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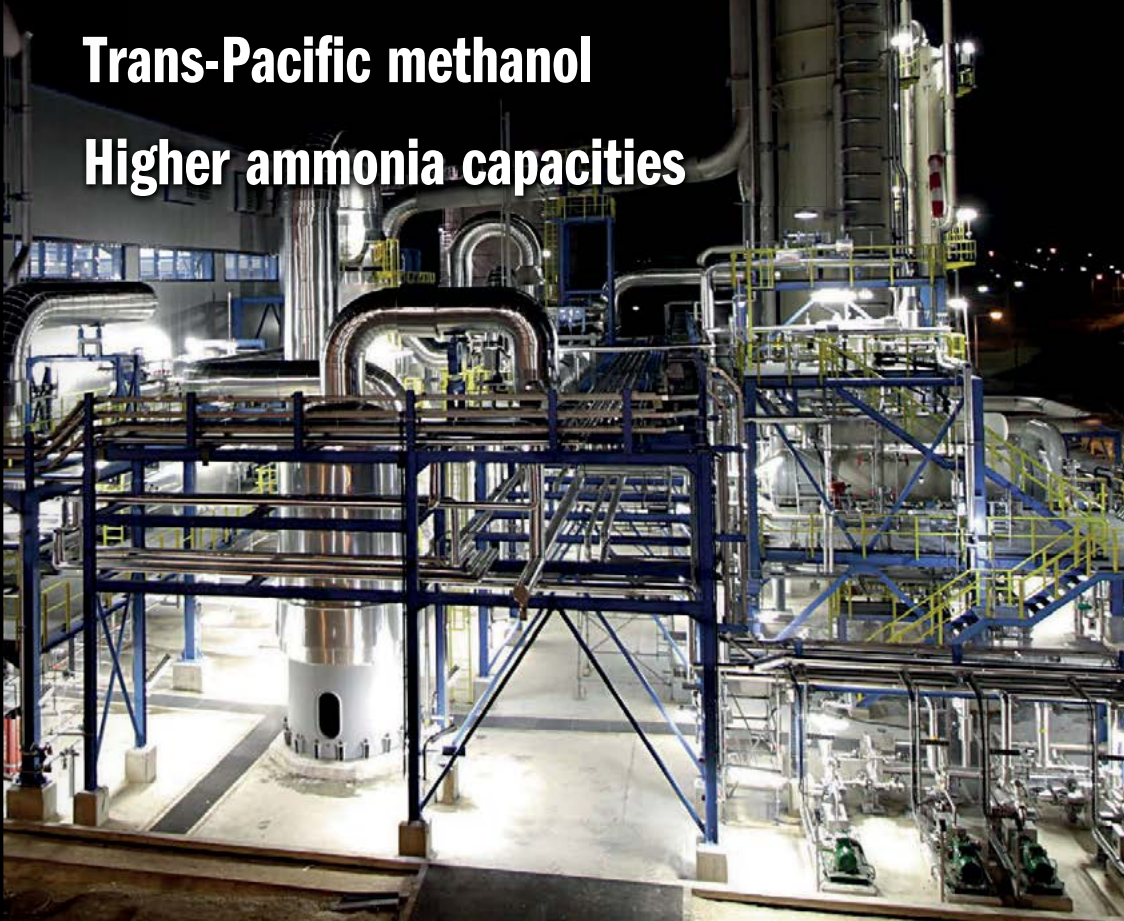
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Nitric acid technology trends

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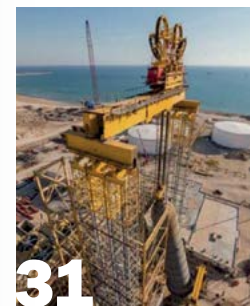
NEW!



Cover: New nitric acid plant at Nitrogénművek in Hungary.
thyssenkrupp Industrial Solutions



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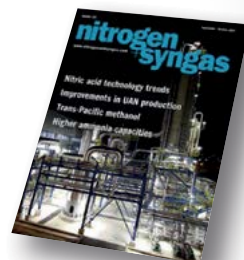
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Can India replicate China's coal gasification boom?



India has always had the potential to travel down the same route.

