership for EHS and Potash," said Joc O'Rourke, president and chief executive officer of Mosaic. "With more than two decades of operational and functional leadership roles at The Dow Chemical Company and Holcim, Walt is uniquely qualified to lead the next phase of our company's phosphates journey."

At the same time, **Bruce Bodine**, who currently leads Mosaic's supply chain, will become vice president for potash on 1 April. Bodine will move to Regina, Saskatchewan, in June to head-up Mosaic's potash business. He has held leadership positions in potash, phosphates and supply chain during his 15 years at Mosaic.

"Bruce is an insightful and engaging leader of both people and process, and he is driven to help Mosaic compete in the potash market," O'Rourke said.

Gary 'Bo' Davis, after six years as Mosaic's senior vice president of phosphate operations, will take on a senior advisor role, prior to his planned retirement in January next year. Davis has worked at Mosaic since the company's formation in 2004 and has 41 years experience in the industry. He has been Mosaic's senior vice president, phosphate operations since 2009. He also served on the board of the Wa'ad Al Shamal Phosphate Company, the joint venture between Mosaic and Ma'aden.

"Bo led the transformation of our phosphates business unit and achieved levels of operational excellence not previously reached in the phosphates industry," O'Rourke said. "His strategic contributions have led Mosaic's phosphate business to be the largest and strongest in the world, and we thank him for his years of dedicated service."

Rio Tinto has announced several board changes. **Richard Goodmanson**, non-executive director of the company, who joined the board in December 2004, will be retiring from the board this year. He will not seek re-election as a non-executive director of Rio Tinto plc and Rio Tinto Limited, according to the company, and will retire from the board at the conclusion of the Rio Tinto Limited annual general meeting in Brisbane on 5 May 2016. **Megan Clark** will be appointed as chairman of the Sustainability Committee upon Richard Goodmanson's retirement on 5 May 2016, and will also become a member of the Remuneration Committee with effect from 1 May 2016.

Rio Tinto chairman Jan du Plessis said "I am very grateful to Richard for his considerable contribution to Rio Tinto over many years. He provided tremendous support during his tenure, notably as chairman of the Sustainability Committee. I wish him well for the future."

Royal Dutch Shell (Shell) says that Unconventional Resources Director and US Country Chair **Marvin Odum**, will leave the company at the end of March, 2016. At the same time, the Athabasca Oil Sands Project and the Scotford Upgrader in Canada will join the global Downstream organisation under Downstream Director **John Abbott**; and the Shale Resources business will join the global Upstream organisation under Upstream Director **Andy Brown**. As a result of these changes, The Unconventional Resources Directorate will cease to exist.

Since joining Shell as an engineer in 1982, Marvin has held a number of commercial and technical leadership roles of increasing responsibility. He has held the position of U.S. Country Chair and President of Shell Oil Company since 2008, and joined Royal Dutch Shell's Executive Committee as Upstream Americas Director in July 2009. Royal Dutch Shell Chief Executive, Ben van Beurden commented: "Marvin has had a long and distinguished Shell career and I'm grateful to him for the future central role he's played in the company's success. He leaves our important businesses in the Americas well positioned for the next phase of their development."

Calendar 2016

MARCH

13-15

Phosphates 2016, PARIS, France
Contact: CRU Events
Chancery House, 53-64 Chancery Lane,
London WC2A 1QS, UK.
Tel: +44 20 7903 2167
Email: conferences@crugroup.com

20-22

AFPM Annual Meeting,
DALLAS, Texas, USA
Contact: Yvette Brooks
Email: ybrooks@afpm.org
Web: www.afpm.org

20-24

SOGAT 2015, ABU DHABI, UAE
Contact: Dr Nick Coles, Dome Exhibitions
Tel: +971 2 674 4040
Email: nick@domeexhibitions.com

APRIL

11-13

TSI World Sulphur Symposium,
VANCOUVER, Canada. Tel: +1 202 331 9660
Email: sulphur@sulphurinstitute.org
Web: www.tsi.org

MAY

22-26

2nd Annual Brimstone Sulphur Symposium,
ABU DHABI, UAE
Contact: Brimstone STS Ltd
Tel: +1 909 597-3249 Fax: +1 909 597-4839
Email: mike.anderson@brimstone-sts.com

30 - 1 JUNE

84th IFA Annual Conference 2016,
MOSCOW, Russia
Contact: IFA Conference Service
Tel: +33 1 53 93 05 25
Email: conference@fertilizer.org
Web: www.fertilizer.org

JUNE

10-11

40th AIChE Annual Clearwater
Conference 2016,
CLEARWATER, Florida, USA
Email: chair@aiche-cf.org
Web: www.aiche-cf.org

NOVEMBER

7-10

Sulphur 2016 Conference and Exhibition,
LONDON, UK
Contact: CRU Events
Chancery House,
53-64 Chancery Lane,
London WC2A 1QS, UK.
Tel: +44 20 7903 2167
Email: conferences@crugroup.com

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Morocco's phosphate boom

Morocco is in the middle of a $16 billion investment programme which will see the country double its phosphate rock and triple its processed phosphate output, with a corresponding increase in sulphur consumption.

Fig 1: **Morocco's phosphate industry**

The segment of sulphuric acid demand represented by phosphates is around 55%, most of that – around 90% – for fertilizer use, the remainder being consumed in various industrial uses, mostly in the food and animal feed industries as well as detergents, cleaners, metal finishing, toothpaste and many others. As the largest segment of demand, sulphuric acid and by extension sulphur demand continue to be driven mainly by the phosphate fertilizer market. And in the world of phosphate, while China is the largest producer, Morocco holds by far the world's largest phosphate reserves – at around 50 billion tonnes P_2O_5, the country claims around 75% of all of the world's phosphates, according to the IFDC's re-rating of world phosphate reserves in 2010.

Although its mining and processing of phosphates is dwarfed by China and smaller than that of the United States, the lack of domestic demand for phosphate has left Morocco as the world's largest exporter of phosphates, while its lack of natural gas, refining or metal smelting capacity means that it also has little domestic sulphur or sulphuric acid production, and hence is also the world's largest importer of sulphur. Its importance to the future of the phosphate and sulphur markets is thus a key one.

OCP

Morocco's phosphate industry is controlled via Office Cherefien des Phosphates (OCP), a company 95% owned by the Moroccan state. OCP has been in the phosphate business for over 100 years, with a history extending back into the French colonial era. It is the largest and richest company in Morocco, given that phosphates represent one of the most important sectors of the economy, employing 23,000 people and with revenues in 2014 of $4.2 billion and earnings of $920 million, accounting for one quarter of Morocco's exports and about 3.5% of its GDP. Indeed, OCP is the world's largest single phosphate producing company, and lays claim to 30% of the international merchant market for rock phosphate and 47% of that for phosphoric acid. However, the company's market share in processed phosphates (mono- and di-ammonium phosphate – MAP/DAP – and triple superphosphate – TSP) has lagged behind this, increasing from 15% in 2013 to only around 17% of the international market in 2014, and while OCP has announced significant increases in phosphate mining as part of its strategic development programme, the real aim behind the country's current massive phosphate expansion is to expand its position in downstream phosphates.

Upstream, the company operates three main mining sites; Khourgiba and Gantour, in the north of the country, and at Boucraa in the south (see Figure 1). Prior to the current expansion beginning, capacity at Khourgiba, where three mines are operational, was 19.5 million t/a of rock. There are six mines at Gantour, split between two sub-sites at Benguerir and Youssoufia, with a combined capacity of 7.6 million t/a. Boucraa has one mine, with 3 million t/a of capacity. Reserves are overwhelmingly concentrated in the northern sites, with 53% at Khourgiba and 45% at Gantour – only 3% of reserves are at Boucraa.

Expansion plans

Because of this concentration of resources, as part of its development plan, OCP intends to open three large new mines on the Khouribga site, increasing the site's annual production from 19.5 million t/a to 38 million t/a, with new mines at El Halassa, Ouled Fares and

the so-called North Central Extension Zone. A fourth mine, Benguerir Sud, is to be opened at Gantour. However, in the longer term, OCP is also aiming to exploit a new despoit, Meskala, further south. The Meskala deposit represents about 15% of Morocco's reserves and is the largest new deposit to be developed since the 1970s. Overall, the first phase of OCP's $16 billion expansion programme will boost Moroccan rock output by 10 million t/a from 2008-2016, and the second phase another 12 million t/a to 2020-22.

As part of the production increase, OCP has been increasing the capacity of its washing plants at Khourgiba, and following the completion of the first phase in 2013, a second phase is now under development with a completion date set for 2016 to increase capacity to 12 million t/a, and tie the beneficiation plant into the Khouribga-Jorf Lasfar pipeline. This 187 km pipeline has been one of the key components of the expansion plan, aiming to drastically reduce the cost of phosphate production by cutting the transport cost from mine to processing site. The pipeline is purely gravity driven – an environmental bonus as well – it begins 650 m above sea level, and transports a phosphate slurry consisting of all of the phosphate mined and washed at Khouribga to the Jorf Lasfar chemical plants for processing into phosphoric acid and DAP, or to the port of Jorf Lasfar for export. The pipeline was officially inaugurated in October 2014, and has a total capacity of 38 million t/a of phosphate slurry, sufficient to cope with the entire projected expansion of the Khourgiba mines.

In addition to this main pipeline, a further slurry pipeline is now planned to run between Gantour and Safi to take another 10 million t/a of beneficiated phosphate, and when the Meskala deposit is developed, that too will get a slurry pipeline to carry beneficiated rock from the mine site the 95km to Safi.

Downstream expansions include expanding port capacity at Safi for shipping phosphates, and also expanding capacity at the port of Jorf Lasfar to cater for additional volumes when the four new fertilizer plants are all on stream. Export capacity at Jorf Lasfar will rise to 10.5 million t/a of phosphate rock, 10.8 million t/a of fertilizer and 2.6 million t/a of phosphoric acid. Improvements at Safi include new sulphur shiploading equipment with a capacity of 4,000 t/h and 200,000 tonnes of additional sulphur storage, plus 4.0 million t/a of sulphur melting capacity. Solid sulphur import capacity will rise from 2.3 million t/a to 7.6 million t/a. Port capacity at Safi is being tripled as part of the expansion programme.

Table 1: **World phosphoric acid production, 2014** (million tonnes product)

China	33.3
USA	14.1
Morocco	8.4
Russia	4.7
India	2.7
EU	2.4
Brazil	2.1
Saudi Arabia	2.0
Tunisia	1.9
Jordan	1.2
Israel	1.0
Others	6.2
Total	**80.0**

Source: IFA

Table 2: **Phosphoric acid capacity increases, 2014-2019** (million t/a P_2O_5)

China	+2.2
Morocco	+3.6
USA	-0.8
Saudi Arabia	+1.5
Brazil	+1.1
India	0
Tunisia	+0.1
Jordan	+0.1
Indonesia	+0.2
World	+7.8

Jorf Lasfar

At the heart of OCP's ambitious investment programme is its flagship project, the Jorf Lasfar Phosphate Hub (JLPH), from which OCP hopes to conquer the processed phosphates market. Fertilizer production capacity at the Jorf Lasfar complex stood at 3 million t/a prior to the expansion, but it is slated to rise to 10 million t/a by 2020. OCP is working closely with Jacobs Engineering on the Jorf Lasfar Hub, via a joint venture company formed in 2010, Jacobs Engineering SA (JESA), 50-50 owned by Jacobs and OCP. The expansion will be in two phases, the first, due to be complete by the end of this year, will build four new DAP/MAP complexes, each with a capacity of just over 1.0 million t/a, and each consuming 500,000 t/a sulphur each to feed sulphuric acid and phosphoric acid capacity. In total, MAP/DAP capacity at the site will ruse to just under 8.0 million t/a, including the 375,000 t/a of the Bunge Maroc Phosphore joint venture at the site, developed initially with US-based Bunge fertilizers. OCP bought Bunge's 50% stake in that plant in 2013.

Work is well advanced on the first phase of the Jorf Lasfar hub. The first plant came on-stream in 2Q 2015, and the second towards the end of the year. The final two plants are set to come onstream in 2Q and 4Q 2016 respectvely. After this, a second, even more ambitious phase will begin. This will involve a further six MAP/DAP plants at the Jorf Lasfar site, all to the same design, with target completion date around 2020-22.

There are also plans for boosting phosphate capacity at Safi and ultimately at Laayoune in the south (to where rock is transported from Boucraa).

Financing

The overall cost of the development programme is put at $15.8 billion, of which OCP spent an estimated $3.6 billion during 2013, and another $3.5 billion for 2014. There have been questions as to how OCP will finance this huge investment, especially at a time when global phosphate prices are at their lowest levels for several years; the company posted a 12% fall in profits during 2014 as compared to 2013 because of falling prices, although the figures for the first three quarters of 2015 (full year figures are not yet available) show an increase in earnings from $3.7 billion to $3.8 billion for the period, and, more importantly, an increasing in net earnings from $1.0 billion to $1.4 billion, with margins increasing from 28% to 37%.

OCP has also taken some foreign loans to help provide additional finance, signing a $150 million loan agreement in 2013 with the Saudi Arabian Islamic Development Bank to finance infrastructure upgrades at the port of Jorf Lasfar, and in 2014 it signed a $270 million loan agreement with German state development bank KfW. For a while there were suggestions of a while or partial IPO of the company to generate cash, but the government moved to deny these rumours in October last year. The company has been active on foreign bond markets, however, with a $1.55 billion dollar bond issue in 2014, and a further $1 billion issue in April 2015. Ratings agency Fitch rates OCP bonds as BBB-; essentially the same as that of the Moroccan govern-

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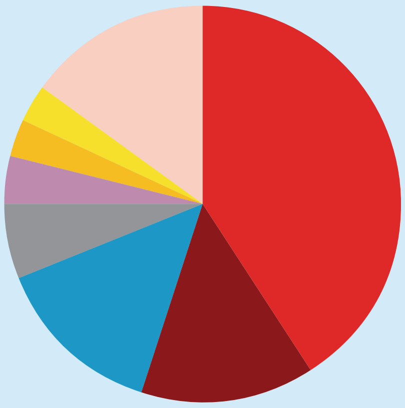
Fig 2: **Phosphate rock production, 2014**



Total: 197 million t/a (tonnes product)

Source: IFA

ment, and for the time being it seems that the company is comfortably able to keep the investment cash flowing.

Market share

One of the major trends in the phosphate industry over the past few decades has been towards greater downstream integration between phosphate rock production and phosphate processing, and OCP is clearly moving in this direction as well. Aside from India, which buys phosphate rock and phosphoric acid to run its own domestic phosphate industry, most countries prefer to buy processed phosphates. As noted above, OCP's expansion is focused squarely on integrating produciton in this way. Not only is it hoping to expand its downstream phosphate base, it is also aiming to reduce costs via integration, and improving that integration through its new slurry pipelines. OCP already has what are widely regarded as some of the lowest prices for phosphate rock production (with the exception of Boucraa, where the geology and logistics are more difficult), and clearly hopes to be at the bottom of the cost curve for MAP/DAP production as well with its large, integrated phosphate complexes. his comes at a time when US production is in relative decline due to mature deposits and falling ore grades, which the strong dollar has exacerbated this. Indeed, as Figure 2 shows, in 2014 Morocco actually just edged ahead of the US in terms of phosphate rock production.

Nevertheless, OCP faces a challenge in moving over the next few years from 17% of the traded processed phosphate market to a projected 28% by 2022, according to CRU. Though previously focused on American and Asian markets, OCP has been actively trying to move into African markets. It announced a partnership in 2014 with Gabon's Société Equatoriale des Mines to build two factories in Gabon, with $2 billion allocated for the projects, and joint output aiming to reach 2 million t/a by 2018. Earlier this month it also announced the formation of a new company: OCP Africa. OCP says that "this new entity aims to contribute to meeting the challenge of creating structured, efficient and sustainable agriculture on the continent of Africa, by providing agricultural producers with all the resources they need in order to succeed: suitable, affordable products, services and partnerships, logistics and financial solutions." It plans to open about fifteen subsidiaries in Africa over the coming months.

Sulphur demand

In spite of the scope of its amibitions then, and the current cyclical depression in global commodity markets, OCP looks set to realise the first phase of its planned mega-development according to the revised schedule, and at the moment the omens are also looking good for the second phase, running to 2022. As noted earlier, because the size of its phosphate industry and lack of any correpsonding oil/gas/metal industry, Morocco has already become the world's largest sulphur importer, importing 4 million tonnes in 2014, according to IFA figures. With the new sulphuric acid plants coming onstream in 2015 and 2016, Morocco's sulphur requirements are expected to reach 5 million t/a by 2017, and as the second phase ramps up, this is set to rise to 6 million t/a by 2020 and potentially 8 million t/a by 2022. At a time when world sulphur production is seeing large new incremental production from countries like Abu Dhabi, China and Kazakhstan, this can only be good news for sulphur producers.

Iran and the Gulf

The easing of sanctions on Iran offers the chance for the increasing volumes of sulphur from the South Pars project to find a wider market.

Above: A view across the Gulf: the South Pars project, Iran.

In January, the UN International Atomic Energy Agency (IAEA) confirmed that Tehran had fulfilled its obligations under an agreement to limit its nuclear enrichment programme. This was the trigger for the progressive easing of international sanctions against the country which have steadily ramped up from 2006 as concerns over the country's nuclear ambitions have grown. The UN had mandated an arms embargo and asset freeze, and had added travel bans on individuals and sanctions on the provision of financial services. The US and EU, as well as some other states, had also imposed further sanctions on oil sales and oil equipment, and the US in particular had not only mandated sanctions on all companies doing business with Iran, it had also caused some difficulties by trying to extend those to companies outside the US that dealt with Iran that had US subsidiaries or operations.

Unravelling this complex web of international agreements and sanctions will take some time, but the thawing in US-Iranian relations begun by the replacement of more hawkish president Ahmedinejad with more moderate president Rouhani seems to be finally bearing serious fruit. Iran's central bank says lifting banking sanctions will allow $30 billion of foreign reserves currently frozen in accounts around the world to be brought back, and the Iran's GDP is expected to see a boost to 5% in 2016-17 from an estimated 0% in 2015.

Oil production

One of the first areas where the effects have been seen has been on Iran's oil production and exports. Prior to the Iranian Revolution of 1979, Iran was one of the world's largest oil producers, with output running at over 6 million bbl/d. This fell to 1.5 million bbl/d in the early 1980s, but recovered to over 4 million bbl/d during the 2000s. The sanctions regime saw this fall from 4.3 million bbl/d in 2011 to as low as 2.6 million bbl/d in 2013, however, and only recovered to 2.8 million bbl/d in 2014 and 2015.

There is clearly massive potential, though. The country's reserves were put at 157 billion barrels at the end of 2014, making it the fourth largest holder of reserves after Saudi Arabia, Venezuela and Canada, and just ahead of Iraq, though much of Venezuela and Canada's reserves are held as oil sands.

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Iran was reported in January to have 47 million barrels of oil sitting in tankers offshore, and it was believed that it would release this onto the market at around 500,000 bbl/d initially, slowing as Iranian production ramps up by 500,000 bbl/d over six months, with around 100,000 bbl/d likely in the first month. The net result would be that, between stored oil and production increases, Iran would in effect be able to add 500,000 bbl/d to the market immediately. The country wishes to increased production by 1 million bbl/d to 3.8 million bbl/d by the end of 2016 and further still in the longer term, but it is unclear the effect that years of sanctions have had on oilfield maintenance and how much investment may be required to achieve this.

Iranian oil exports were at 2.5 million bbl/d in 1979, but had fallen to 1.4 million bbl/d in 2014, and as low as 1.2 million bbl/d in 2015 – sanctions actually capped them at 1.1 million bbl/d, but there was some evasion of this regime. The additional volumes from storage and reinstated production should take this to just under 2 million bbl/d by the end of 2016. This has of course not been good news for an oil market already oversupplied due to increased US shale oil production, a fall in Chinese demand and the reluctance of Saudi Arabia to play swing supplier this time. Iran has been selling at a discount to market rates in order to gain back market share. Costs of production are estimated at only $12/bbl in Iran, giving it a reasonable margin even at current oil prices.

Gas production

Iran actually has the largest natural gas reserves in the world, put at 34 trillion cubic metres in 2014, and representing around 18% of the world's gas. Most of this is held in the huge offshore South Pars Field (which extends across the maritime border with Qatar, where it is called the North Field). Development of South Pars has proceeded in spite of the various difficulties incurred by sanctions, with around 28 phases projected for the overall development of the field and associated onshore facilities, including gas and condensate production and downstream petrochemical works. Phases 1-16 are now operational, and the pace of development has begun to accelerate as the prospects of sanctions being eased has made access to international markets more likely. Phases 17 and 18 are producing gas and due to be completed this year, and the gas processing/sweetening sections of Phases 20-21 are also expected on-stream soon, initially processing sour gas from Phases 6-8. Iran says that all of the gas phases should be up and running by the end of 2017 and the oil producing sections by 2018.