India shares many common features with China – large populations; 1.3 billion in India to China's 1.4 billion; a reliance for energy on large domestic coal deposits, and relatively small oil and gas reserves leading to large-scale imports; as well as intensive agriculture based on large scale domestic urea production. However, the differences are also significant – India's nitrogen sector was based on naphtha feedstock and now has switched to almost equally scarce and expensive imported liquefied natural gas, while China has instead made full use of its coal reserves via gasification technology. China has also extended the use of coal gasification to produce both more natural gas (synthetic natural gas – SNG), and of course to manufacture methanol on an unprecedented scale. This abundance of methanol has allowed China to blend it as a vehicle fuel, and convert it to dimethyl ether to blend into LPG (something China, like India, makes great use of as a domestic heating and cooking fuel), and, most recently, even to use methanol to olefins technology to produce large volumes of polypropylene from coal. In this way China has managed to control at least to some degree its ever-increasing imports of oil and gas and monetise its own coal reserves. Coal gasification is also far more efficient than simply burning coal, and lowers the carbon emissions per tonne of coal used.

While India has always had the potential to travel down the same route, there has been considerable inertia, much of it caused by India's own rather unfortunate experiences with coal gasification: the country's two coal-based ammonia plants having suffered from all manner of technical issues and eventually being closed down. Attempts to develop a new, more modern large scale coal-based ammonia plant have dragged on for years and at one stage became mired in political scandal over the allocation of coal blocks. However, India seems to be – belatedly – coming around to the idea of trying to pursue a Chinese model of coal-based development. As discussed in our Syngas News section on page 16, India's Transport Minister Nan Gadkari is now pushing

the use of coal-based methanol as a domestic fuel for vehicles. The first step, according to a report Gadkari commissioned by think-tank NITI Aayog, would be a large-scale integrated coal gasification based complex which produced both power, ammonia/urea and methanol, allowing its scale to bring down the cost of each individual commodity.

In theory, it seems a perfectly reasonable plan, and China's success has meant that there is now more international experience than ever available to any developer of a coal gasification project. However, some argue that while China has a very centralised government bureaucracy which is able to push through large scale infrastructure projects, and which often has to pay little heed to local objections or competing policy objectives, India's democratic and decentralised system often has a way of complicating even the simplest things. India's coal industry is already finding it hard to meet domestic needs and the world's fourth largest holder of coal reserves ends up importing coal to meet its requirements, hampered by mining rights, regulatory hurdles and land acquisition problems. Likewise, other countries that have tried large-scale coal gasification have run into difficulties – the US and Australia with costs and financing, Vietnam with technical as well as financial issues. Nevertheless, both South Africa and China have proved that it can work, if the political will is there. The big question is whether such a programme can survive India's fractious politics. ■

Richard Hands, Editor



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Price trends



MARKET INSIGHT

Laura Cross, Senior Analyst, Integer Research, assesses price trends and the market outlook for nitrogen.

NITROGEN

The urea market remains chronically over-supplied and demand expectations are below previous years as we have commented on in previous issues. A small signal of change appeared in July, however, as Egyptian price ideas increased and sentiment elsewhere began to follow. Furthermore, MMTC India announced an import tender which allayed concerns that Iranian supply would flood into markets such as Europe and Turkey over the summer months. Most Iranian material in July and August was expected to be shipped to India under the MMTC tender which closed on 20 July. This absorbed a healthy amount of liquidity from Iranian producers, although there is also Arab Gulf product to contend with and if India chooses not to buy further volumes before September this would leave Arab Gulf producers with little other choice for export destinations. Under a new buying approach, India is looking to purchase around 500,000 tonnes under each tender. The country's import agencies plan to tender more frequently and buy smaller quantities. However, speculation is already rife that the buyer may pick up more tonnes if prices are low. While demand from India has provided some upward sentiment for the urea market, it is however, only a tem-

porary reprieve. Beyond the higher price ideas and import tenders, which are not expected to last, the outlook for the urea market continues to be extremely bearish, with a tentative stabilising in prices expected to play out over the next quarter. There has been renewed talk of Chinese import requirements however we do not expect substantial quantities to be required given the imminent ending of the domestic season and the likelihood of increased production over resorting to imports.

The Chinese Ministry of Industry and Information Technology (MIIT) has announced a research programme aimed at optimising the country's fertilizer capacity. As part of the programme, fertilizer capacity considered to be outdated would be forced to close but new and advanced capacity will be encouraged to expand. The government will subsidize the construction and development of more advanced capacity, if the owners are willing to switch off obsolete capacity.

The China Nitrogen Fertilizer Association has reported that about 13 million t/a of urea capacity is likely to close between 2015-2020. Under the MIIT research programme, it is envisaged that overall nominal fertilizer total capacity volume won't change significantly, despite these closures, if replacement and expansion of other capacity is encouraged and subsidized.

In general, fertilizer capacity considered to be outdated is mainly located in central cities or close to residential areas. These plants are required to be relocated because of uneconomic and emission issues.

In the ammonia market, the latest Tampa contract price announcement at press time, for August, has been the main talking point, as the price was dropped drastically by \$50/tonne to \$190/tonne c.fr. This level is the lowest the contract price has been since July 2009, highlighting the fundamental weakness of the market in its present state. Weak demand throughout the summer has left prices in the US slashed in order to move tonnes, and it is unlikely that any significant upward trajectory will be seen until further fill-buying begins to take place in the Autumn.

In Southeast Asia, there have been a variety of outages, both planned and unplanned, which could serve to add some support to global price benchmarks over the coming months but this has yet to come to fruition and ammonia prices remain far weaker in nitrogen terms than urea. Malaysian producer Petronas' planned outage at its Kerteh facility has been extended to a new start-up date in early September and will have no spot availability as a result. This was echoed in Indonesia, where KPI is also extending its planned outage due to technical problems and will similarly be delayed until September.

Middle Eastern producers are the only global producers experiencing some respite from the downtrend in pricing, as the region has been experiencing some strong buying interest, aiding in a slight uplift in pricing. This began when Indian producer MCFL was offline in July and will likely continue over the coming months as buyers may be looking for volumes due to the outages in Malaysia and Indonesia. Pricing in the region averaged \$184/tonne in July but had moved up to \$207.5/tonne by the second week of August.

In July, typical margins for producers across all regions continued to decline dramatically, as global pricing benchmarks continued to deteriorate. Typical margins for producers in Russia fell to \$6/tonne off the back of falling ammonia prices, which is a drop of \$40/tonne, keeping many producers in the region focusing on domestic clients instead of exports. Producers in NW Europe fell to below breakeven, with the typical margin resting at -\$45/tonne, due to weak pricing and some slight increases in feedstock costs.



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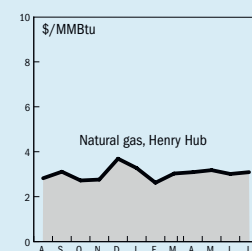
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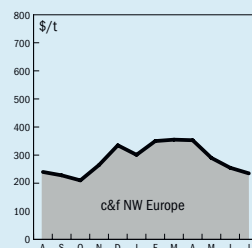
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END OF MONTH SPOT PRICES

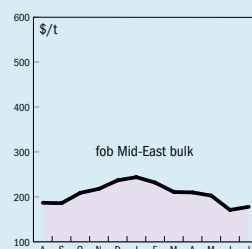
natural gas



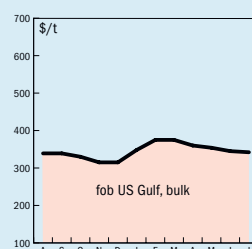
ammonia



urea



diammonium phosphate



MARKET INSIGHT



Mike Nash, Global Business Director, IHS Chemical, assesses the market for methanol.