Fig 1: **Iran's sulphur producers**

New investment

With much oil and gas around the world held by states and state-owned companies, International Oil Companies have been excited at the prospect of the possible opening up of Iran's upstream sector to foreign investment. The Iranian government has been keen to attract investment to achieve its long term target of reaching 5.7 million bbl/d of production, and consequently has hinted at relatively generous terms. An initial foreign investment of around $25 billion is targeted, and several leading E&P companies such as BP, Eni, Repsol, Shell, Statoil and Total are believed to have been in discussions. An auction of 50 oil and gas projects and 18 E&P blocks is expected to be held in May, and will provide IOCs with an opportunity to access lower cost oil prospects and help re-balance their portfolios and books and remain competitive in a low oil price environment. The licensing round will include onshore and offshore, as well as early and late stage projects, with varying degrees of complexity.

As well as upstream investments, there are many plans for gas export pipelines, and Iran would also love to develop LNG export facilities, as its options for export of natural gas are currently limited, one reason perhaps why it has focused on downstream urea and methanol production for export instead.

Caveats

However, enthusiasm has been tempered with caution on various fronts. Firstly on a purely technical level, a number of key sanctions still remain. The US is lifting its so-called "secondary sanctions" - the ones that apply to non-US individuals or companies, but US "primary sanctions" will still

Table 1: **Iran's major sulphur producers**

Producer	Sulphur recovery units	Capacity t/d	Sulphur forming/ granulation units
Petrochemical complexes			
Razi Petrochemical Co	Claus	2 x 630 t/d	2 x 800 t/d
	Claus	1 x 700 t/d	
Kharg Petrochemical Co	Claus	515 t/d	-
Gas refineries			
South Pars gas refinery Phase 1		606 t/d	-
Phases 2 & 3, 4 & 5, 9 & 10		3 x 400 t/d	-
Phases 15-18		4 x 400 t/d	-
Khangiran (Shahid Hasheminejad Gas Refinery Co)	Claus	4 x 650 t/d	6 x 750 t/d (pastillation units)
Ilam Gas Refinery Co	Claus	2 x 173 t/d	2 x 300 t/d (Sandvik rotoform pastillation units)
Oil refineries			
Tehran Oil Refinery Co	Claus	100 t/d	-
Tabriz Oil Refinery Co	Claus	20 t/d	-
Bandar Abbas Oil Refinery Co	Claus	130 t/d	-
Isfahan Oil Refinery Co	Claus	103 t/d	-

Source: National Iranian Gas Company

ban US nationals and companies from engaging in business with Iran. On the financial side, dollar-based clearing restrictions remain in place, and US restrictions on dealing with the Republican Guard and its affiliates, including the Khatam al Anbia company, which is Iran's largest contractor in industrial and development projects, controlling over 800 subsidiaries including oilfield service companies. It will certainly be difficult if not impossible for US companies to deal with Iran, and project finance may need to be routed through countries with fewer restrictions, like China. These obstacles are not insuperable, but have the potential to cause delays and costs for investors.

Secondly, the political truce remains fragile and subject to shocks from either side. Iran faces presidential elections in 2017, and while Rouhani should win a second term provided that people are seeing the benefits of the sanctions regime being lifted, there are many conservative elements in Iran opposed to a deal with the west. Likewise the rhetoric from some Republican contenders for the US presidency, especially Donald Trump, has in effect threatened to tear up the existing deal. These political risks could loom large for investors in the short term. Iran is said to be inserting a clause in contracts that foreign companies will not be released from contractural obligations should the US, EU or UN re-impose sanctions.

Sulphur output

Iran produces sulphur from four refineries, at Tehran, Tabriz, Bandar Abbas, and Esfahan, as well as the Razi and Kharg petrochemical complexes, but most of the country's sulphur production has come from its natural gas processing. There are three sour gas processing complexes – at Khangiran (Hasheminajad) near Mashhad in the northeast of the country, at Ilam in the west near the Iraqi border, and at Assaluyeh, where the gas from South Pars is brought ashore, and Assaluyeh and Khangiran are the two largest of these. Figure 1 shows the location of these complexes and Table 1 their approximate capacities. Total output ran at about 1.7 million t/a from a capacity of 2.2 million t/a a few years, and has slowly increased as new phases of South Pars come on-stream.

> **"The country has suffered from a chronic shortage of refining capacity.**

Iranian sulphur consumption is of the order of a few hundred thousand tonnes per year, for sulphuric acid production – some for some small phosphate processing, as well as other industrial uses, but this runs behind production, so Iran has been a major sulphur exporter, of the order of 1.2 million t/a for the past few years.. Iran has exported this mainly (ca 75%) to China, which trade has not been affected to the same extent by sanctions, and to a lesser extent to India, but as sanctions are lifted the prospect becomes more open to exports elsewhere.

As well as extra production from South Pars, which could in theory take sulphur capacity to 3 million t/a, Iran says that it has plans to revamp its ageing oil refineries – the country has suffered from a chronic shortage of refining capacity and has actually often had to resort to importing gasoline even while it was exporting oil. These would obviously also be a major boost to sulphur capacity – Iran currently mainly produces gasoline to Euro-4 standards at Shazand Arak, Isfahan and Tabriz refineries, and Euro-3 at the others, but there is an upgrade in progress at Bandar Abbas and Lavan. NIOC puts Iranian refining capacity at 2.1 million bbl/d, but the government aims to increase this to 3.4 million bbl/d by 2025, and Iran is also looking at investing in refinery capacity overseas, in Spain and Brazil.

Uncertainty still remains

Iran has grand plans, which have moved a step closer with the beginning of the lifting of sanctions. While years of underinvestment will take some time to put right, the resources are certainly there – it is worth remembering that Iran was a bigger oil producer than Saudi Arabia prior to the revolution. There has been some justified scepticism about the pace of development that Iran seems keen to try and undertake, not least because of factors such as the availability of skilled engineers and fabricators, and according to companies that deal with Iran, payments still remain a fraught area at present – undoubtedly the greatest boost will come when Iran reconnects to the SWIFT international payments system. But provided that politics do not intervene once more (as they yet might), Iran's resurgence will add more oil, gas and sulphur output to the already considerable production of the Gulf.

Sulphur recovery project listing 2016

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.

Company	Site	Process	Capacity	Licensor	Contractor	Type of project	Start-up date
AZERBAIJAN							
SOCAR	Garadagh	H_2S, 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 t/d	WorleyParsons	n.a.	New	2017
BELGIUM							
ExxonMobil	Antwerp Refinery	SWS	n.a.	Fluor	Amec Foster Wheeler	New	2017
ExxonMobil	Antwerp Refinery	O_2 enrich, amine TGT	325 t/d	WorleyParsons	Amec Foster Wheeler	Revamp	2016
BRAZIL							
Petrobras	Premium I	SuperClaus	2 x 240 t/d	Jacobs	n.a.	New	2017
Petrobras	Premium II	SuperClaus	240 t/d	Jacobs	n.a.	New	2017
Petrobras	REDUC	SuperClaus	2 x 62 t/d	Jacobs	n.a.	Revamp	Cancelled
Petrobras	Maranhao Premium	12 x Claus, NH_3	238 t/d H_2/amine TGT	Amec Foster Wheeler	n.a.	New	2017

KEY
BTX = BTX destruction
Fuel = Fuel gas supplemental burning
O_2 = Oxygen enrichment
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TGT = Tail gas treatment unit
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47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

Company	Site	Application	Capacity	Licensor	Contractor	Type of project	Start-up date
CAMEROON							
SoNaRa	Limbe	SRU, SWS	17 t/d	Amec Foster Wheeler	KT Kinetics Technology	New	2017
CANADA							
Suncor Energy	Montreal	Claus, SCOT	2 x 100 t/d	Jacobs	n.a.	New	2016
CHINA							
CNPC/Chevron	Chuadongbei	Claus, SCOT	2 x 687 t/d	WorleyParsons	n.a.	New	2016
Inner Mongolia Manshi	Ordos	SRU	51 t/d	Jacobs	n.a.	New	2017
Jiutai Energy	Linyi, Shangdong	EuroClaus	32 t/d	Jacobs	n.a.	New	2017
Shijiazhuang Yingding Gases	Shijiazhuang, Hebei	EuroClaus	12 t/d	Jacobs	n.a.	New	2016
ECUADOR							
Petroecuador	Esmeraldas	Claus	50 t/d	Prosernat	n.a.	New	2016
EGYPT							
MIDOR	Alexandria	Claus	410 t/d	n.a.	n.a.	New	2018
FRANCE							
Total	Donges	Claus	n.a.	n.a.	n.a.	Revamp	2017
INDIA							
HPCL	Visakh Refinery	Claus, NH_3, amine	300 t/d	WorleyParsons	n.a.	New	2016
Reliance	Jamnagar	O_2, NH_3, amine TGT	4 x 1,300 t/d	WorleyParsons	n.a.	New	2016
INDONESIA							
PT Medco E&P	East Aceh	EuroClaus	48 t/d	Jacobs	n.a.	New	2019
Pertamina	Balongan	Claus, NH_3, H_2, Amine TGT	1,100 t/d	Amec Foster Wheeler	n.a.	New	n.a.
IRAQ							
Midland Ref. Co	Daura	Claus, TGT, Aquisulf	2 x 125 t/d	Lurgi	n.a.	New	2015
North Ref. Co	Kirkuk	Claus, TGT, Aquisulf	3 x 135 t/d	Lurgi	n.a.	New	2015
South Ref. Co	Maissan	Claus, TGT, Aquisulf	3 x 272 t/d	Siirtec Nigi, Lurgi	Lukoil	New	2015
Gazprom Neft	Basra	SRU, LT-SCOT, amine	2 x 157 t/d	Jacobs	n.a.	New	2016
Petrochina	Halfaya	Claus, amine	3 x 60 t/d	WorleyParsons	n.a.	New	2017
Lukoil	Yamana	Claus, SCOT	n.a.	WorleyParsons	n.a.	New	2017
ISRAEL							
Bazan	Haifa Refinery	O_2 enrich	3 x 140 t/d	WorleyParsons	n.a.	Revamp	2017
ITALY							
Total E&P	TempaRossa	Claus, TGT	80 t/d	Siirtec Nigi	Tecnimont	New	2016
KAZAKHSTAN							
Agip KCO	Kashagan	Claus + TGT	2 x 1,900 t/d	WorleyParsons	Black & Veatch, Petrofac	New	2017
Pavlodar Oil Chem	Pavlodar Refinery	Claus, TGTU	180 + 260 t/d	Siirtec Nigi	Rominserv, Technip	New	2015

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Company	Site	Application	Capacity	Licensor	Contractor	Type of project	Start-up date
KUWAIT							
Chevron	Wafra	Claus, amine	2 x 218 t/d	WorleyParsons	n.a.	New	2018
CGUP	Wafra	Claus, amine	2 x 400 t/d	WorleyParons	n.a.	New	2018
KNPC	Mina al Ahmadi	Amine	2 x 400 t/d	WorleyParsons	n.a.	New	2016
KNPC	Al Zour Refinery	Claus	1,500 t/d	Amec Foster Wheeler	n.a.	New	2019
MALAYSIA							
Petronas	Johor	SuperClaus	3 x 470 t/d	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 t/d	WorleyParsons	n.a.	New	2017
NIGERIA							
Dangote Oil	Lekki Refinery	SuperClaus	2 x 115 t/d	Jacobs	n.a.	New	2017
OMAN							
OOC	Duqm Refinery	NH_3, H_2/Amine TGT S degas	3 x 355 t/d	Fluor	n.a.	New	2015
PDO	Yibal Khuff Sudair	Claus + TGT	250 t/d	WorleyParsons	n.a.	New	2016
PERU							
Repsol	La Pampilla	2 x Claus, NH_3, O_2, H_2/amine, TGT	83 t/d	Amec Foster Wheeler	SAINC	New	2016
POLAND							
Grupa Lotos	Gdansk Refinery	O_2 enrich	2 x 72 t/d	WorleyParson	Tecnimont	Revamp	2017
QATAR							
Qatar Petroleum	Mesaieed	Sour gas, AGE, Claus+ TGT	310 t/d	Worley Parsons	Petrofac/ Black&Veatch Prosernat	Revamp	2016
RUSSIA							
Bashneft	Ufa	Amine, SWS	n.a.	Amec Foster Wheeler	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 t/d	WorleyParsons	n.a.	New	2017
Lukoil	Volgograd	NH_3, H_2/Amine TGT, D'GAASS	2 x 76 t/d	Fluor	n.a.	New	2015
Lukoil	Kstovo Refnery	Claus, TGTU	2 x 290 t/d	Siirtec Nigi	Tecnicas Reunidas	New	2015
Mariisky	Mari El Republic	SRU+TGT, amine	n.a.	Shell	Amec Foster Wheeler	New	n.a.
Orsknefteorg	Orsk	EuroClaus	2 x 99 t/d	Jacobs	n.a.	New	2017
OOO Ilskii NPZ	Krasnodar Krai	EuroClaus	86 t/d	Jacobs	n.a.	New	2016
Varino Refinery	Varino	Smartsulf	15 t/d	WorleyParsons	n.a.	New	2017

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Company	Site	Application	Capacity	Licensor	Contractor	Type of project	Start-up date
SAUDI ARABIA							
Luberef	Yanbu	SRU, SCOT	2 x 81 t/d	Jacobs	n.a.	New	2016
PetroRabigh	Rabigh	EuroClaus	292 t/d	Jacobs	n.a.	New	2019
Saudi Aramco	Fadhili				Petrofac	New	2019
SINGAPORE							
Singapore Refining	Singapore Refinery	O_2 enrich	145 t/d	WorleyParsons	n.a.	On hold	n.a.
SPAIN							
Petronor	Muskiz	EuroClaus	86 t/d	Jacobs	n/a/	Revamp	2018
TURKEY							
STRAS	Aliaga/Izmir	SRU+TGT, amine, SWS	463 t/d	Tecnimont KT	Amec Foster Wheeler	New	2017
Turkish Petroleum	Mansuriya	Claus, amine	230 t/d	WorleyParsons	n.a.	New	2018
TURKMENISTAN							
Turkmenbashi Oil	Turkmenbashi City	SuperClaus	25 t/d	Jacobs	Hyundai	New	2019
UNITED ARAB EMIRATES							
IPIC	Fujairah	SRU, SWS, amine TGT	330 t/d	Amec Foster Wheeler	n.a.	New	2018
UNITED STATES							
Hydrogen Energy California	Kern County, CA	O_2 enrich, NH_3/H_2, Amine TGT, D'GAASS	100 t/d	Fluor	n.a.	New	2017
Leucadia hold	Chicago, IL	Claus, TGT	2 x 215 t/d	Black & Veatch	n.a.	New	On
NCRA	McPherson, KS	D'GAASS	194 t/d	Fluor	n.a.	Revamp	n.a.
Sinclair Oil	Sinclair, WY	Claus	n.a.	Amec Foster Wheeler	Amec Foster Wheeler	Revamp	2016
UZBEKISTAN							
Lukoil	Bukhara, Karasul	SuperClaus, TGT	2 x 405 t/d	Jacobs	n.a.	New	2018
VENEZUELA							
PDVSA	El Palito	SRU/amine TGTU, SWS	250 t/d	Shell	Amec Foster Wheeler	New	2018
PDVSA	Monagas	Amine reg, SWS	54 t/d	Amec Foster Wheeler	Amec Foster Wheeler	New	2016
PDVSA	Puerto La Cruz	Claus, NH_3, amine	2 x 225 t/d	WorleyParsons	n.a.	New	2018

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Sulphur World Symposium 2016

The Sulphur Institute (TSI)'s Sulphur World Symposium 2016 will be held from April 11-13th, 2016 in Vancouver, Canada.

The Sulphur Institute is proud to host the Sulphur World Symposium 2016, an annual event that attracts sulphur industry leaders for two days of expert speakers, networking events, and industry tours. Over 150 delegates from over 30 countries are expected at this year's event. Whether it is a review of leading industry practices or an analysis of the sulphur supply and demand forecast, TSI's speakers incorporate information on the entire industry value chain.

www.SulphurInstitute.org/Symposium16

MAIN SESSION PRESENTATIONS

Global sulphur and sulphuric acid outlook

Meena Chauhan, Integer Research Ltd.

For many years the industry has been predicting a significant global surplus of sulphur as large scale oil and gas projects were expected to come to fruition, but this has yet to emerge. However, a long awaited gas based project in the Middle East was commissioned in 2015, following numerous delays, and sulphur exports increased significantly from the UAE as a result. New projects in the Middle East, Asia and FSU are expected to add increasing volumes to the export market in the years ahead and will likely lead to changes in sulphur trade routes as producers seek to place tonnage in new or existing markets.

The collapse of oil prices has led to questions over the sustainability of some high cost sulphur producers, leading to the cancellation or postponement of projects in some regions including North America. The sustained downward trend of global commodity market prices including copper, zinc and nickel has also shifted the outlook for sulphuric acid markets across both supply and demand. Meena Chauhan, Sulphur Research Manager at Integer Research will explore these issues and present an outlook for supply, demand and trade for sulphur and sulphuric acid and present a view on the potential for the sulphur market to move into oversupply.

Global phosphate supply and demand overview

Jeff Holzman, PotashCorp

Challenging macroeconomic conditions have impacted all fertilizer products over the past year. This presentation will review the major factors impacting the phosphate market and provide an outlook for supply and demand over the next five years. Factors addressed in the outlook for phosphate demand include agriculture market fundaments, government policy and global economic conditions. From a supply perspective, the presentation will highlight major proposed capacity additions by region and provide an outlook for regional changes in production and trade.

Mosaic's sulphur melter in Florida

Hermann Wittje, The Mosaic Company and Mark Gilbreath, Devco USA LLC

The driver and rationale for Mosaic's construction of a new state of the art melter in Florida will be reviewed. Mosaic's need for 4.3 million t/a of sulphur and the challenges to accumulate this large quantity of sulphur on a rateable basis steered it to the decision to move forward with this project.

Mosaic needed an efficient and environmentally friendly facility to match the high internal standards it has for its operations. We will also take a closer look at the layout and construction of the facility and review what makes the melter reliable, efficient and state of the art.

As the world turns: what's next for energy markets?

Josh McCall, BP North America Gas and Power

The global energy markets have shifted dramatically over the last 18 months. What's next for the oil, gas, and NGLs markets? What is the impact of the current price environment on upstream investment and drilling activity? Will North American supply growth and exports continue to impact the global energy markets? Or, will the current price environment ultimately cause a shift in investments and production trends?

Supply is only half the equation. Energy demand has generally responded well to lower prices. Does this denote a structural shift, or is it just a temporary reaction to low prices? Global economic conditions will be key in determining the future of energy demand. Understanding how the next year plays out is critical, as short term energy market dynamics could impact the market for years to come.

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Metals market/ leaching outlook

Peter Harrisson, CRU

Sulphur and sulphuric acid consumption for the production of non-ferrous metals (copper, nickel, uranium, zinc, cobalt) has increased significantly over recent years. The emergence of the new demand routes, particularly in the case of nickel, has provided rapid and significant demand growth in new geographical markets.

The base metal sector has failed to avoid the dramatic slump in the commodity price over the last year and in some cases have been the leading losers of value. As market prices have approached marginal cost the question of sustainability of the new, and sometimes technically challenging, leaching technologies has come into question. The announced cutbacks in the copper leaching sector are a key example of how new demand can quickly become old demand.

This presentation will set out the current metals market landscape and highlight the areas where sulphur and sulphuric acid consumption are at risk, along with a view of what happens beyond the bottom of the cycle. The presentation will provide an overview of the key drivers of the various industries and how this translates into sulphur and sulphuric acid demand around the global market.

Dry bulk freight market – a survival mode perspective for all participants

Brian Malone, MID-SHIP Group

The dry bulk freight market has entered 2016 at its lowest point in history. Not only are freight rates at historical lows, but the collapse in commodity prices and vessel fuel prices have further reduced voyage rate costs for charterers while creating significant distress for the vessel operators and uncertainty of performance on all sides of the market.

The question on the supply side in 2016 is will we see the forecasted additions to the fleet significantly reduced by low market levels and financial constraints. What percentage will slip back and or face outright cancellation? Demand growth is expected to slow and remains heavily reliant on China and India. Will the current oversupply of vessels continue and for how long? Or will scrapping and layup off set in sufficient enough numbers to allow more balance?

This presentation will assess the current market from the charterers perspective (are we at or near the bottom?), examine the mitigation of risk in these extreme conditions and aims to identify the elements required for an eventual freight market recovery.