METHANOL

Methanol prices increased slightly in July; an increase in both ethylene and propylene prices went some way towards maintaining MTO affordability despite higher Asian methanol price – in July \$13/t higher, in the range of \$280-288/t, c.f.r. China average prices were up \$18/t at \$278-290/t, c.f.r. Methanex's posted APCP for August is \$320/t, a rollover of June and July. European spot prices (T2 f.o.b. Rotterdam) for July were down €10/t from June at €258/t over the month. Methanex posted its 3Q 2017 West European Contract Price at €320/t, f.o.b. Rotterdam T2, a decrease of €130/t on the previous quarter. The 3Q 2017 West European Contract Price was settled at €315/t, f.o.b. Rotterdam T2, down €90/t. The ongoing suspension of duty on methanol arriving into the EU implemented by the European Commission is likely to remain for the foreseeable future. The official posted reference prices from the two main US producers for August were \$1.13/gal for Methanex (a rollover from July) and \$1.10/gal for Southern Chemical Co. (up 2 cents from July); these prices equate to \$376/t and \$366/t respectively. IHS Chemical's contract net transaction price for August was officially posted up 0.9 cents at \$1.116/gal (nominal \$371/t).

On the supply side, global utilisation rates in July were around 1% higher than June 2017, at around 72% of nameplate capacity, or 85% of effective capacity. In Iran, gas issues resulted in all units operating at slightly reduced rates in July. In Saudi Arabia, Ar Razi V re-started in mid-June after a planned maintenance outage and the one line at Sirte in Libya ran at full rates during the month. Viromet in Romania is idled due to high gas prices. Tomsk re-started on 2nd July after a maintenance outage that included a capacity increase to 1.0 million t/a. In Trinidad, with not all units running, the average operating rate was around 89% for July. Methanex's Chilean unit is estimated to have run at around 76% capacity for most of July, before dropping to around 27% at month end as seasonal gas allocations took their toll. North American units operated at 90% overall, with maintenance at

one of Methanex's Geismar units continuing into July. In Southeast Asia, Petronas's no. 1 unit resumed normal production in early July. Brunei Methanol Company came offline unexpectedly in 2H July but was operating at high rates by the end of the month. In China, overall capacity utilisation was relatively stable compared with June, at around 56%. Shaanxi Yulin Natural Gas Chemical's gas-based unit and Boyuan Unichem remained offline for the whole month. Yankuang Rongxin shut down unexpectedly in early July. Shandong Mingshui's new 600,000 t/a coal-based capacity started up in the first week of July.

Methanol demand in the Americas remained strong, especially into fuels and formaldehyde. In India, port prices started the month at an average of \$225/t c.f.r. T1 and finished the month \$17.5 higher at \$242.5/t due to a gradual reduction in inventory towards the end of the month following temporary gas restrictions in Iran, which restricted supply. The ongoing monsoon continued to have a negative impact on consumption, particularly towards the end of July, when flooding resulted in extremely low demand. The new system of taxation, introduced on 1 July, caused trading to be suspended for some of July, as taxes must now be paid by importers of product rather than the buyers, leading to contract terms having to be re-written and importers attempting to amend payment terms from 75 to 45 days.

European demand for methanol was stable in July. Demand into formaldehyde was seasonally healthy, although this typically declines towards the end of the third quarter. Rhine water levels at Kaub rose by the end of July, starting the month at around 150 cm and finishing the month at a much more comfortable level of 265 cm.

Chinese methanol consumption into the MTO sector healthy at an average rate of around 76%. Zhongmei Mengda re-started in mid-month after a turnaround. Fund Energy Changzhou has been offline since the end of March, with no re-start date announced. Shandong Levima resolved its issue with an air separation unit and its MTO unit was back to high rates by mid-July. Demand for formaldehyde continued to be impacted by a seasonal downturn and by environmental pollution controls in north China.