Shah project update: from sand to sulphur

Patricia Wories, Enersul Inc.

In the overall picture of the oil and gas industry sulphur forming is a small component in a very large complex operation. However, if this component has an upset or becomes damaged, the ramifications to high volume gas and liquid producers like Al Hosn Gas in the Shah Field would be hugely impacted. The decision to develop and bring on stream one of the world's largest sulphur forming facilities to full production has taken many years to successfully engineer, deliver, construct and start all of the equipment components. This presentation covers this very large diverse undertaking from conception, design/engineering, construction to start-up operations. Enersul is proud of its involvement in this large undertaking and will share with you our experiences with AlHosn to make their vision a reality.

Global economic outlook

Richard Koss, International Monetary Fund

Supply chain management/safety session

The Sulphur Institute has increased programming this year to add an informative new presentation session providing delegates with operations and logistics-centric positions a unique opportunity to learn from expert speakers. Papers will include:

Navigating North American transportation regulations for sulphur/sulphuric acid

Harold H. Weber, The Sulphur Institute

This presentation will summarise steps to follow when determining how to ship sulphur/sulphuric acid within North America. Beginning with a brief overview of United Nations Recommendations on the Transport of Dangerous Goods, it will provide insight on navigating through U.S. Code of Federal Regulations, Title 49, which is in harmony with Transport Canada's "Clear Language" Regulations. It also will exemplify how changes to regulations can originate at any time and from many sources; why regulatory reciprocity of shipments between the U.S. and Canada is important, and summarize recent changes resulting from rail tank car accidents. This presentation also will include descriptions of where to find additional resources for "beyond the book" answers, such as special provisions located within the regulations and TSI's guidance documents prepared in cooperation with regulatory agencies and referenced within the regulations. In addition, this presentation will include a few examples of typical problems encountered during transportation that demonstrate TSI's liaison role as a representative of our collective sulphur and sulphuric acid industry.

Hydrogen sulphide – training personnel to work safely

Chuck Simpson, CSP, Epic Brokers

Hydrogen sulphide (H_2S) is one of the few chemicals that has few warning properties at dangerous concentrations and can lead to immediate incapacitation. One must be particularly aware of the hazards associated with H_2S whenever working where hydrogen sulphide is present or presumed present. After engineering controls, a well-executed training program is arguably the most effective way of protecting personnel against hydrogen sulphide exposure. For several years, The American National Standards Institute (ANSI) has maintained a training standard titled Accepted Practices for Hydrogen Sulfide (H_2S) Training Programs or ANSI Z-390. The latest revision of Z390 emphasises the importance of providing instruction specifically applicable to the job site and work activities. The standard makes clear that a single annual, generic training session may satisfy some of the knowledge requirements, but training is not complete until job site and work activity component is provided.

Private/public partnering during emergency response

Amy Blanton, Chemours and Joe McCann, CSX Transportation, Inc.

Effective emergency response planning demands persistence and consistent liaison and coordination among a large diversity of governmental agencies, response organizations, and shippers of hazardous materials. This session is a departure from the traditional presentation format. Subject matter experts will provide detailed information on emergency response to accidental sulphur and sulphuric acid releases during transport. Topics addressed throughout the presentation will include shipper's responsibilities, actions of the first responders, independent agency interface during the incident, and the issues local or regional governments may examine following the incident. ■

Sulphuric acid project listing 2016

Sulphur's annual survey of recent and planned construction projects in the sulphuric acid industry includes several large-scale acid plants both for phosphate processing and to capture sulphur dioxide from smelters.

Company	Site	Application	Capacity	Licensor	Contractor	Type of project	Start-up date
AUSTRALIA							
Nyrstar	Port Pirie	Smelter off-gas	n.a.	Outotec	Outotec	Revamp	2017
CANADA							
Vale	Sudbury	Nickel smelter off gas	1,400 t/d	Jacobs	Jacobs	New	2015
CHINA							
Henan Zhongyuan	n.a.	Smelter off-gas	4,820 t/d	MECS	n.a.	New	2015
Hubei Huaqiang	n.a.	Sulphur burning	667 t/d	MECS	n.a.	New	2015
CHILE							
Codelco	Mejilones	Smelter off-gas	100 t/d	n.a.	Foster Wheeler	New	2015
Codelco	Potrerillos	Smelter off-gas	n.a.	Outotec	Outotec	Revamp	2018
FINLAND							
Boliden	Harjavalta	Smelter off-gas	n.a.	n.a.	n.a.	New	2019
GERMANY							
Grillo Werke	Duisberg	Spent acid regeneration	+120 t/d	Grillo Werke	n.a.	Revamp	2015
INDIA							
FACT	Kochi	Sulphur burning	2,000 t/d	n.a.	n.a.	New	2016
Paradeep Phosphates	Paradeep	Sulphur burning	2,000 t/d	MECS	Jacobs	New	2016
KAZAKHSTAN							
Kazatomprom	Stepanogorsk	Sulphur burning	450 t/d	MECS	Desmet Ballestra	New	2015
MOROCCO							
OCP	Jorf Lasfar	Sulphur burning	4,200 t/d	MECS	n.a.	New	2017
NAMIBIA							
Namibia Custom Smelter	Tsumeb	Smelter off-gas	1,000 t/d	Outotec	Outotec	New	2016
PERU							
SCC	Tia Maria	Smelter off-gas	1,640 t/d	Outotec	n.a.	New	2017
Votorantim	Cajamarquilla	Smelter off-gas	n.a.	Outotec	Outotec	New	2016
RUSSIA							
Ural Mining	Svyatogot	Smelter off-gas	n.a.	Outotec	Outotec	Revamp	2018
Norilsk	Nadezhda	Smelter off-gas	n.a.	Outotec	Outotec	Revamp	2015
SAUDI ARABIA							
Ma'aden	Umm Wual	Sulphur burning	3 x 5,050 t/d	MECS	SNC Lavalin	New	2016
SERBIA							
RTB Bor	Bor	Smelter off-gas	1,820 t/d	MECS	SNC Lavalin	New	2015

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 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

Company	Site	Application	Capacity	Licensor	Contractor	Type of project	Start-up date
SOUTH KOREA							
Korea Western Power	Taean	WSA	96 t/d	Haldor Topsoe	n.a.	New	2015
TUNISIA							
Groupe Chimique Tunisien	Gafsa	Sulphur burning	1,800 t/d	MECS	n.a.	New	2017
TURKEY							
Cengiz Group	Samsun	Smelter off-gas	n.a.	Outotec	n.a.	Expansion	2016
Toros Tarim	n.a.	Sulphur burning	2,200 t/d	MECS/ Ballestra	n.a.	New	2015
TURKMENISTAN							
Turkmenchimia	n.a.	Sulphur burning	1,500 t/d	MECS	n.a.	New	2016
UGANDA							
Sukuru Phosphate	Tororo	Sulphur burning	600 t/d	n.a.	n.a.	New	2016
UNITED STATES							
Mississippi Power	Kemper, MS	Gasification	400 t/d	n.a.	n.a.	New	2016
Freeport McMoRan	Miami, AZ	Smelter off-gas	n.a.	n.a.	n.a.	On hold	2017
UZBEKISTAN							
Ammophos-Maxam	Almalyk	Smelter off-gas	2,000 t/d	n.a.	n.a.	New	2018
Navoi Mining	Uchkuduk	Sulphur burning	2,000 t/d	n.a.	n.a.	New	2019
ZAMBIA							
Kansanshi Mining	Solwezi	Smelter off-gas	4,400 t/d	Outotec	Outotec	New	2015

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 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

The seven deadly sins of sour water stripping

Sour water systems should be designed to minimise operating problems, maximise on-line factor and optimise the quality of the feed gas to the sulphur recovery unit. **D. Engel, P. le Grange, M. Sheilan** and **B. Spooner** of Sulphur Experts describe the process of sour water stripping and focus on the most common mistakes (the seven deadly sins) made in operating and designing these units.

Sour water stripping is the first step in the treatment of 'process' wastewater in many industrial operations, particularly in refineries. Water streams from a number of process units throughout a refinery complex are typically sent to the sour water stripper (SWS), which is designed to remove hydrogen sulphide (H_2S) and ammonia (NH_3) from the process water. There are several variations in the designs of sour water strippers, all playing upon the same theme of using heat to break the bonded ions in the NH_4SH salt contaminant in the wastewater. This liberates gaseous ammonia and hydrogen sulphide in a produced sour water acid gas (SWAG). In some designs the NH_3 and H_2S are separated in separate columns and sent to individual destinations, but in the majority of SWS applications the effluent acid gas from a sour water stripper overhead is processed in a sulphur plant.

As oil and gas processing facilities deal with increasing sulphur content in their feedstock in combination with enhanced environmental pressures to remove sulphur from finished hydrocarbon products, the volume of "sour" water containing H_2S, ammonia and other contaminants is increasing. Additionally, the concentration of contaminants is increasing and exerting higher demands for sour water processing capacity. Simultaneously, more stringent environmental legislation and tougher fines for non-compliance have led to increased focus on the availability and reliability of sour water treating units.

A correctly designed, properly operated and well maintained sour water stripper (SWS) unit is critical to these operations. If the SWS unit is ever out of service, the facility must often run at reduced throughput or even temporarily shut down. As a result, sour water must be stored in a holding tank until processing is re-established and must often use tank capacity not designated for water storage. Quite often sour water composition is unknown (especially the contaminants other than H_2S and NH_3), which can make correctly setting operating conditions quite difficult. In other situations, SWS units with fluctuating hydrocarbons in the feed create problems for the downstream sulphur recovery unit. In these cases the acid gas is sent to a furnace or flare system for incineration, resulting in significant SOx and NOx emissions. Finally, some units do not make product specification treated water and therefore a proper understanding of the fundamentals of the process can help the operator come to a rapid and effective optimisation of the unit.

Sour water stripping process chemistry

The purpose of a sour water stripper is to remove components that are toxic or cause undesired odour. The most important ones are H_2S and NH_3 but other components like CO_2, HCN, mercaptans, phenols, hydrocarbons and solids are removed to varying degrees.

The SWS process entails contacting the sour water flowing down a stripper column with steam flowing up the tower. When sour water enters the stripper it is heated, causing the ionically-bonded ammonium sulphate salt (NH_4SH), which is in the aqueous phase and has no vapour pressure, to decompose into H_2S and NH_3. H_2S and NH_3 have a vapour pressure and thus can be stripped into the vapour phase and separated into a different process stream. Any other volatile species in the sour water such as carbon dioxide, hydrogen cyanide, mercaptans and 'light' hydrocarbons are also stripped.

$$H_2S + NH_3 \rightleftharpoons NH_4SH \rightleftharpoons NH_4^+ + HS^-$$
$$\longleftarrow \text{ temperature } \longleftarrow$$

There are many process wastewater sources in a refinery, all of which have different contaminant compositions, flow rates and pressures. In addition, some sources may be continuous while others are intermittent. As a result, without proper upstream equilibration, design, and operation, the chemical composition and flow of water to the SWS may vary significantly. This can result in frequent and severe operational upsets both for the stripper and the downstream sulphur plant, the destination for the gases stripped from the water.

The most common process feed water sources are from: atmospheric crude columns, vacuum crude towers, steam crackers, fluid catalytic crackers (FCC), hydrodesulphurisation (HDS) units, hydrocracking (HCU) units, atmospheric desulphurisation (ARDS) units, coker units, amine reflux purges and TGTU quench towers. H_2S and NH_3 concentrations are the highest in water from the HDS, HCU, ARDS, and FCC units. Any water stream containing 10ppm or more of H_2S requires treatment before leaving site limits.

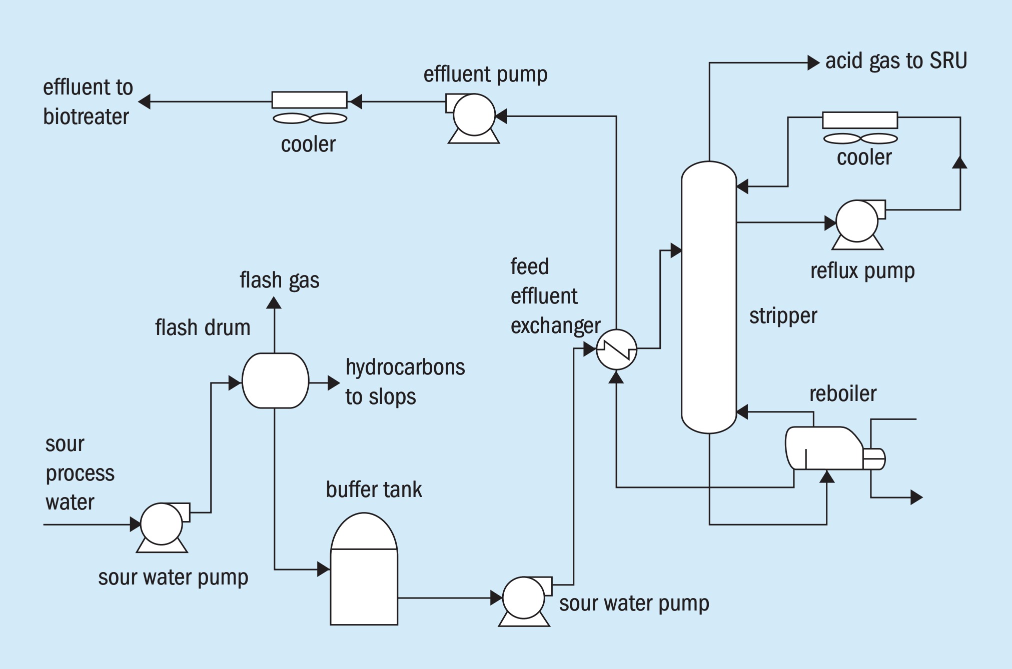
Meeting specification on NH_3 and H_2S in the treated water is extremely important as subsequent steps in wastewater treatment usually involve biological treatment which cannot operate under high hydrogen sulphide levels.

Sour water process description

The process can be viewed in Fig. 1 and summarised as follows:

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 34 35 36 37 38 39 40 41 42 43 44 45 46
47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

Fig 1: **Generalised sour water process scheme for a pumparound SWS**



Various sour water streams are collected from throughout the refinery and sent to the flash drum. The flash drum removes entrained and dissolved gases by allowing the water to de-pressurise and settle.

The degassed water flows to a holding, or settling tank (buffer tank). In this tank there may be further degassing as well as the separation of some liquid hydrocarbons that float to the surface of the water. If the residence time is long enough, the composition of the water stabilises and allows for a consistent flow and composition of water to the stripper.

If any filters are present in the process loop, they may be present upstream or downstream of the settling tank.

The water exiting the settling tank is heated in an exchanger (feed/effluent exchanger) by hot, stripped water exiting the stripper.

The heated sour water enters near the top of the stripper tower where it flows downwards and is stripped of H_2S, NH_3, 'light' hydrocarbons and other volatile species, by the steam rising from the bottom. Steam is produced in a reboiler or introduced into the column directly as live steam.

The overhead of the stripper may consist of a pump-around cooling section which cools the stream to a minimum of 85°C. Alternatively, a reflux system is used for the same purpose. These systems recover a portion of the water in the overhead stream, decreasing the amount being sent to the SRU.

The stripped water is cooled in the feed/effluent exchanger, then pumped to various areas for further use or processing (crude desalter, tail gas unit quench tower, biological treatment, etc.).

Sour water sources

In a refinery setting or any plant in general, sour water can be generated in many locations. Water for process applications is used in many ways such as: quench water, steam, wash water and is also generated by the various distillation fractions where water is co-distilled with certain hydrocarbons. Figures 2 and 3 show the diversity of a number of sour waters originating from various units and Fig. 3 provides an indication of the quality of the treated water in relation to the feed quality.

Primary sour water sources include:

- Amine system reflux water purges. TGTU quench water.
- Atmospheric and vacuum crude towers: water is produced by condensation of steam in overhead streams. Vacuum towers may also contribute sour water from ejectors and barometric condensers.
- Thermal and catalytic cracking units: sour water originates from wash water injection, stripping and aeration.
- Hydrotreater and hydrocracker wash water from high and low pressure separators.
- Cokers, delayed and fluid type plants. Water is produced from decoking and quench water.
- Flare seals and knock out drums.
- Hot condensates from throughout the refinery which may have had contact with hydrocarbons (often the concentration of contaminants in these streams is low).
- Any refinery water draw boot: each contains a different sour water composition and flow, depending on crude type and the severity of the process. Manual level controls can also affect the hydrocarbon content of the water especially if they are accidently left open for too long.

According to a previous study by the American Petroleum Institute[2] covering process water consumption estimates, the summary level quantities of water used in refineries were as presented in Table 1.

Table 1: **Refinery conversion unit process water estimates[2]**

Refinery conversion unit	Estimated process water use US Gal/1000 bbl	Water use, US gpm per 100,000 BPSD crude
Distillate hydrotreater	1,500	31
Cat Feed hydrotreater	2,400	66
Vacuum unit	2,000	69
Crude unit	1,400	97
Coker	8,000	112
FCCU	4,500	125
Total		**500**

Seven deadly sins of sour water stripping

Over the years, multiple problems and deficiencies have been uncovered. These have been compiled in the form of a list of the seven most deadly sins of sour water stripping:

1. incorrectly designing the sour water stripper column;
2. incorrectly controlling the overhead and acid gas temperatures;
3. poorly managing the sour water;
4. poorly operating or designing the flash vessel and feed stabilisation tank;

5. inadequate removal of solids and liquid hydrocarbons;
6. lack of a detailed sour water analysis;
7. neglecting the sour water stripper metallurgy.

No. 1: Incorrectly designing the sour water stripper column

In designing a sour water stripper tower, there are several options available depending on the treated water specification of the particular plant. The downstream destination of the water determines the allowable amount of H_2S and NH_3. The typical design basis can be described as follows:

Feed
$[H_2S] \approx 500 – 24{,}000$ ppm
$[NH_3] \approx 250 – 12{,}000$ ppm

Product
$[H_2S] \approx 1 - 25$ ppm
$[NH_3] \approx 10 - 50$ ppm

Generally, the options for design relate to a trade-off between the number of trays or height of packing and the quantity of steam required for stripping the contaminants from the process feed water. That is, the more contact stages available in the tower, the less steam required for stripping. Also, since ammonia is less volatile than hydrogen sulphide (therefore, harder to strip) it is usually the component that sets the quantity of contact stages. In general, the more the ammonia content in the feed stream, the more contact stages required or the higher the steam rates for stripping.

There are also options related to the regeneration medium; stripping steam generated within a reboiler or live stream injection directly into the base of the tower. If a reboiler is used, there is an option to go with a kettle-type or thermosyphon, or one of these reboiler types in combination with the option of incremental live steam injection. If live steam is used, it must be understood that this will increase the treated water content by 10-15%, which could increase treatment costs, which are normally linked to volume.

Lieberman[3] has stated that it may be possible that the extra water generated in the live steam mode could reduce the make-up water requirements for processes such as de-salters and hydrotreater effluent washes. Since external water is required for make-up for the above mentioned units, the increased water produced by the live steam injection could off-set the fresh water make-up. The typical energy usage in the stripper is in the range of 15% steam on a mass basis to the pounds/kilograms of sour water; 1.3 – 1.5 lb steam per US gallon of sour water[1].

Fig 2: **Various process water streams feeding an SWS**

Fig 3: **Water quality in feed and effluent of an FCC (L), Coker (C) & HT (R) SWS**

There are several options related to the type of reflux section in the top of the tower. Options to choose from include:

- no reflux at all (generating enough steam in the reboiler (or via live steam) to produce a stripper overhead temperature of around 88°C)[3];
- standard refluxed sour water stripper, with condenser/cooler, accumulator and pump;
- pumparound reflux, with externally cooled and pumped water system in a discreet top section of the tower.

The majority of the sour water strippers around the world use either the pumparound or refluxed condenser methods, with close to an even split between the two methods. Non-refluxed strippers are not favoured in modern industry as they can have excessive water content in the SWAG if overhead temperatures are not diligently monitored. Further they may experience significant capacity limitations in the event that feed temperature drops, because of exchanger fouling for instance. A feed temperature drop in a non-refluxed stripper needs to be compensated for via increased boil-up which in turn leads to higher vapour and liquid traffic below the feed tray with potentially lower flood points in the column as a result.

The pumparound process has potential for less corrosion than a refluxed system because it is liquid filled and not as prone to solid salt deposition; the relative concentration of ammonium salts is less in the pumparound as long as the temperature does not drop too low. The general target for the reflux temperature is >185°F (>85°C) to eliminate the potential for ammonium salt precipitation in the water loop and the associated piping to the SRU.

There are also options on the tower internals themselves. Historically, the towers have been trayed, with an option for sieve or valve trays. Anecdotal evidence seems to indicate that either sieve trays or grid trays will handle the inherently fouling service best. Valves can become stuck to the tray deck, which will promote plugging and flooding. In recent years, some operators/designers have tried using random

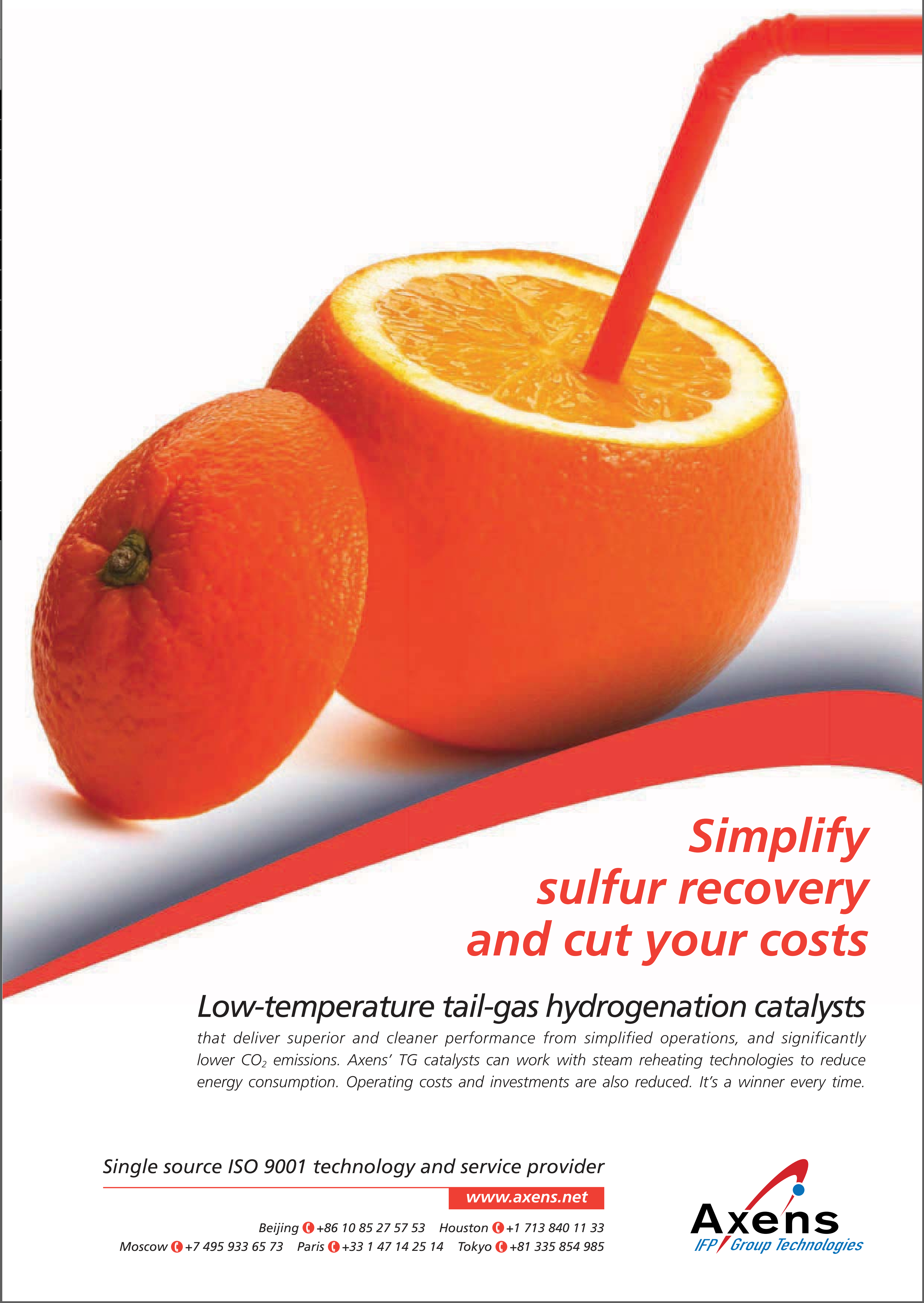

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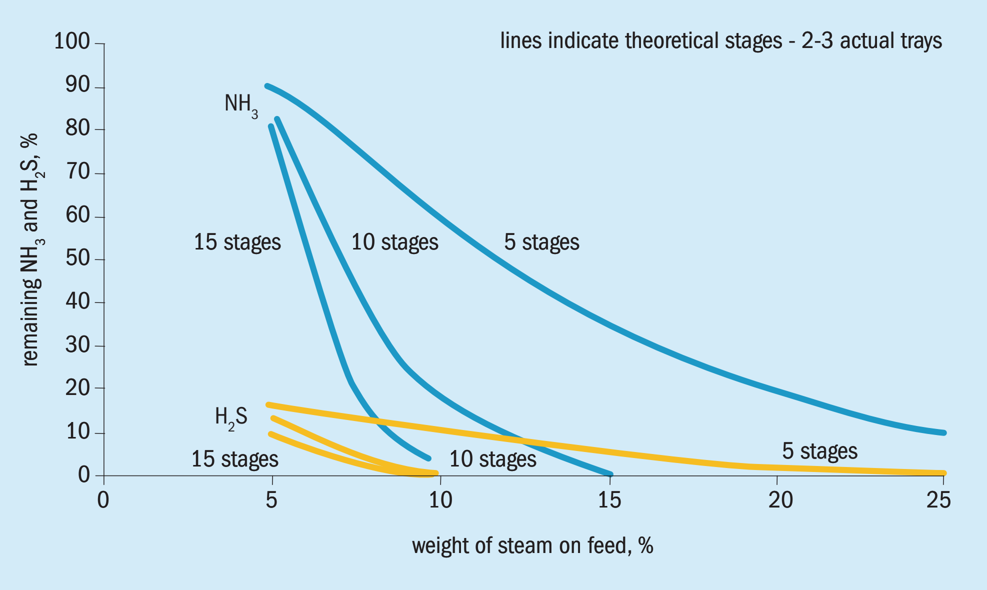
Fig 4: **Tray and steam effect on H_2S and NH_3 removal**



packing. Random packing has much lower pressure drop, thus providing a higher tower capacity than a trayed vessel. Unfortunately, a packed bed requires exceptionally good liquid and vapour distribution within the tower and given the potential for fouling in sour water service, this option may prove to be troublesome. To be safe, Sulphur Experts recommends that packing (either random or structured) not be used in sour water service.

The core question becomes how many stripping trays or how high a packing height and how many reflux trays are required to properly strip the sour water. Figure 4 provides a graphical overview of the effect of theoretical stripping stages and weight of steam to feed on the treated water H_2S and NH_3 content. Tray efficiencies have been loosely regarded as being somewhere between 30-45%, so around three trays per theoretical stage. There are rate-based software simulators available that can accurately predict sour water performance on an actual tray-by-tray basis, allowing the design engineer to feel confident in accurately determining the tray-to-steam ratio that will most economically and efficiently treat the sour water.

The pH of the sour water plays a very significant role in the ability of the steam to strip the H_2S and NH_3. Because H_2S is a weak acid in solution, it remains dissociated under alkaline conditions and is difficult to strip from the water. If the pH is low (<5.5) it returns to its gaseous form much more readily and it is possible to remove almost all of the H_2S from the water. NH_3 is basic in nature, so it would require a high pH of the water to return to its gaseous form. Full dissolution can be achieved if the pH is in excess of 10 and sufficient steam is introduced to the regenerator or stripper bottoms.

Theoretically, there would be two strippers, one operating with a low pH for maximum H_2S removal and another operating at high pH for maximum NH_3 removal. Because most refineries do not have that luxury, the single stripper may need to be modified to improve the potential for success of removing these disparate contaminants from the water. Because H_2S is easier to strip than NH_3, operations should err on the side of improving the NH_3 removal so the target pH of the sour water is slightly basic (around 7.5 to 8.5) to try to improve the removal of the less volatile ammonia fraction.

A possible modification would be the addition of a strong base such as caustic. Historically, the addition point has been recommended at some point lower in the tower to reduce the likelihood of "binding" the H_2S before it has had a chance to be stripped from the sour water. The ideal location of addition may even be in the sour water feed itself, but that is best decided upon with rigorous rate-based modelling for any particular application. Associated with the addition of the caustic is the need for an accurate means of measuring the water pH in both the sour water feed and effluent. Any excess addition of caustic can be detrimental and result in caustic deposition, binding of H_2S and poor stripped water performance in the de-salter as emulsion formation is elevated. It is important that the lag time between dosing and pH measurement is minimised to avoid overdosing the unit.

Figure 5 provides a review of a particularly poorly designed sour water system.

There was no buffer tank after the flash drum, leaving no opportunity to moderate flow rate and composition. Lack of a buffer tank also means there is no place to store sour water in the event of a stripper outage.

With no feed/effluent exchanger, the feed temperature to the SWS is too cold, requiring an inordinately high amount of stripping steam.

Because there is no reboiler, this extra steam is all live steam injection, which increases the effluent water volume dramatically.

Finally, and most troubling, is the routing of the flash gas vapours (primarily hydrocarbons) to the SRU feed stream. The flash tank is present to remove hydrocarbons from the feed to the SWS because they will naturally end up in the SWS overhead stream feeding the SRU. But this design sends these removed hydrocarbons right back into the SRU feed. There is practically no reason to have the flash tank in this design. These hydrocarbons wreak havoc in the SRU as far as air demand in the reaction furnace, side reactions to unwanted species and major coking on the sulphur catalyst.

Major capital outlay was required to bring the unit up to 'best practices' guidelines (installed a buffer tank, a reboiler and a feed/effluent exchanger) and the flash gas was re-routed to a low pressure refinery absorber. Currently, the unit runs virtually trouble-free, steam consumption has been more than halved, SRU operations are smooth and sulphur quality is excellent.

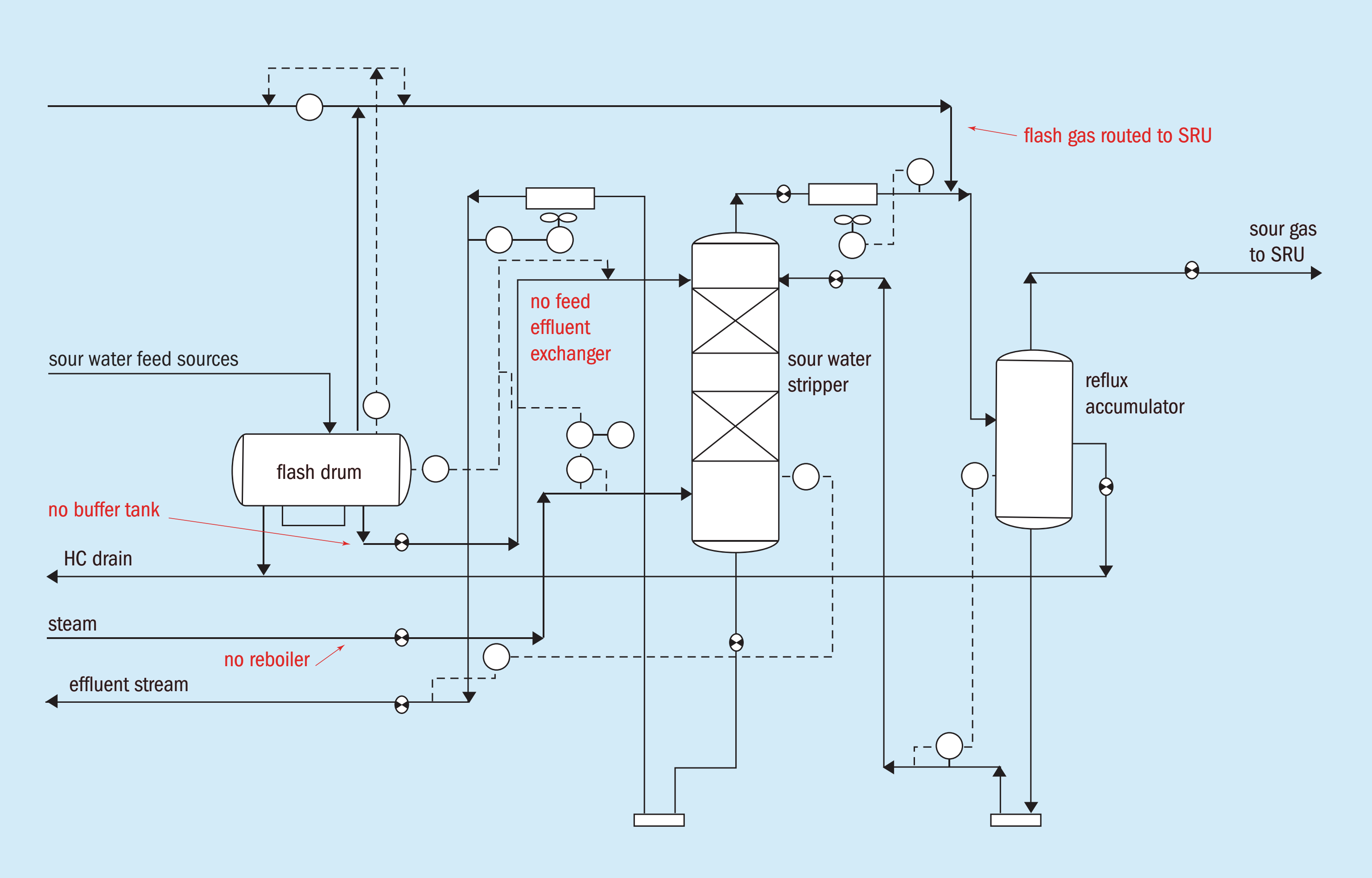
No. 2: Incorrectly controlling the overhead and acid gas temperatures

Heat is the primary component in effective SWS operation. Heat is required for:

- raising the water temperature from the feed temperature to the boiling point (reboiler temperature); sensible heat load;
- providing the temperature for the reaction of the ionic salts back into pure components;
- providing the heat to transfer the pure components from the liquid to the vapour phase;
- providing a diluent environment by lowering the partial pressure of the stripped gases by providing excess steam vapour (produces the reflux flow).

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 34 35 36 37 38 39 40 41 42 43 44 45 46
47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

Fig 5: **Improperly designed sour water system**



Without sufficient heat, the stripping of H_2S and NH_3 will not occur. In a well operated and designed sour water stripper there is always sufficient heat available in the column for the final stripping of H_2S and NH_3 to occur.

There are three temperature effects in the overhead system that the SWS operator should be aware of:

- ammonium salt sublimation;
- corrosion due to high salt content in the reflux water;
- high temperature polymerisation.

Fig 6: **Ammonium salt sublimation**

Ammonium salt sublimation

Ammonium carbonate and bicarbonate sublime in the temperature range of 55 to 75°C (130 to 167°F). When SWS overhead gas is cooled too much, salts precipitate and foul instruments, control valves and lines. This has been experienced on numerous sites throughout the industry, some typical examples of this are shown in Fig. 6. Sulphur Experts recommends a minimum temperature of 85°C to prevent fouling of the system due to salt deposition. Checking the instruments and overhead lines to the sulphur recovery unit (SRU) for cold spots should be done on a regular basis. It is standard industry practice for these lines to be insulated and steam traced but steam jacketing is preferred.

Corrosion due to high salt content in the reflux water

Most metallurgy is not rated for the high salt contents (>35 wt-%) that can be found in reflux water if the temperature is not maintained (Fig. 6). As H_2S and NH_3 are more volatile than H_2O, operating the overhead system at a higher temperature will decrease the salt content in the reflux water. This, unfortunately, increases the water content of the SWAG gas to the SRU which has an adverse effect on its operation. Salt content in the reflux system is SWS specific. With good test data a safe operational temperature can be set.

The pump shown in Fig. 7 is from the pumparound reflux of the first stage of a two stage stripper unit. The highly corroded stainless steel impeller was found only six months after the unit was commissioned along with multiple other leaks in the pumparound system. The pumparound was operating at 59°C. Simulation revealed that the ammonium salt content in the reflux was at 35 wt-%; subsequent

Fig 7: **Reflux pump corrosion due to elevated salt content**

lab analysis revealed 38 wt-% of NH_4HS in the pumparound.

After extensive simulation study the best solution (which was later implemented) was to take the pumparound out of service and utilise a water wash in the top of the column to keep the stripper top temperature at a reasonable level (~85°C).

High temperature polymerisation

Steam tracing/jacketing for winterisation or fouling prevention should not exceed 150°C as this increases polymerisation reactions which can result in fouling.

No. 3: Poorly managing the sour water

Poor water management is, unfortunately, quite prevalent in the industry. Broadly, this falls into four categories:

- cross contamination;
- dilution;
- segregation of phenolic water;
- bulk hydrocarbon ingress.

Cross contamination

Contamination of the SWS system with improper water streams needs to be avoided. There is no reason for cooling, fire or ballast water in a SWS system, as the Ca/Mg hardness in these streams will foul the reboiler and trays below the feed nozzle. Spent caustic or waste from an alkylation unit should also not be sent to a SWS, as these streams contain strong bases or acids which will bind H_2S or NH_3, resulting in off specification stripped water. It is critical that only "process" water streams be routed to the SWS. Note that spent caustic could potentially be used for caustic dosing, however the caustic strength must be quantified and dosing rates or PH controller tuning should be adjusted accordingly.

Dilution

Dilution has the following negative impacts:

- causes poor energy efficiency in the SWS due to unnecessary processing of inappropriate waters
- increases the cost of downstream treatment and disposal due to greater treated water volumes
- uses SWS capacity unnecessarily, which may affect plant flexibility

Common sources of SWS water dilution include:

- direct steam injection: this traditional (low capital) design approach should not be used because it increases the effluent water quantity by 10 to 20%;
- routing uncontaminated or low contamination streams to SWS (streams feeding the stripper should be tested for H_2S, NH_3 and phenols);
- dumping of other (non-sour) water streams into the SWS system.

Segregation of phenolic water

Phenolic water primarily comes from refinery cracking units such as cokers or FCCs. It is important to understand that only a small fraction of phenols in water will be removed in a SWS. One African refinery was off-specification on phenols for several years as a result of not been cognisant of this. This is largely a result of the low phenol volatility. A good simulation study of this was published by Hatcher et al[14].

Most of the phenols can be removed in the crude desalter unit, which is downstream of the SWS. It is important to route all the phenolic effluent water to the desalter, as the phenols tend to partition into the oil phase, thus reducing their phenol content in the effluent stream. If possible it is also recommended to use a separate SWS for phenolic waters so that the non-phenolic water can stay segregated from the phenolic waters.

Bulk Hydrocarbon Ingress

The best solution to minimising hydrocarbons in sour water feeds is to ensure that the hydrocarbons are not in the water in the first place. This is carried out by a comprehensive and thorough evaluation of sour water generation points. It is important that all SWS feed streams are analysed, regardless of flow rate. An example of this is shown in Fig. 8, which shows a sample from a water stream which contributed less than 10% of the feed flow, but was responsible for over 90% of the hydrocarbon contamination.

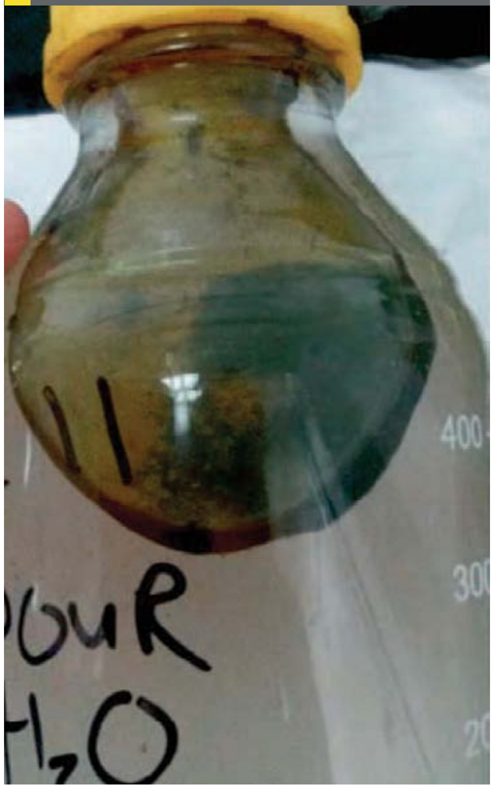


Fig 8: **A low-flow sour water system with high hydrocarbon levels**

At the commissioning of a Middle Eastern refinery, there was a significant amount of bulk heavy diesel fraction hydrocarbon in the sour water. This led to fouling of the feed/effluent exchanger within three months of start-up and reduced performance on the stripper.

No. 4: Poorly operating or designing the flash vessel and feed stabilisation tank

Good flash vessel operation and design is vital. The flash vessel serves to remove the light hydrocarbons and the bulk of the heavier liquid hydrocarbons. Without a proper flash vessel the SWAG cannot be safely sent to an SRU.

Preventing hydrocarbons from entering the SWS will prevent hydrocarbon from entering the sulphur plant (SRU). There are several reasons why it is advantageous to minimise hydrocarbon in the SRU feed, the most important being:

- difficulties maintaining stable SWS (and consequently SRU) operation;
- decreased capacity;
- lower efficiencies;
- potential catalyst deactivation and sulphur quality issues due to soot formation in the downstream Claus reactors.

Hydrocarbons in the sour water feeding the stripper will also significantly increase the fouling of stripper internals. The "black shoe polish", which is found on sour water

stripper internals, typically contains heavy hydrocarbons at varying degrees.

The flash vessel is a three phase separator; its purpose is to separate water, oil and gas. This is achieved through pressure drop and residence time; the greater the pressure drop or greater the residence time, the better the separation of the three phases. Flash vessel operation is therefore at its peak when the pressure in the vessel is as low as possible and residence time maximised. The minimum recommended residence time is 20 minutes at the normal operating level of 50-60%. The lower the pressure, the more likely hydrocarbons will flash off, as pressure has a direct effect on the vaporisation point of hydrocarbons.

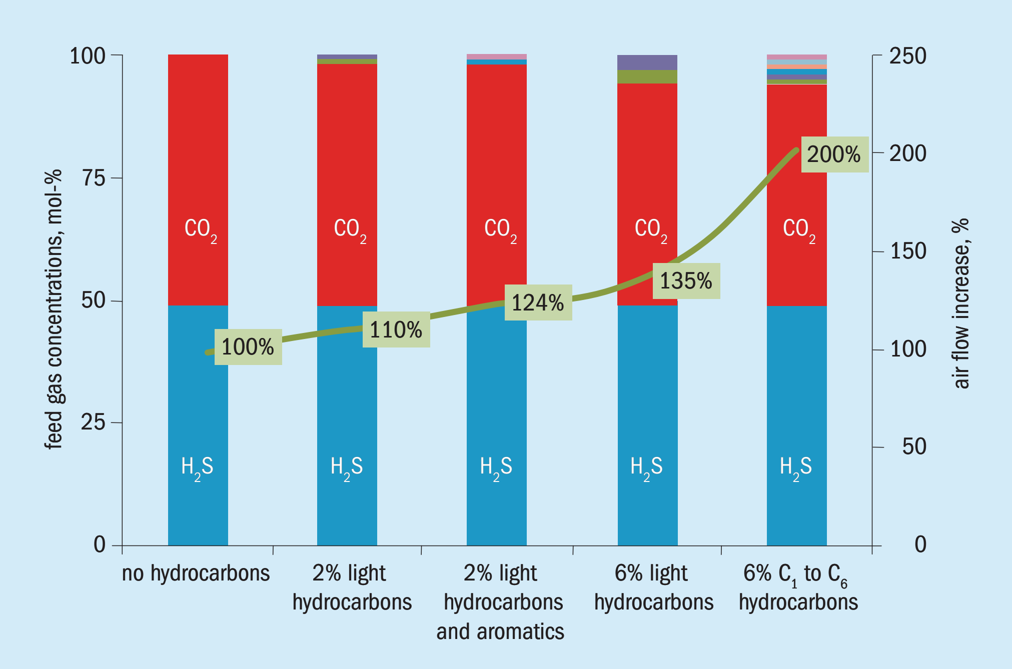
The pressure is set by the destination pressure of the flash gases. These gases are normally sent to flare, incineration, or a low pressure fuel gas amine absorber. Under no circumstances should flash gas be routed to the SRU, as the flash gas will have a continuous fluctuation in both flow and composition. Figure 9 presents a graphic visual of the potential effect of hydrocarbon load on reaction furnace operation. With excess heavy-end hydrocarbons the required air flow for combustion can double.

If the flash vessel uses weirs, they will typically be set at a height of 50-60% of the vessel. The water level should be maintained at 7-8 cm below the weir height, allowing liquid hydrocarbons to then flow over into the oil side of the weir. When both the size of the tank and liquid level are set, then the only option to increase residence time is to reduce the water flow to the vessel by critically evaluating all streams feeding this vessel.

One of the prevalent causes of SWS unit upsets is from large fluctuations in the composition and quantity of the sour process water. These fluctuations are inherent to refinery operation and can be prevented by a properly sized SWS feed stabilisation/buffer tank, with the water feed on one side of the vessel and the exit on the other.

The stabilisation/buffer tank also serves to partially remove suspended solids and liquid hydrocarbons. It is essential that the buffer tank has skimming facilities installed to remove these hydrocarbons. The buffer tank can be designed with a gasoil layer floating on the top, as a blanket, to avoid smell problems. A better option is to use an internal floating roof with hydrocarbon skimming facilities. This is a more costly option but it will significantly reduce the serious odour problems of SWS feed water.

Fig 9: Effect of hydrocarbons on reaction furnace air flow



Similar to the flash vessel, this stabilisation tank is normally operated 50% full. The correct level is however, always a compromise between the various functions: feed stabilisation, free storage capacity and separation. This tank must be partly empty, to allow for upsets of the stripper which can last hours or significantly longer. On the other hand, a longer residence time will improve hydrocarbon/solids separation and most importantly will stabilise the SWS feed composition and flow.

Along with the height of the liquid level, the location of the feed and discharge water lines play an important part in the stabilisation role. The inlet and outlet nozzles should be located at opposite ends of the vessel, so as to minimise potential bypassing of contaminants. The outlet nozzle is often 600 mm from the bottom of the tank so that precipitated solids or heavy oils are not pumped out of the tank along with the sour water. The buffer tank should be designed with a bypass to accommodate cleaning.

No. 5: Inadequate removal of solids and liquid hydrocarbons

Poor filtration of solids and inferior liquid hydrocarbon removal can result in fouling and corrosion problems of the SWS unit, which then leads to poor reliability and decreased run lengths between shutdowns. Additionally, hydrocarbons that are not separated at the source of the feed sour water can be present in the sour water acid gas affecting the downstream sulphur recovery process.

Hydrocarbons

Hydrocarbons in water streams can be present essentially in three forms:

- free hydrocarbons;
- soluble hydrocarbons;
- emulsified hydrocarbons.

Free hydrocarbons

These will not interact with the bulk water and will tend to separate within a few minutes in the flash vessel. Free hydrocarbons are normally observed by the formation of a top hydrocarbon layer above the water phase (or below depending on the density difference). The levels of free hydrocarbons can vary from 100 ppmw to percentage levels. Their separation efficiency is calculated via Stokes Law, which has a large effect from the droplet diameter (Figure 10). Viscosity and density difference between the phases also plays a lesser role, with lower viscosities improving separation velocity. Generally, this indicates some benefit from running at slightly higher temperatures as the fluid viscosities decrease at higher temperatures.

Soluble hydrocarbons

All hydrocarbons will have a certain solubility in water phases. The extent of hydrocarbon solubility in water will depend on the pH of the water, water pressure, temperature and the type of hydrocarbon. It is impossible to observe dissolved hydrocarbon in a water phase as it is indistinguishable from pure water. In general, the solubility of hydrocarbon in water can

Fig 10: **Relationship between hydrocarbon droplet diameter & separation time**

Stokes Law:

$$V = \frac{gd^2(\rho_a - \rho_b)}{18\mu}$$

where:
V = separation velocity
g = gravitational force
μ = viscosity continuous phase (water)
ρ_a = density of solvent phase
ρ_b = density of liquid contaminant
d = diameter contaminant droplet

Droplet diameter, microns	Time
160	30 min
106	1 h
75	2 h
43	6 h
22	24 h
16	48 h
8	168 h

water + diesel-like hydrocarbon at the discharge of a centrifugal pump

range from a few ppm to a few hundred ppm. The solubility of hydrocarbon in sour water has a direct relationship with the pH; the higher the pH, the higher the solubility.

Emulsified hydrocarbon

Under normal conditions, hydrocarbons will be either free or dissolved in a water phase. However, when conditions are conducive (including the presence of surfactants and energy), the hydrocarbon contaminants can form very small droplets in the water phase (Fig. 11). These droplets are stabilised by molecular surfactants (similar to soaps or detergents) and also by small size suspended solids. Emulsion droplet sizes can range from a few microns to about 500 microns. Micro-emulsions are the most stable emulsion type and can take weeks to naturally separate. Micro-emulsions are typically found when droplet sizes are less than 10 microns.

Suspended solids

Suspended solids in the sour water feed are fairly common, especially in plants associated with coker units. To some extent these solids will settle in the upstream feed stabilisation tanks (also undesirable), however, a considerable portion can be present in the effluent. The effects of suspended solids can be somewhat similar to hydrocarbons, as they will stabilise foaming and deposit on metal surfaces leading to reduced flows and under-deposit corrosion.

Most refineries use stripped sour water as desalter wash water, particularly in FCC refineries to remove the phenols from the stripped water. The presence of solids will enhance emulsification, impacting the effectiveness of water and crude separation in the desalter unit. This can lead to increased salts in treated crude, generating higher corrosion rates in the crude unit overhead. Many crude unit corrosion problems, desalter upsets and increased additive usage can be tied to improperly stripped water.

Many solids are present at diameters less than the visual acuity of the human eye (<40 microns) and they are not detected until they precipitate out of the solution en masse (Fig. 12). These are the types of solids that can really stabilise a foaming condition, and there is a very good chance that these low micron solids and the larger visible solids will settle in any stabilisation tank, so it is generally recommended to include tank cleaning during any turnaround situation. Leaving excessive solids in the stabilisation tank can lead to the eventual transfer of these solids out of the tank with the process water if they are allowed to rise to the level of the outlet nozzle on the tank.

Possible remedies for the conditioning of sour water

Filtration is the basic technology for removing suspended matter from the sour water. For the removal of emulsified hydrocarbons, the technology of choice is a coa-

Fig 11: **Water and oil micro-emulsion**

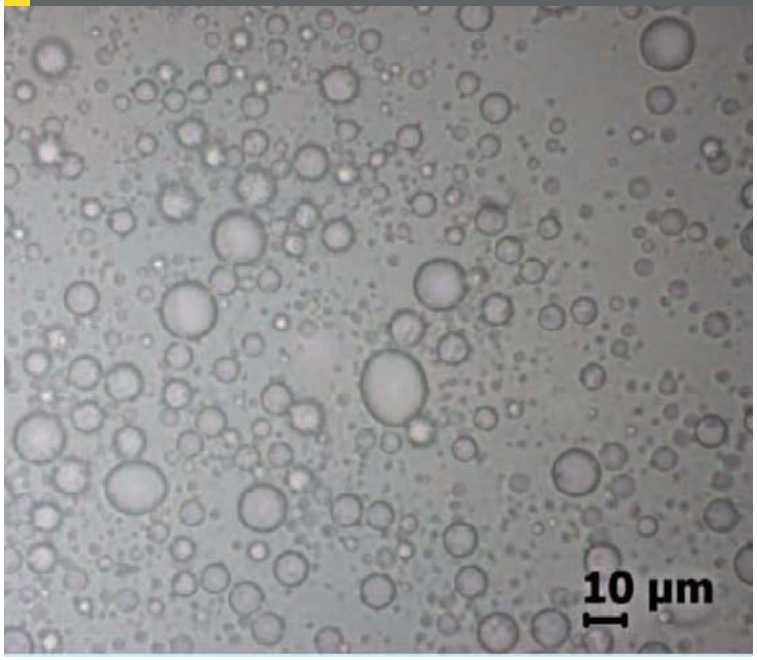


Fig 12: **Solids precipitating out of sour water after 12 hours**

lescer. Coalescence is the recombination of two or more small liquid droplets to produce a single entity larger in size.

A large stabilisation/buffer tank could be used to separate hydrocarbons and suspended solids from the various sour water streams. However, this is quite costly. A buffer tank rarely has sufficient residence time to accommodate effective separation of fine particulate and micro-emulsions (10 microns and smaller). It is therefore recommended to use filters and coalescers.

Due to the particle size and the high fouling properties of emulsified hydrocarbons in sour water streams, only disposable microfiber-based coalescers are able to provide proper emulsion separation. Other systems such as inclined plates and fibre mesh are not effective.

Suspended solids removal upstream of the hydrocarbon coalescer is mandatory. Particulate removal will protect the coalescer elements and also destabilise the emulsion, significantly improving overall system efficiency. The particle filter and liquid coalescer combination system should always be installed downstream of the sour water charge pump and upstream of the heat exchanger. This configuration is illustrated in Figure 13 and the effectiveness of such an arrangement is shown in Fig. 15. Figure 14 presents the deposition in a packed SWS caused by ineffective upstream solids removal.

Fig 13: **Optimal filtering and coalescing of sour water**

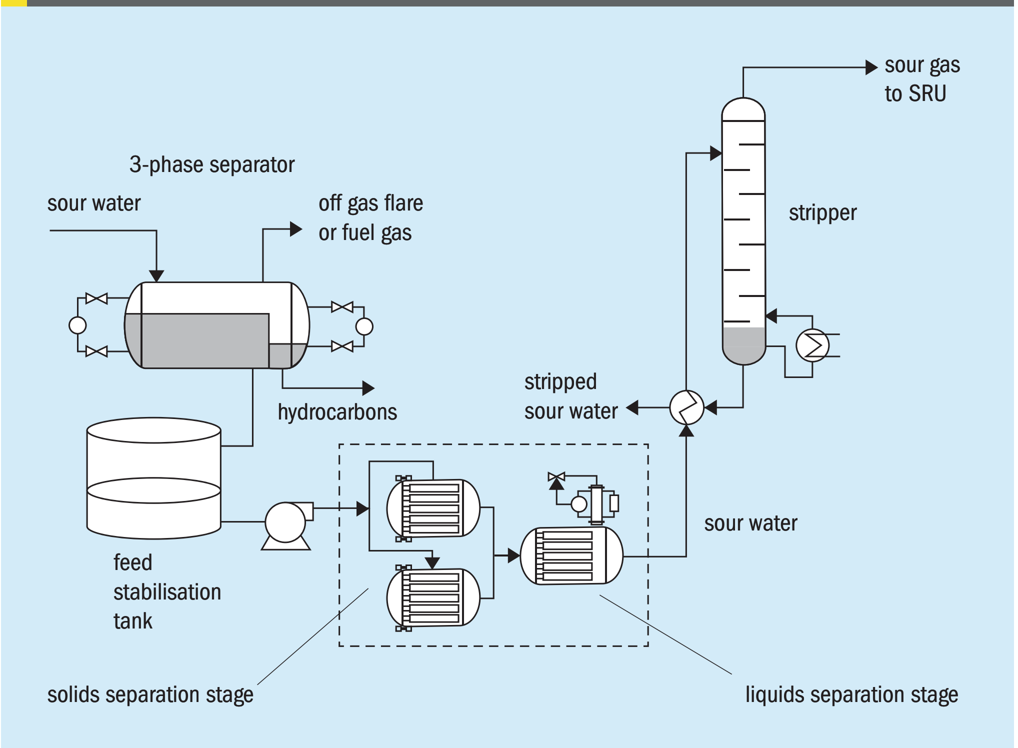


Fig 14: **Fouling in an SWS tower**

No. 6: Lack of a detailed sour water analysis

The design specifications of a SWS often only list the H_2S and NH_3 content of the combined feed. The possible presence of other components is very rarely mentioned. These other contaminants can create significant problems because they could:

- bind H_2S to the water and reduce its tendency to be stripped;
- bind NH_3 to the water and reduce its tendency to be stripped;
- plug up the trays or packing or scale on hot surfaces;
- lead to foaming conditions in the stripper;
- affect the performance of the downstream bio-treaters.

Common "other" sour contaminants are listed below:

- Sulphuric acid, hydrofluoric acid, formic acid and other acids will bind NH_3 in such a strong way that it will be almost impossible to strip. The NH_3 can be liberated and subsequently stripped by addition of a strong base, normally caustic, to neutralise the strong acids.
- Calcium and magnesium can be present if hard-water has been used as process water. It is also possible that fire water or cooling water has been discharged to the SWS unit. As a result, calcium and magnesium carbonates will deposit as scale in the reboiler.
- Elemental sulphur or polysulphides are normally caused by air ingress to the process water system. H_2S will be oxidised to sulphur and polysulphides. This sulphur will deposit as a scale on the bottom trays of the stripper and in the feed/effluent exchanger.
- Phenols are present in the process water from FCC units, cokers and thermal cracking units. Preferably this process water should be segregated in phenolic and non-phenolic water. Phenols are not properly removed in a SWS and therefore the phenolic water effluent should be sent to the crude desalter where the phenols are extracted to the crude.
- Nitrogen components such amines, filming corrosion inhibitors or HCN are very poorly removed in a SWS. Often the stripped effluent of the SWS is only analysed for NH_3 and therefore these other nitrogen components are missed. One of the important environmental goals of the SWS is to remove as much of the nitrogen components as possible.

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 34 35 36 37 38 39 40 41 42 43 44 45 46
47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

Fig 15: **Photograph of coalescer effectiveness in sour water service**

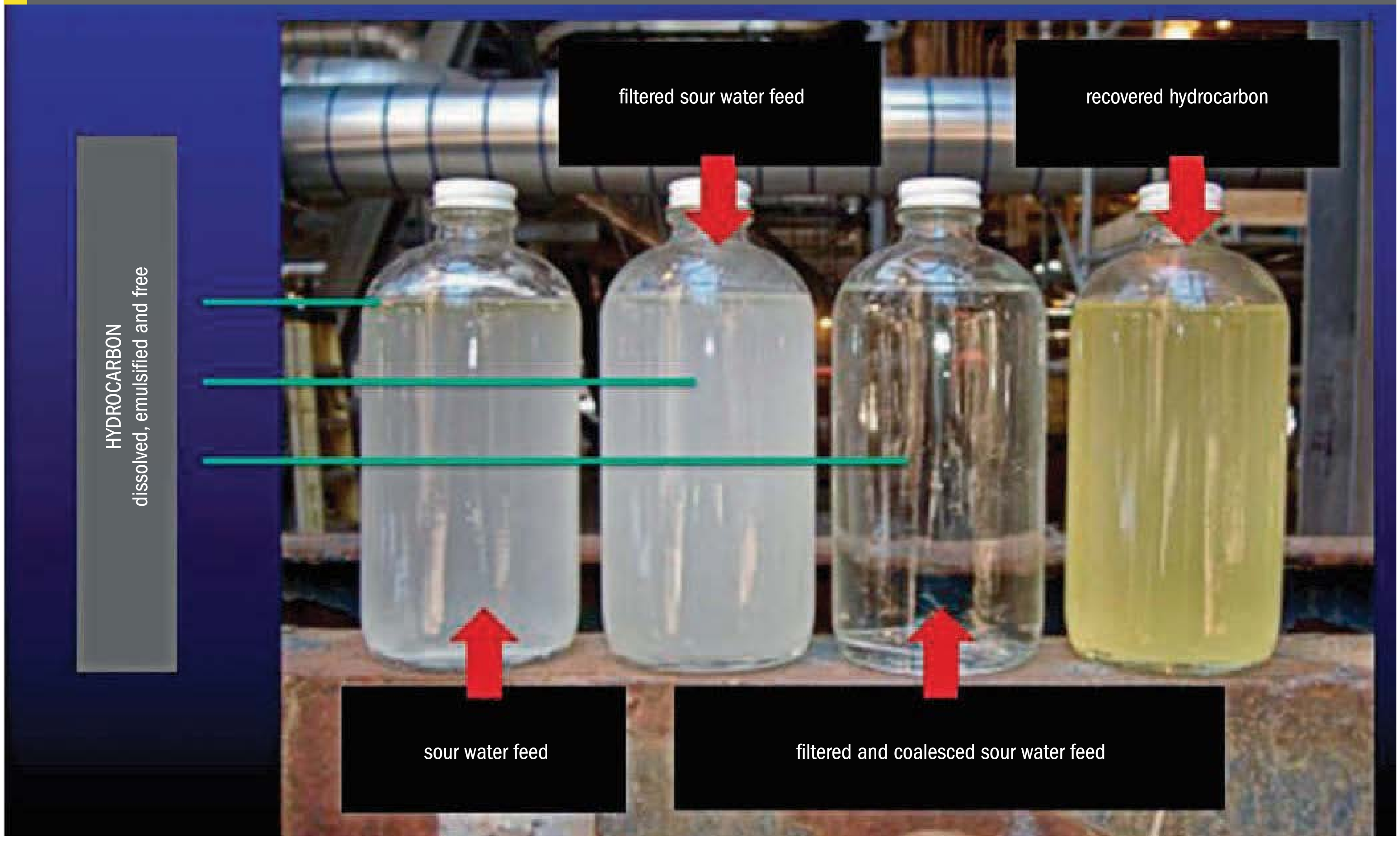
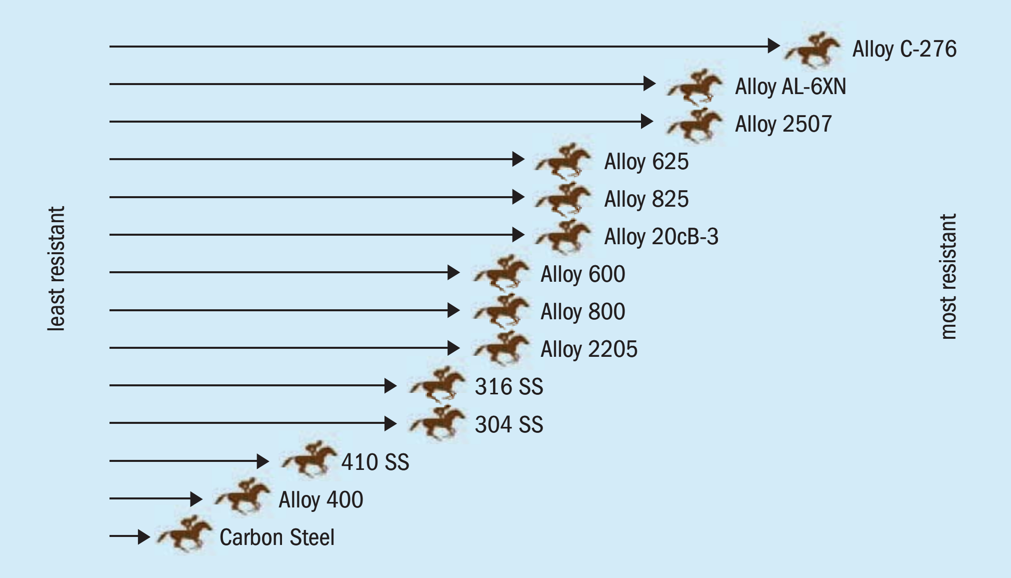


Fig 16: **Relative corrosion resistance for test materials in ammonium bisulphide environments[9]**



- Surfactants will certainly be present in the SWS feed, as they are removed by the various refinery wash-water systems. Analysis of these trace components is close-to impossible. As a consequence, the SWS needs to be designed with a safety margin to allow for foaming upsets.
- Mercaptans may also be present in the SWS feed, and will be removed in the stripper. As a result, they will contribute to the sulphur and hydrocarbon content of the acid off-gas going to the SRU. The overall design of the SWS and the SRU needs to allow for the additional air requirement for the SRU.
- Benzene, toluene, ethyl benzene and xylene (BTEX) will also be present in the SWS feed and removed in the stripper. BTEX compounds are well known to be detrimental to the life of catalyst in the SRU.

No. 7: Neglecting the sour water stripper metallurgy

The presence of ammonium bisulphide in sour water systems where both ammonia and hydrogen sulphide are present is the main driving force for corrosion. Ammonium bisulphide corrosion appears to be enhanced when flow rates are increased and also contributes to under-deposit attack. Flow-enhanced corrosion occurs at impingement points or after flow disturbances. The problem with corrosion in sour water strippers is that the corrosion has been historically hard to predict and most industry guidelines have been based on collected field experience with existing metallurgies. More recently, work has been done to build tools that will better predict sour water corrosion.

The following factors have been found to contribute to corrosion in sour water stripping systems:

- Ammonium bisulphide
  - Increasing concentrations result in increased corrosion
  - Some texts indicate a threshold level of 35 wt-%
  - Corrosion rates increase with water velocity
- Hydrogen sulphide partial pressure

Table 2: **Recommended metallurgy in sour water strippers**

Location	Minimum metallurgy	Notes
Sour water flash vessel	Carbon steel with 6 mm corrosion allowance	
Feed pump	300 series stainless steel internals	
Feed / effluent exchanger	Carbon steel shell with 6 mm C.A. AISI 316(L) SS or Incoloy 825 tubes	
Stripping column	Carbon steel shell with 6mm C.A. Trays are 316(L) SS	400 series metallurgy not sufficient
Effluent pump	Carbon steel casing and internals	
Overhead cooler	Titanium or Avesta 254 tube bundle; headers can be 316(L) SS	
Overhead accumulator	Carbon steel with 6 mm C.A.	It is important to Control reflux temperature well to maintain an acceptable level of NH_4SH
Reflux pump	316(L) SS casing and internals	
Piping	Most piping can be carbon steel with 3 mm C.A.	Piping in the overhead section should have a more robust 6 mm C.A. due to high NH_4SH levels
General		If cyanides are present at ≥30 ppm then HIC resistant steel should be selected (cyanides are usually present in the sour water from FCC units)

  - Corrosion rates increased with an increase in H_2S partial pressure; accentuated by higher velocities and higher ammonium bisulphide concentrations
  - Effect was far more extreme for the least corrosion-resistant materials (carbon steel, Monel 400 and Type 410 SS) – Fig. 16
- Temperature
  - As expected, an increase in temperature increased corrosion rates

Table 2 provides a quick review of recommended metallurgy.

Conclusions

In summary, sour water strippers have been designed and operated from simple systems with no buffer tank, minimal hydrocarbon removal, no filters, towers with few trays and live steam injection to deluxe systems with full pre-treatment stages (including flash tank, buffer tank, coalescer and filters) and segregated H_2S stripping and NH_3 stripping towers.

Ultimately, sour water systems should be designed to minimise operating problems, maximise on-line factor and optimise the SRU feed gas quality. In order to accomplish this the system should be designed with:

- a flash drum with three-phase separation capabilities and a minimum of 20 minutes residence time;
- a buffer tank with a minimum 24 hour residence time at 50% full;
- a filter followed by a coalescer for solids and hydrocarbon removal;
- a feed/effluent exchanger to heat up the feed stream and reduce the reboiler load;
- some consideration to segregating the ammonia fraction from the SRU feed stream (two-stage stripping);
- a reboiler (with the potential for live steam injection; a last resort as "dilution should not be the solution to pollution");
- reflux loop to control SRU feed temperature;
- insulated and steam traced piping to the SRU;
- a detailed feed water analysis;
- the correct metallurgy for the location in the system.

References

1. Weiland R.H. and Hatcher N.A.: "Sour water strippers exposed", LRGCC (2012)
2. The sources, chemistry, fate and effects of ammonia in aquatic environments, American Petroleum Institute (1981).
3. Lieberman N.: "Sour water strippers: Design and operation", PTQ (Q2 2013).
4. Van Hoorn E: "The basics of sour water stripping, Amine Processing Textbook", published by Sulphur Experts (2002-2012).
5. Klint B. "Hydrocarbon destruction in the Claus SRU reaction furnace", Sulphur Recovery textbook, published by Sulphur Experts (2005-2012).
6. Engel D.B.: "Filtration and separation technologies in sulphur recovery", Chemical Engineering Magazine (April 2013).
7. Kister, H.Z., Mathias P.M., Steinmeyer D E., Penney W R., Crocker B B., and Fair J.R.: "Equipment for distillation, gas absorption, phase dispersion and phase separation", "Perry's Chemical Engineers' Handbook", 8th Ed., Section 14, McGraw-Hill, New York (2008).
8. Horvath R.J., Lagad V.V., Srinivasav S., Kane R.D.: "Prediction and assessment of ammonium bisulfide corrosion under refinery sour water service conditions – Part 2;
9. Tems R.D.: "Ammonium bisulfide corrosion in hydrocracker and refinery sour water service", SAER-5942 Final Report on a Joint Industry Program (JIP)
10. "HCN and phenol in a refinery sour water stripper", The Contactor, Optimised Gas Treating Inc., Volume 7, Issue 11, (Nov 2013).
11. Spooner B., Sheilan M., van Hoorn E., Engel D., le Grange P., Derakhshan F.: "Reducing hydrocarbons in sour water stripper acid gas", LRGCC (Feb 2014).
12. Armstrong T., Scott B., Gardner A.: "Sour water stripping; Today's Refinery, (June 1996).
13. Kister H.: "Distillation operation"; pp327.; McGraw Hill, New York (1990).
14. Hatcher, N. A., Jones, C. E., and Weiland, R. H.: "Sour water stripping".

Improved heat recovery in sulphuric acid plants

Sulphuric acid plants are significant producers of energy. In this article Chemetics discusses the main methods of heat recovery for the secondary heat of the sulphuric acid plant, Outotec provides case studies on the application of state-of-the-art solutions for more efficient heat recovery from old sulphuric acid plants and NORAM highlights some of the options available to maximise energy recovery.

Today, sulphuric acid plants are commonly used as a local source of steam, and therefore energy, within a metallurgical or fertilizer complex. Control and management of the heat produced from a sulphuric acid plant is critical to the profitability and operability of the site.

Energy sources from a sulphuric acid plant can be summarised as follows:

1. Combustion of sulphur inside furnace to produce SO_2:

$$S + O_2 \rightarrow SO_2 \ (\Delta H = -295 \text{ kJ/mol})$$

2. Oxidation of SO_2 to SO_3 inside catalytic converter:

$$SO_2 + ½O_2 \rightarrow SO_2 \ (\Delta H = -99 \text{ kJ/mol})$$

3. Absorption of SO_3 into sulphuric acid:

$$SO_3 + H_2O \rightarrow H_2SO_4 \ (\Delta H = -138 \text{ kJ/mol})$$

4. Absorption of moisture in ambient air inside drying tower:

$$H_2O(g) \rightarrow H_2O(l) \ (\Delta H = -141 \text{ kJ/mol})$$

The amount of energy produced is dependent on the type of sulphuric acid plant. Sulphur burning acid plants will have all four forms of energy present. Metallurgical type plants will have types 2, 3 and 4. Regeneration type plants will have 2, 3, and 4 but will also have the heat from the combustion of fuel and spent acid contaminants in the acid regeneration furnace.

In Topsoe Wet gas Sulfuric Acid (WSA) plants, no matter whether the feed stream is H_2S, SO_2, elemental sulphur or spent sulphuric acid, heat recovery is an integral part of the process and not an add-on. In addition to the heat recovered in standard single or double absorption dry gas plants, in WSA plants the heat of hydration of SO_3 to H_2SO_4 vapour, the condensation heat of sulphuric acid and the enthalpy of cooling the process gas to 100°C are recovered. A typical double-condensation WSA plant based on burning of liquid sulphur will export more than 1.5 tonne of steam (54 bar g, 435°C) per tonne of sulphuric acid product. A single-condensation WSA plant handling H_2S gas will export more than 2.1 ton of steam per ton of sulphuric acid product.

The remainder of this article focuses on heat recovery from dry gas sulphuric acid plants.

Energy recovery options

Energy can be recovered from the acid circuit or from the gas circuit of the plant. Different options for each are as follows:

Acid circuit:

- generation of hot water via acid coolers;
- generation of 7 to 10 barg steam via a heat recovery system such as ALPHA®/HEROS/HRS®.

Gas circuit:

- high pressure (HP) boiler after the sulphur burning or acid regeneration furnace;
- medium pressure (MP) boiler, superheaters, and/or economisers after the converter beds;
- puller instead of pusher blower;
- use of strong SO_2 content gas into the acid plant;
- preheat combustion air from back end of plant (say bed 4 outlet).

For a sulphur burning sulphuric acid plant the total energy available is as follows and indicated as a value per metric tonne of plant acid production:

- Total reaction energy released
  ΔH = 1,507 kW/t
- Other energy input (sulphur / blower)
  ΔH = 45 kW/t
- Losses to stack/product
  ΔH = -35 kW/t
- Energy available for recovery
  ΔH = 1,517 kW/t

The assumption in the above calculation is that the plant has selected the option of using a puller type blower where the main process blower is located downstream of the dry tower. This configuration allows the heat and energy added to the stream by the blower to remain in the gas flow rather than be lost to the acid in the dry tower. This energy will be transferred to steam produced in the HP boiler at the end of the sulphur furnace and end up producing an approximate additional 1.4 kW of electrical power per tonne of acid produced.

Without recovering energy from the acid circuit only 73% of the energy can be recovered. This recovery will be done in the form of generation of HP saturated or superheated steam at an appropriate temperature and pressure for the site. This HP steam is sent to a turbine and used to generate electrical power. The choice of steam pressure and temperature is critical in the design of the plant. Higher pressure and temperature boilers and turbines are more expensive but considerable improvements in the quantity of produced power can be realised. For example, 60 barg and 500°C

Table 1: **Comparison of overall thermal efficiency of low grade heat recovery options**

	Hot water	MP steam for heating	MP steam for power
AP Energy recovered, kW/t	402 – 462	308	308
Useable energy, kWt	402 – 462	275	50
Thermal efficiency, %	max 100	max 90	max 17

Fig 1: **Double absorption sulphuric acid plant**

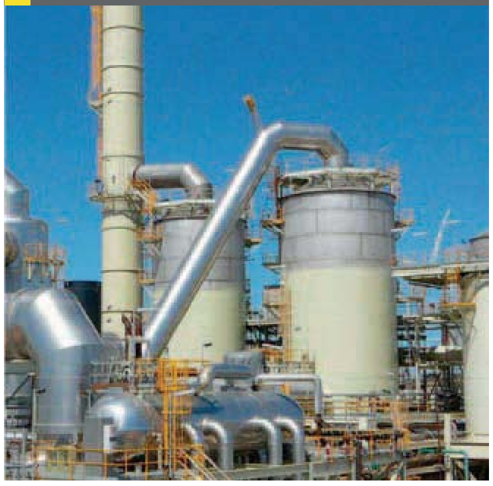


PHOTO: CHEMETICS

steam will produce about 15% more electrical power from the same sulphuric acid plant capacity when compared to 40 barg and 500°C steam.

Stronger SO_2 gas being sent to the converter can also allow greater heat recovery. Stronger gas will reduce the volumetric flow through the plant and will increase the heat recovery in each boiler, superheater and economiser. However, stronger gas will also result in more catalyst being required to achieve the same SO_2/SO_3 conversion. Strengthening the gas slightly has an economic benefit, but making it too rich will typically be a detriment.

For additional heat recovery from the acid circuit, options are available to allow production of 90-92°C hot water (99% heat recovery) and/or production of medium pressure (MP) steam (up to 93% heat recovery). Hot water heat recovery has higher efficiency because it captures the heat from both the final and inter absorption acid towers while the MP steam system only captures heat from the inter tower in a double absorption plant. In summary, for a sulphur burning acid plant:

BFW heating & hot water from acid circuit

HP steam, kW/t	1,047	(69%)
BFW heating, kW/t	60	(4%)
hot water (90 °C), kW/t	402	(26%)
cooling water, kW/t	18	(1%)

Heat recovered to hot water = 462 kW/t (30%)

BFW heating and MP steam from acid circuit

HP steam, kW/t	1,047	(69%)
BFW heating, kW/t	79	(5%)
MP steam, kW/t	308	(20%)
Cooling water, kW/t	93	(6%)

Heat recovered to MP steam and BFW = 387 kW/t (25%)

The energy that can be recovered in both cases can be viewed as low grade energy. To recover this energy costs money and the best return on investment choice is nearly always the one that recovers the energy in the cheapest form it can be effectively used.

Table 1 indicates the relative comparison of the thermal efficiency percentage for:

- heat recovery to 90°C hot water; versus
- heat recovery to MP steam directly; versus
- heat recovery to MP steam and then using that MP steam to produce power.

Recovery to hot water allows up to 68% more usable heat compared to MP steam. The capital cost of the required sulphuric acid coolers and systems to recover the heat as hot water is about 20% of the cost of an ALPHA® (Acid Low Pressure Heat Absorption) system to recover to MP steam. If a local use can be found to use the heat in the form of 90°C hot water, this choice will very likely be the best return on investment for the site. Examples where hot water can be effectively used in many phosphate fertilizer sites include:

- production of desalinated water via an multi-effect distillation (MED) system;
- phosphoric acid concentrators;
- ammonia heaters/vaporisers, etc.

Figure 1 shows a double absorption sulphuric acid plant with hot water heat recovery used to generate desalinated water.

In integrated chemical complexes or smelters there may be other available uses for hot water.

If hot water cannot be used effectively then the next best choice is to create and use MP steam. In a phosphate fertilizer complex the MP steam can be used for phosphoric acid concentration making this an attractive economic option. If some or all of the steam cannot be used there, it can be used via a steam turbine to produce electrical power. The thermal efficiency of the conversion to electrical power is quite low at 17% so this option should be only used as a last resort. From a return on investment perspective the high cost of capital to create the MP steam and generate power will be in the region of 0.035 to 0.06 $/kWh amortised at 6 to 10% respectively over 20 years. The local sale price of power to the grid should be considered before making this investment. Often the sale price is in the 0.04 to 0.06 $/kWh range so the payback time on the incremental investment can be quite long. However if the power produced can be used on site, the power it is replacing can often be in the 0.08 to 0.12 $/kWh range so the rate of return is reasonable.

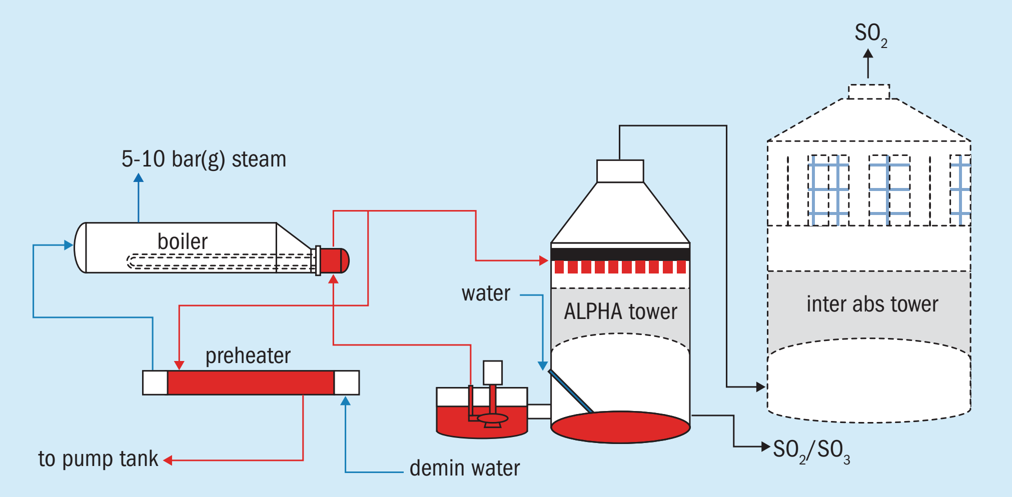
Chemetics ALPHA® System

Chemetics has been long recognised as one of the leading innovators of process and equipment technology in the sulphuric acid industry. In many cases Chemetics has introduced completely new and revolutionary products to the market that have changed the design of the plants throughout the industry. In other cases the innovations have been incremental in nature where Chemetics has looked to improve its own or existing technologies in the market to allow them to be more reliable, more flexible in operation, easier to maintain, easier to operate, less expensive and/or easier to install. In all of these cases the overall goal is to make the client's plant make more money. To make money the plant must operate reliably. Every day of unnecessary shutdown can eliminate a substantial portion of the revenues generated per year by supplemental low grade heat recovery. For most sites a single day of shutdown of the main sulphuric acid plant will lose more money to the site operator than the gain provided by five weeks of operation of the ALPHA system or hot water heat recovery system. With this financial reality it is imperative that the heat recovery system be designed into the plant to allow the main acid plant to operate whether the heat recovery system is operational or not operational. This will ensure that the only revenue lost to the site is the bonus revenue from the heat recovery, not the main process plant. This is even more important for integrated sites where the power, steam, sulphuric acid, hot water produced

Fig 2: **Chemetics ALPHA® system**

PHOTO: CHEMETICS

Fig 3: **Chemetics ALPHA® system process schematic**



Source: Chemetics

by the main sulphuric acid plant is critical for operation of the linked chemical plants. You want to avoid your entire site going off for weeks or even months due to failure of a supplementary, non-core system.

Heat recovery has been a reality in sulphur burning sulphuric acid plants in the fertilizer industry for almost 30 years and the units work quite well. However there are certainly areas that need improvement. In the development of the ALPHA Chemetics focussed on keeping the good aspects of the existing technology while improving on all the other key areas where the existing technology was somewhat lacking. Chemetics has investigated and developed the ALPHA system via almost 25 years of operational, lab and field testing. Key aspects of the ALPHA system improvements are as follows:

- ALPHA shutdown does not shut down main process plant;
- independent start-up/shutdown of main acid plant and ALPHA;
- brick-lined ALPHA tower and pump tank, more resistant to weak hot acid than alloys;
- instrumentation tracks acid process parameters and if not in safe range, all acid and water immediately dumped to acid and water tanks respectively, maintains equipment integrity and forces correct operator training for long term operational stability;
- SARAMET® HT alloy developed for greater longevity of boiler, piping, heat exchangers and more operational range;
- more steam production in higher humidity location than competition;
- safer water dilution into the base of the brick-lined ALPHA tower.

Fig 4: **Hot water heat recovery acid coolers**

PHOTO: CHEMETICS

The ALPHA system (Figs 2 and 3) is designed to be easily integrated and retrofitted into existing or new sulphuric acid plants.

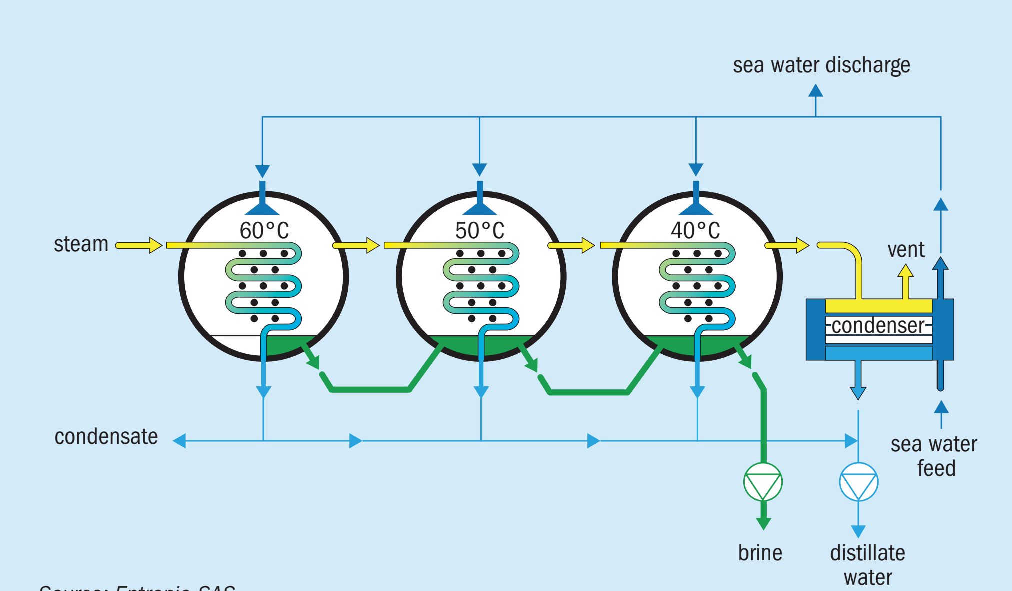
There are many ways to recover and use heat from a sulphuric acid plant. Selection of the most appropriate methods or a particular site is a critical decision for the long term site economics. Chemetics recommend that care should be taken to ensure that no matter which designs are chosen, the plants are designed to allow independent operation of the heat recovery system from the main sulphuric acid plant. Adding in this critical feature is a very small additional cost but will pay for itself within one or two unscheduled plant shutdowns if this flexibility were not provided.

Hot water heat recovery

Hot water heat recovery has been used in sulphuric acid plants for over 40 years. The initial form of this heat recovery was to generate hot boiler feed water for enhanced HP steam generation. Chemetics invented the anodically protected stainless steel shell and tube acid cooler and is still the largest global supplier. Chemetics designs and manufactures all of the Chemetics acid coolers at its Pickering, Canada fabrication shop. By tightly controlling the design parameters and the manufacturing process Chemetics is able to design reliable, long lasting acid coolers even for the more difficult applications like BFW heating and hot water heat recovery (Fig. 4).

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 34 35 36 37 38 39 40 41 42 43 44 45 46
47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64

Fig 5: **Multiple effect distillation (MED) flow schematic**



Source: Entropie SAS

In some site locations fresh water is a resource that is not readily available and is costly to obtain. For these locations it is often viable to install an MED desalination plant that uses the low grade heat in the form of 90°C hot water from the sulphuric acid plant to generate fresh water. The complex can be designed to allow the desalination plant and the acid plant to operate together or be operational when the other plant is off line. This site configuration will be significantly more reliable than a site that first generates MP steam and then uses the power produced to create fresh water via a reverse osmosis type plant. A pair of acid coolers is far less complicated to operate than an ALPHA system. The operating risk is also significantly lower as acid leak detection and mitigation strategies are much simpler and more reliable in heat recovery acid coolers when compared to steam generation systems using hot sulphuric acid. MED distillation is far simpler to operate and maintain than a sea water reverse osmosis (SWRO) plant. Operational availability over the lifetime of the plants can be in excess of 355 days per year as a long term average with the hot water and MED combination. This compares to the currently achieved industrial norms of slightly higher than 330 days per year operation for common MP steam based heat recovery systems.

MED distillation is a process where a series of effects are used in a progressively lower vacuum to vaporise steam from salt water. The water that is created is relatively pure and can usually be used directly as process or boiler feed water without chemical treatment. Figure 5 shows a typical flow schematic for an MED system. As the number of effects is increased it is possible to generate greater quantities of fresh water using the same quantity of heat as an input o the process. 0.5 barg steam is the normal input to the process. This steam can be generated from the 90°C hot water from the sulphuric acid plant by flashing off the steam in a vacuum chamber. It is possible to generate up to 6 m³ of fresh water per tonne of sulphuric acid produced.

Outotec case studies

While modern sulphuric acid plants use technologies to apply the energy available from sulphuric acid plants for the generation of steam or electricity older plants built some two or three decades ago were not optimised to achieve the same levels of heat recovery efficiency.

In old sulphuric acid plants often the energy from conversion and absorption is wasted partly or totally to the environment reducing the efficiency of the plant significantly.

The Sankey diagram of a typical sulphur burning plant shown in Fig. 6 illustrates the enormous loss of energy when only the energy from sulphur combustion is used to generate high pressure steam. This steam amounts to only one third of the available energy whereas approximately another 25% could be gained from the conversion processes and about 40% from the absorption processes.

In the following two recently executed case studies, Outotec reports on how state-of-the-art technologies have been applied to facilitate this potential to increase plant efficiency.

The first case study refers to a metallurgical acid plant built more than 20 years ago. The project began with a study to identify the options to recover excess process heat available from the sulphuric acid plant and to convert such heat into steam or other useable heat carriers. The heat recovery options were considered for the conversion gas section and for the absorption circuits and heat recovery data were evaluated for different plant loads, considering varying gas flows and SO_2 concentrations typical for the daily operation.

Within the gas circuits the energy is available at a fairly high level, allowing the generation of steam of up to approximately 80 bar. Economisers for boiler feed water preheating may be used to "shift" energy to such a high level.

The absorption tower acid circuits are typically operated at temperatures between 70 and 120°C; as a consequence the temperature level for such heat recovery is limited to approximately 80-100°C. The "consumption" of such low level energy therefore is limited to boiler feed water preheating and to other heating purposes like refinery liquid circuits etc.

On the basis of these limiting factors the following targets were defined:

- high pressure steam production using direct heating from the process gas after a catalyst layer;
- production of high pressure steam >40 bar g.

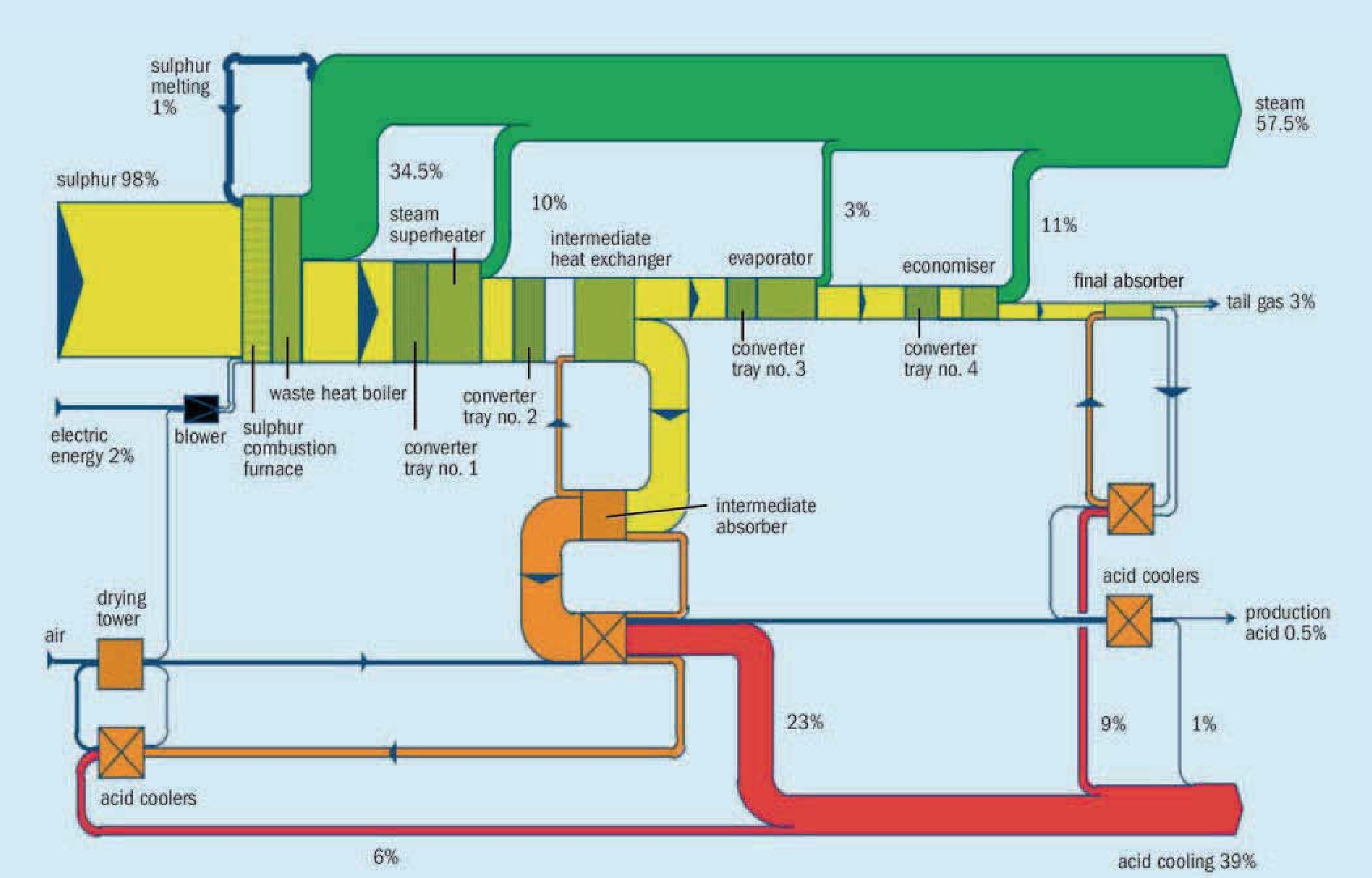
With respect to the existing design the following heat recovery systems were recommended, agreed and the engineering executed.

The high-pressure steam system uses the heat generated in a catalyst layer to produce steam. The heat is transferred from the SO_3 gas in a water tube type boiler to water, which is evaporated at the desired pressure of >40 bar g.

The design also includes the original SO_3 cooler (gas-gas heat exchanger) as a backup option.

Additional heat is recovered as the feed water is preheated by a heat exchanger in the absorption tower acid system. By preheating the water from approx. 65°C to 100°C, the acid is cooled down and delivered back upstream of the original acid cooler.

Fig 6: **Sankey diagram of a typical sulphur burning plant**



Source: Outotec

Further preheating of the boiler feedwater is performed using the available heat of the SO_3 gas in parallel to gas/gas heat exchangers. Also included in this step is the superheating of the generated high pressure steam by the same gas. The SO_3 gas is cooled down to about 160°C and is then conveyed to the final absorption tower.

The resulting steam production that can be achieved varies with the operational state of the sulphuric acid plant considering the specifically available heat in the different modes.

The different modes vary in gas flow rate and SO_2 concentration of the feed gas. All operational scenarios have been considered, special care was taken in the design not to impact the autothermal operational capacities of the plant.

For basic and detail engineering operating records provided by the client were used to adjust the design accordingly to maximise production.

The second case study, also in a metallurgical complex, consists of two steps. The objective of the first step was the optimisation of energy recovery from a 20-year old metallurgical sulphuric acid plant in combination with a capacity increase project. The second step consisted of a new metallurgical sulphuric acid plant with optimised heat recovery. In both cases Outotec HEROS forms the heart of the heat recovery in the absorption section accompanied by feeding the energy from the catalytic section into the steam system as high pressure steam.

Similar to the first case study, in this plant from the 1990s all excess energy from the catalytic conversion section was rejected to the atmosphere by means of gas coolers, while cooling water was used to remove the energy raised at the drying and absorption section. Therefore in addition to the unused energy additional energy was employed for air fans, water pumps and cooling tower fans to obtain the energy balance of the acid plant. On top of this, the mentioned equipment was high maintenance. As with the capacity increase project, the main waste heat boiler upstream of the sulphuric acid plant was replaced and a high pressure steam superheater and evaporator was integrated into the boiler system to capture the energy from the catalytic section.

The changes to the absorption sections were realised with the Outotec HEROS solution (Fig. 7) to produce low pressure (LP) steam from the heat generated by the exothermal reaction in the intermediate absorption tower. The existing tower was replaced by a new tower combination, consisting of a venturi co-current absorber and downstream conventional absorption system based on a packed tower design. To be able to produce LP steam the HEROS is operated with 98.5-99% H_2SO_4 fed into the venturi at around 200°C. In this operation mode the largest part of the SO_3 contained in the gas is absorbed in the venturi section, which means that the sulphuric acid concentration has to be monitored and controlled in the venturi circuit. For the absorption of the residual SO_3 contained in the gas exiting the venturi only a small amount of acid is fed into the packed bed part of the interabsorption.

The uniform distribution of this acid is performed by means of a special patented,

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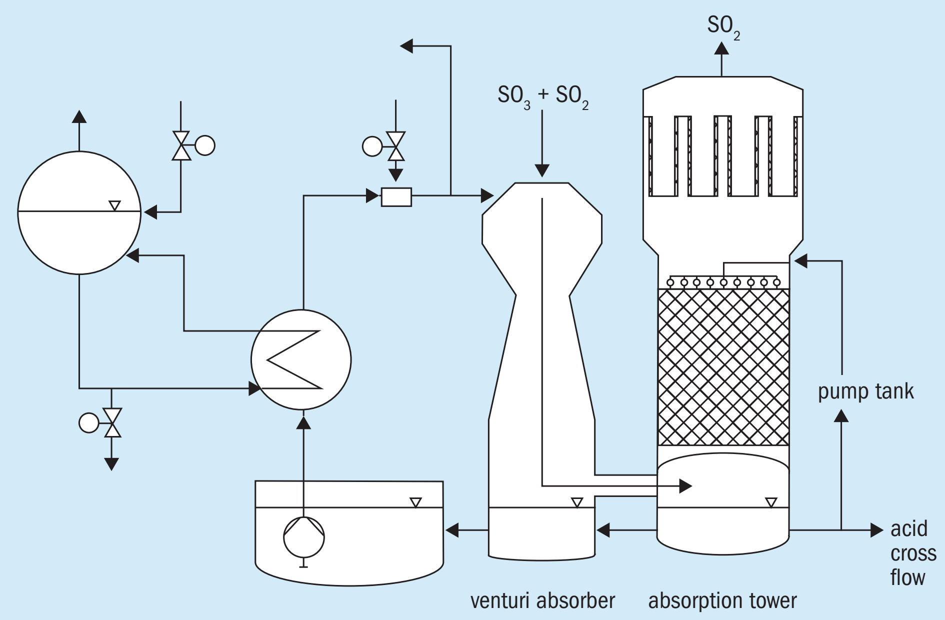
Table 2: **Risk mitigation for heat recovery systems**

			HEROS feature
Availability	Material of construction	All vessels are acid-bricklined to minimise corrosion. Use of Alloy 3033, which allows for wide operational window where no bricklining is possible,	✔
	Flow sheet	Fundamental principle is that HR plant failure will not lead to SAP plant shutdown. The hot acid circuit is completely separated from rest of the acid plant. Intermediate absorption tower is designed for complete SO_3 absorption at 100% plant load.	✔
Safety	Flow sheet Material of construction Automation	In fail-safe mode, automatic draining of hot acid by gravity into brick lined equipment, reducing corrosion to minimum levels.	✔
Operability	Support tools	PORS system for additional high level trending/critical process information requiring operator action to safeguard operating plant integrity. Operator training simulator as optional training measure.	✔

Fig 7: **Outotec HEROS solution**



Source: Outotec

dual operation Outotec FiDi system, a combined system with two different acid headers that is required to achieve an essential benefit of the HEROS solution:

To secure the availability of the sulphuric acid plant and therefore also the upstream metallurgical operation the acid plant can also be operated when the HEROS system is shut down.

A major challenge of any heat recovery processes for LP steam production is the need to operate the absorption plant with very high temperature concentrated sulphuric acid, typically 200-220°C. This temperature level is thermodynamically required for producing saturated LP-steam of e.g. 10 bar. Concentrated acid at this temperature is extremely corrosive unless a very strictly defined acid concentration window is secured.

Regardless of all efforts, e.g. extensive instrumentation, failure of process control cannot be entirely avoided and a large number of acid plants equipped with heat recovery processes have experienced serious problems, in the worst case leading to catastrophic hydrogen incidents that have caused shutdowns for several months.

Acknowledging that the system can fail despite all instrumentation and precautions Outotec's approach with the HEROS solution incorporates a design that ensures any damage can be mitigated in such scenarios.

In addition to the improvements in operational efficiency that are directly linked to the specific flow sheet features and allowing operation of the sulphuric acid plant at full capacity and full high pressure steam production even with the HEROS system in shutdown mode, other risk mitigation factors such as the chosen material of construction (MOC) can be highlighted. Hydrogen incidents, nowadays widely acknowledged in the acid industry, are mitigated by the incorporation of brick-lined HEROS vessels, thus reducing corrosion levels to an absolute minimum and providing inherent safety for the operating personnel. In a world of ever increasing electronic warnings, operators are becoming numbed to the flashing lights and buzzers informing them that some part of the plant requires attention. There is however a need for high level trending/critical process information requiring operator action to safeguard the integrity of the operating plant. Outotec has therefore developed a Plant Operability Reliability and Safety system (PORS) to meet these requirements and a module that is specifically applicable for the HEROS solution is available.

Table 2 summarises the risk mitigation principles of the Outotec HEROS approach.

These features were taken into account when modernising the existing metallurgical sulphuric acid plant from the 1990s with the new heat recovery systems including HEROS along with the capacity increase project.

The successful start-up of the HEROS system proved the operational flexibility optimising plant availability for the whole complex. With HEROS being a reliable solution for heat recovery as LP-steam the customer decided to choose the same configuration for its new sulphuric acid plant in a complete new metallurgical line.

These case studies serve to illustrate that heat recovery can be optimised for each sulphuric acid plant ranging from small add-on modules for feed water heating or for increasing HP-steam production or larger scopes to produce LP-steam with a HEROS solution. This applied to both new lines and the optimisation of a running plant.

NORAM heat recovery options

Process studies can be used to identify opportunities for better energy integration and recovery.

NORAM develops a number of flowsheet alternatives to maximise the efficiency of the plants and to increase energy recovery. Findings can be materialised as reduced electrical power consumption, increased steam production or increased production of electricity.

The Implementation of low pressure drop equipment reduces the energy consumption by the plant blowers (for instance, lower electrical consumption by the main blower) and allows for increased capacity. Examples of such equipment are HP packing, which has about half the pressure drop of conventional packing and radial flow heat exchangers.

Fig 8: **NORAM RF™ gas exchanger**

PHOTO: NORAM

Heat transfer rates in the NORAM RF™ gas exchanger (Fig. 8) are maximised because all the heat transfer surface is fully utilised. Shell gas flow is always perpendicular to the tubes providing the larges film coefficients.

Heat recovery from hot sulphuric acid coolers can be used to preheat water for process use, or for boiler feed water preheating. NORAM SX™ alloy acid coolers provide a very simple, reliable and low maintenance solution for acid cooling since the excellent corrosion resistance of NORAM SX™ eliminates the need for anodic protection.

NORAM designs SO_3 coolers that remove process heat by indirect heat transfer to air. The hot air product can be used to feed combustion furnaces for increased energy production, or it can be used for indirect heat transfer to produce steam. The SO_3 cooler would be a radial flow exchanger.

In addition NORAM can optimise the design of steam systems in acid plants to maximise energy recovery and equipment longevity.

References

1. Fenton M (Chemetics): "Heat sources in a sulphuric acid plant", written for *Sulphur* magazine (Feb 2016).
2. Bräuner S, Schüdde J and Storch H (Outotec): "Case studies on solutions from Outotec for more efficient heat recovery in sulphuric acid plants", written for *Sulphur* magazine (Feb 2016).

Improved WSA plant layout for smelter applications

M. Thellefsen, **M. Møllerhøj** and **E. Eriksson** of Haldor Topsoe describe a new WSA layout for smelter applications featuring an improved heat exchange layout, which replaces the molten salt system with a combination of gas/gas heat exchangers and a high pressure steam system. The new layout significantly improves process control and plant operation, especially for fluctuating flows and SO_2 concentrations.

Due to the nature of the metallic ore and smelter unit operations, smelter off gases fluctuate in both flow rate and SO_2 concentration and the downstream sulphuric acid plant must be able to operate smoothly under these challenging conditions, maintaining low SO_2 emissions and keeping the operation of the plant within the design limits.

At the same time the plant must be designed for lowest capex and opex to ensure a cost-efficient cleaning of the off gases.

Topsoe has sold 14 WSA plants to the non-ferrous industry, focusing on SO_2 concentrations in the low to middle range, i.e. <8 vol-% SO_2, typical for smelter gases from lead, molybdenum, platinum and zinc smelters.

The WSA technology is characterised by efficient heat recovery, which results in so-called autothermal operating points as low as ~2.0 vol-% SO_2, i.e. at this SO_2 concentration the heat released from SO_2 oxidation, SO_3 hydration and H_2SO_4 condensation is sufficient to heat up the cold SO_2 feed gas from the gas cleaning plant without using support fuel.

Traditional WSA-layout

The traditional WSA-layout for treatment of SO_2 smelter gases is shown in Fig.1. The cold SO_2 feed gas is first preheated in a corrosion resistant heat exchanger by hot air from the WSA condenser. The smelter gas can contain some acidic mist, which must be evaporated before reaching the process gas blower, and a small flow of preheated feed gas is recycled to raise the temperature of the feed gas by 10-20°C.

Process gas heating and SO_2 conversion

In the process gas heater, the feed gas temperature is increased by means of heat exchange with molten salt. For start-up situations and operation below the autothermal point, the heating by molten salt is not sufficient, and then the support heater must provide the remaining heat up to ~400°C. This is the optimal temperature for SO_2 oxidation in the first catalyst layer in the SO_2 converter. The heat released in the first catalyst layer is transferred to the molten salt in the interbed cooler, lowering the process gas temperature to the second catalyst bed to ~380°C to ensure the lowest possible SO_2 emission.

The WSA condenser

In the process gas cooler, the SO_3 to H_2SO_4 hydration energy and gas cooling duty is transferred to the molten salt. The process gas temperature to the WSA condenser must be controlled in the range between the sulphuric acid dew point temperature of typically 220-260°C and 290°C, which is the maximum operational temperature for the acid resistant material in the WSA condenser.

In the WSA condenser, the process gas is cooled and the H_2SO_4 is condensed in vertical glass tubes, separating concentrated sulphuric acid product from the cleaned process gas which is sent to the stack. The gas cooling and condensation duty is transferred to hot air, which is used to preheat the cold feed gas. The highly integrated heat recovery system provides the lowest possible autothermal point.

The molten salt system

The molten salt system consists of a salt buffer tank, a pump, process gas cooler, interbed cooler, process gas heater, salt cooler and control valves. The salt cooler keeps a stable salt temperature in the tank and surplus energy from the SO_2 converter is exported as LP steam. The salt temperature in the tank must be kept above the sulphuric acid dew point in the process gas. The salt pump ensures sufficient cooling capacity in the process gas cooler, while the interbed cooler heats up the salt to a maximum temperature of 450°C, such that the process gas heater can increase the process gas temperature to the 400°C at the inlet to the SO_2 converter. The salt solidifies at 145-190°C and thus it is important that the process gas entering the process gas heater is above this temperature.

If controlled correctly the process is robust and highly energy efficient. On the down side, due to the large salt volume, the molten salt system does have a long response time to changes and the start-up of the plant can be time consuming if

Fig 1: **Traditional WSA plant layout for cold SO_2 smelter gas with molten salt as heat transfer media**

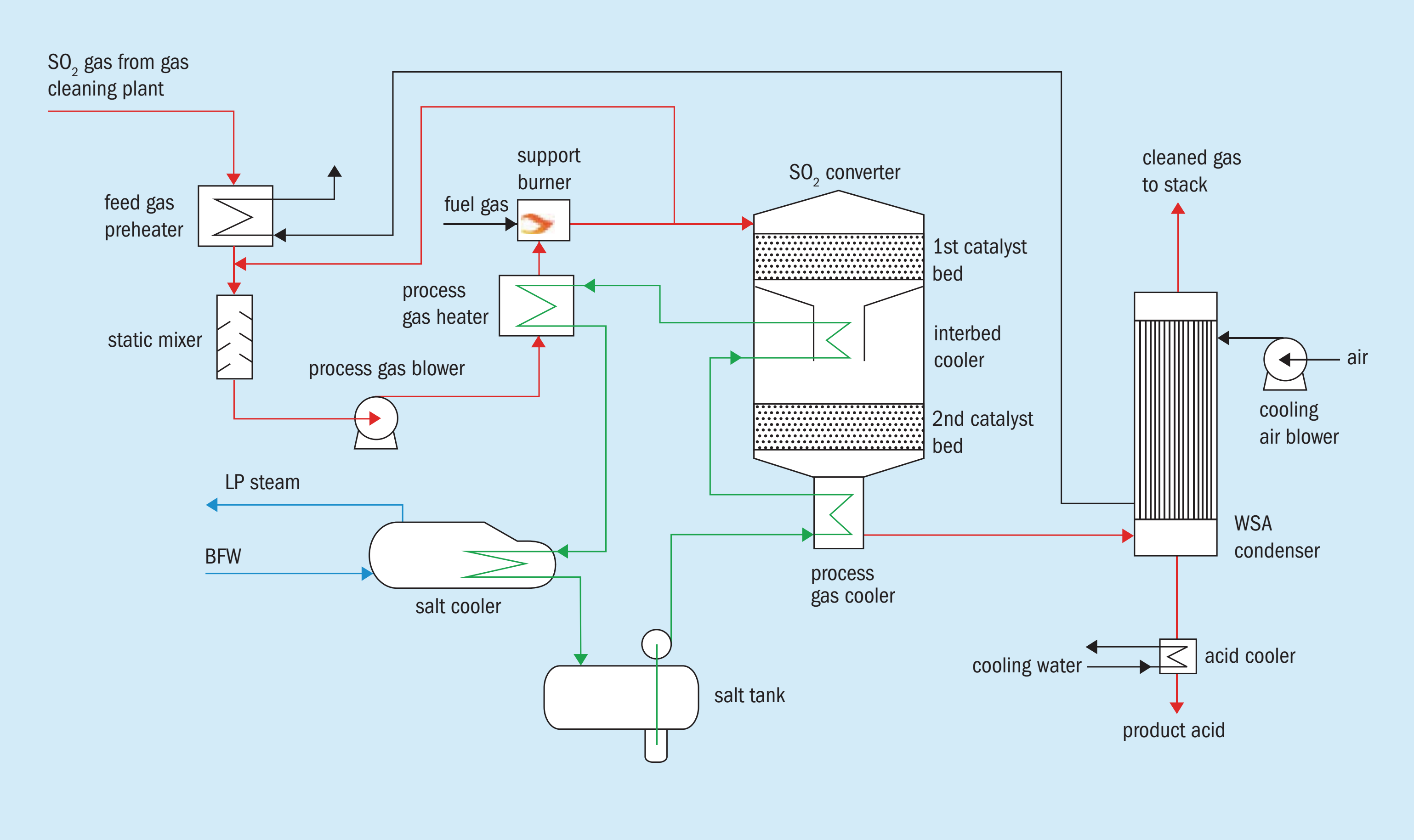
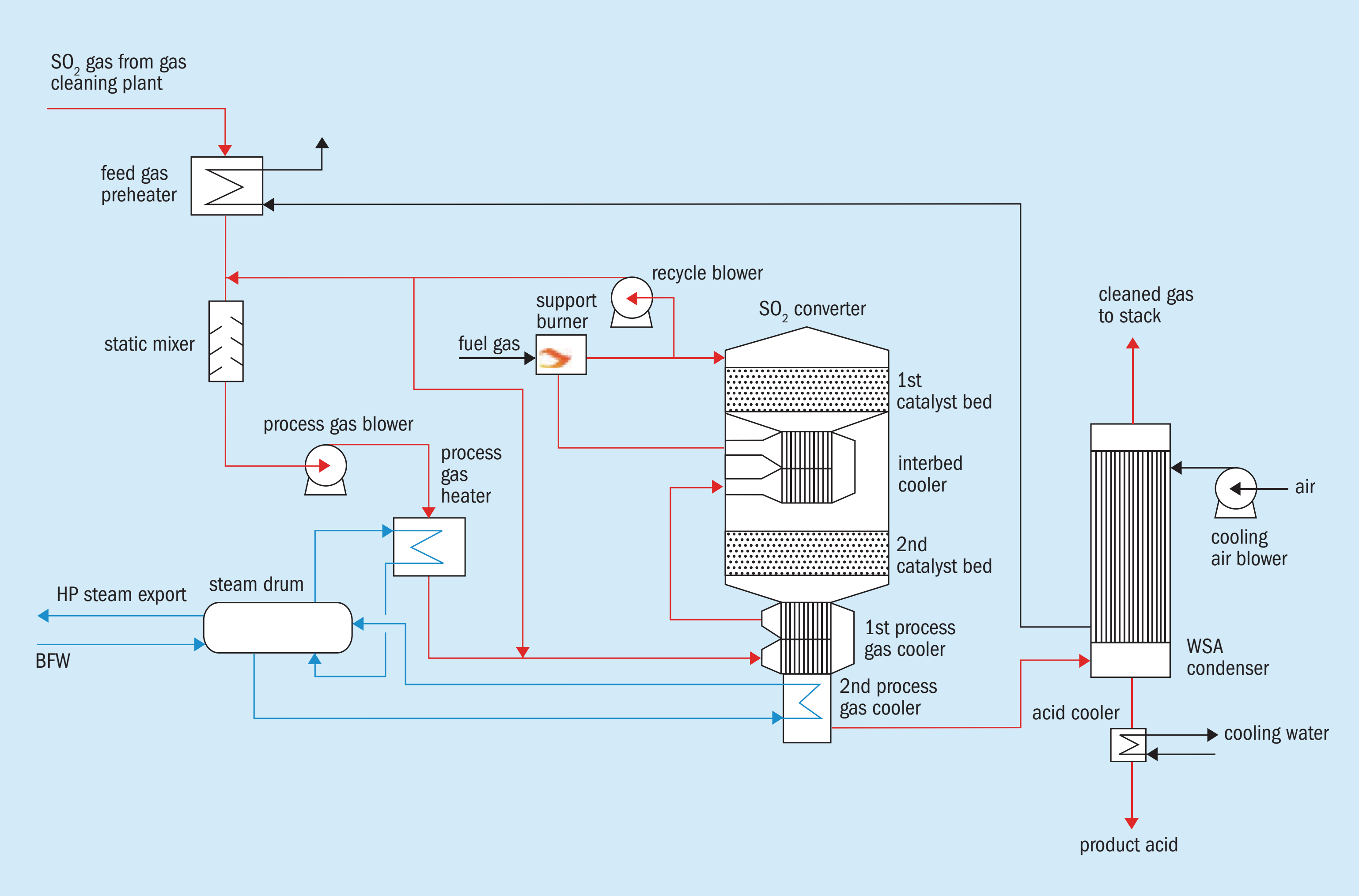


Fig 2: **New WSA plant layout for cold SO_2 smelter gas combining high pressure steam and gas/gas heat exchangers**



the salt has solidified during the shutdown. Also care should be taken to ensure that there are no cold spots where the salt can solidify, thus blocking pipes, valves and heat exchangers. It is therefore mandatory that all salt piping and control valves are heat traced.

Salt leakages do occur from time to time and must be stopped immediately. Over time the melting point of the salt will increase, which results in increased risk of salt solidification in heat exchangers, and eventually the salt must be replaced.

The new WSA-layout

The layout of the newly developed WSA-layout for smelter gases is shown in Fig. 2.

Process gas heating and SO_2 conversion

In many ways the layout is similar to the traditional layout: the cold SO_2 gas is preheated in the feed gas preheater by hot air from the WSA condenser and a little hot process gas is added to the preheated process gas to ensure complete evaporation of any acidic mist present in the feed gas.

The process gas is then further preheated to 245-260°C in the process gas heater, by means of condensation of saturated high pressure steam produced in the second process gas cooler. A flow of hot unconverted process gas is then added to increase the temperature by 20-30°C to be well above the sulphuric acid dew point temperature of the fully converted process gas. The unconverted process gas is used for heat exchange with the fully converted process gas from the second catalyst bed. The unconverted process gas then goes to the interbed cooler, where it cools the converted process gas to the desired temperature to the inlet to second catalyst bed, while being heated up to the optimal temperature at the inlet to the first catalyst bed.

During start-ups and operation below the autothermal SO_2 concentration it is necessary to increase the temperature of the unconverted process gas by firing fuel gas in the support burner.

The WSA condenser

The fully converted process gas is cooled to the 275-290°C inlet temperature to the WSA condenser by means of the first and the second process gas coolers. In the WSA condenser the process gas is cooled and sulphuric acid condensed by means of air cooling. The hot air from the WSA condenser is used for process gas preheating, decreasing the need for support heat.

The steam and gas/gas heat transfer system

At the first glance, the new layout looks more complicated than the traditional layout: an extra heat exchanger is added. However, when analysing the layout it becomes clear that the new layout will be simpler to control as the high pressure steam system is more or less self regulating: the second process gas cooler is a natural circulation boiler, which has a high flexibility for steam production and therefore is very robust against fluctuations in inlet temperature and flow of process gas. This means that the process gas temperature out of the second process gas cooler is much dampened compared to the fluctuations at the inlet of the cooler. Similarly the process gas heater is also self-regulating as the flow of saturated steam is controlled by the required duty for the process gas heating. Any excess heat from the WSA plant is exported as saturated high pressure steam.

The interbed cooler and first process gas cooler are gas/gas heat exchangers of proprietary design and designed for low pressure operation. By intelligent control of bypasses of unconverted process gas around the coolers, it is possible to maintain the correct temperatures to the inlet of the catalyst layers, ensuring optimal SO_2 conversion.

The new layout is easier controlled during process gas fluctuations as e.g. variations in process gas flow will immediately effect both the heat required for heating up the unconverted process gas and the cooling of the converted process gas and with the gas/gas heat exchangers the temperature variations in and out of these units will have less fluctuations compared to the salt heat exchangers. Also the steam circuit is only mildly affected by the change in e.g. process gas flow as the delay in steam production compared to steam consumption is very short and the second process gas cooler has a large capacity for steam production.

Comparison of the two WSA-layouts

The two layouts provide similar SO_2 conversion efficiencies and the heat recovery is almost identical, i.e. the autothermal SO_2 concentration is ~2.0 vol% SO_2 for both layouts.

The main difference lies in the start-up, control and maintenance of the plant, where the new layout is superior in all three categories.

Another benefit of the new layout is a reduced cost of equipment. Even if the new layout has two process gas coolers, the total heat exchanger area is significantly reduced. With the lower operating pressure of the gas/gas heat exchangers and simpler design compared to the salt coolers, the overall cost is reduced by 5-10% for the new layout. This value is based on a WSA plant treating 30,000 Nm^3/h process gas with 3.4 vol-% SO_2; for other process gas flows and SO_2 concentration the saving may be different.

Due to a slightly higher process gas pressure drop, the power consumption of the process gas blower will increase, but that extra cost of power is countered by replacement and disposal of spent heat transfer salt. Since high pressure steam has a value, the new layout could become better than the traditional layout with regard to both capex and opex.

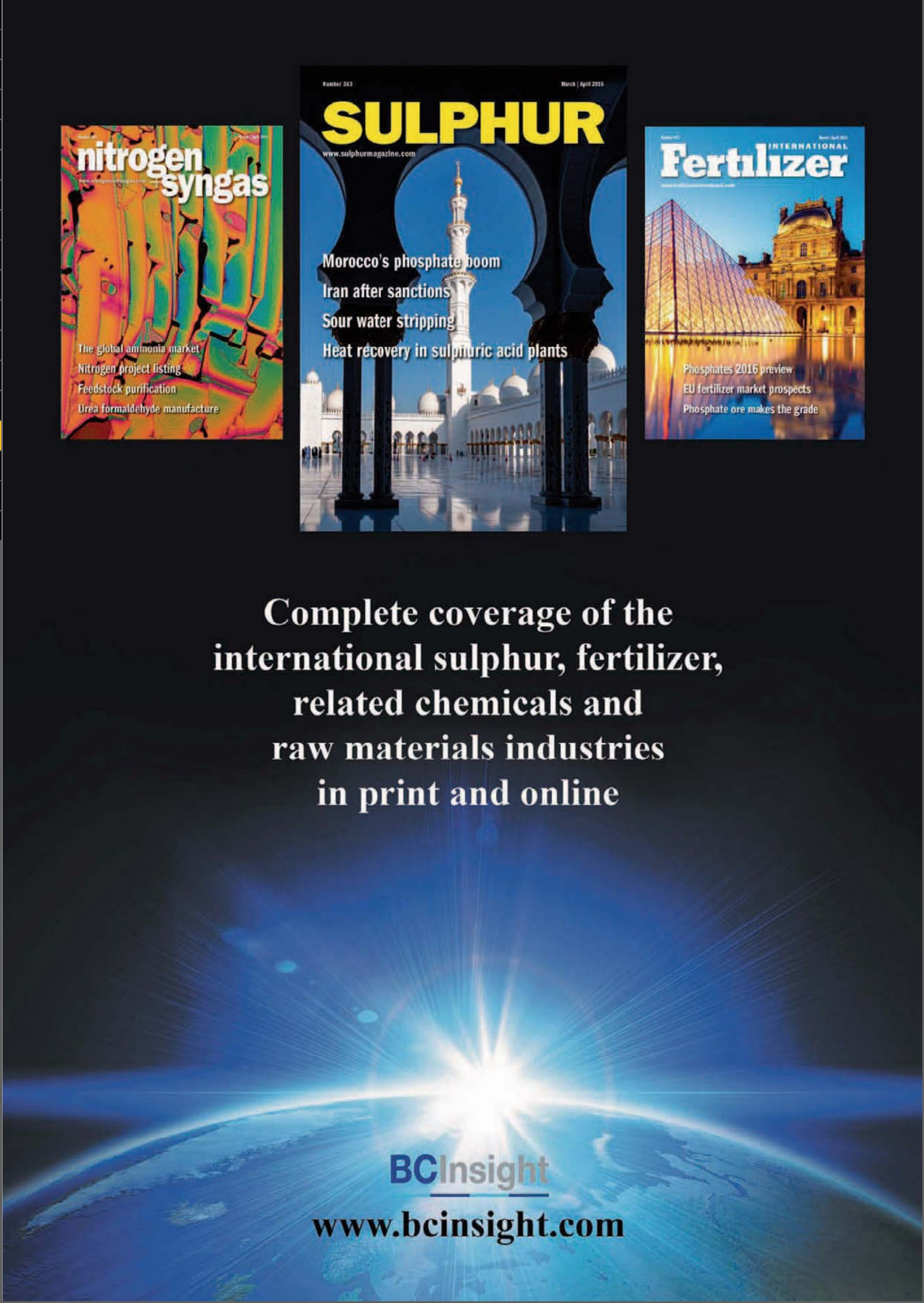
Since the steam system is operating at high pressure, the requirements for the boiler feed water is increased, whereas the traditional layout could accept poorer quality water in the salt cooler. On the other hand, the higher quality steam produced may prove beneficial for being used in other units outside the battery limit. If there is no need or market for high pressure steam, the steam will be condensed and the condensate recycled.

For most of the 140 WSA plants licensed so far, the high pressure steam system is the backbone of the heat exchange layout and the system has proven very reliable in operation.

Conclusion

Topsoe's newest WSA plant layout for treating cold SO_2 smelter off-gases provides a more stable operation than the traditional layout with molten salt as heat carrier. The new combined steam and gas/gas heat exchange solution offers a much faster response to changes in the feed gas flow and SO_2 concentration at a lower cost, while maintaining the high energy efficiency of the plant.

The new layout is Topsoe's preferred option and is already being offered to clients.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Japan: (also subscription enquiries)
KOICHI OGAWA
O.T.O. Research Corporation
Takeuchi Building
1-34-12 Takadanobaba
Shinjuku-Ku, Tokyo 169, Japan
Tel: +81 3 3208 7821
Fax: +81 3 3200 2889

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Fax: +44 (0)20 7793 2577
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