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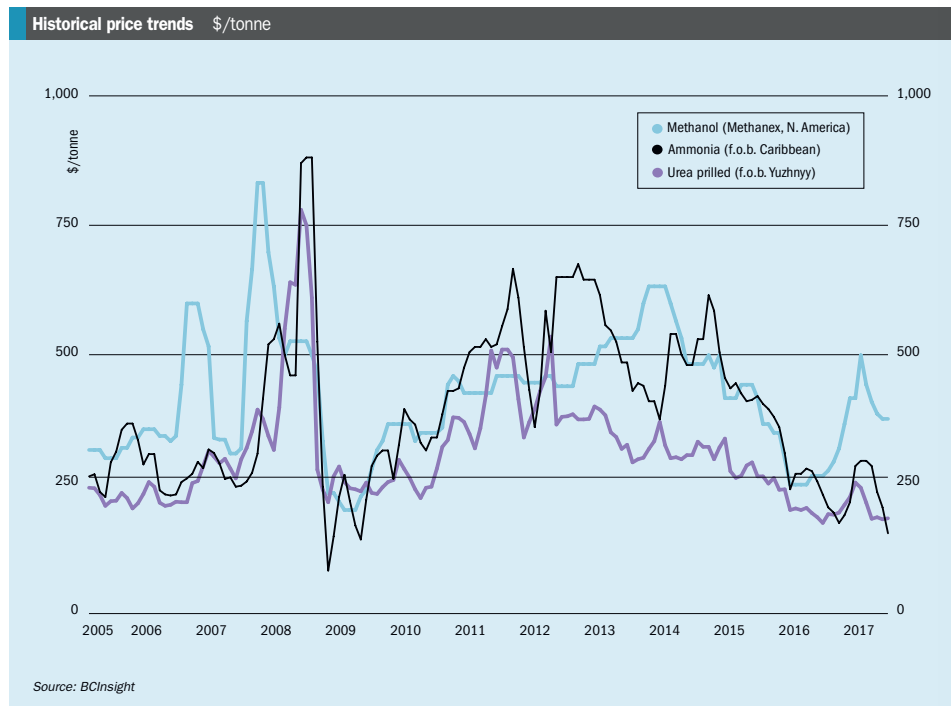
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Market outlook



AMMONIA

- Ammonia prices are at their lowest levels since 2009 as weak demand has led to price cutting to move tonnages. Ammonia market sentiment remains weak and prices continued to trend downwards in early August.
- Demand remains weak in most regions globally, and sellers are still reported to be long in most areas as a result.
- However, some unexpected supply outages in southeast Asia could spur some minor upticks in demand over the coming months, and could therefore lead to some upward movement in pricing, especially for producers in the Middle East which are the supplier of choice East of Suez. Petronas has extended a planned outage to September and Kaltim Parna Industri in Indonesia is also prolonging a maintenance shutdown for technical reasons.
- US ammonia demand is forecast to remain weak until seasonal buying kicks in after the summer.

UREA

- Several urea capacity expansions are due for completion by the end of 2017 and this will extend the gap supply-demand gap.
- It is expected to further weaken price sentiment despite recent increases in price ideas.
- However, seasonal buying and curtailments elsewhere could offset the new supply additions in the short term.
- Indian import activity is forecast to be lower than in previous years owing to healthy opening stocks this year which has reduced the country's overall import requirement.
- Any Indian tenders that are announced are highly likely to be met by Iranian and Arab Gulf product, with urea producers in these countries having the highest levels of liquidity and most competitive price ideas.
- Closures to Chinese capacity are likely to be balanced by larger, more modern facilities elsewhere.

METHANOL

- Increasing ethylene and propylene prices have helped support Chinese MTO demand and hence regional demand for methanol and methanol pricing across Asia.
- New MTO capacity continues to come on-stream, and is likely to lead to further step changes in Chinese methanol demand.
- In general however methanol prices remain at relatively low levels, and European 3Q contract prices have suffered a very sharp correction after being overpriced for much of 2Q 2017.
- Methanol availability in the Americas has not been badly impacted by shutdowns in Trinidad. In spite of the ongoing situation in Venezuela, methanol output remains steady there, and North American capacity is operating at high rates.
- Chinese gas-based plants are setting the price floor at around \$260-280/t, but coal prices are rising there.

Nitric acid catalysts and process plants ...

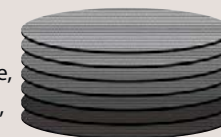
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RUSSIA

New joint venture urea plant for Togliatti

PJSC KuibyshevAzot, a leading Russian fertilizer producer, will partner with Italy's Maire Tecnimont, to build and operate a joint-venture greenfield urea plant at KuibyshevAzot's existing site at Togliatti, in Russia's Samara region. KuibyshevAzot will hold a 68% share of the new company, and Maire Tecnimont the remaining 32%, via its fully owned project development company MET Development. Total investment cost for the project is estimated to be around €160 million. The urea plant will have a capacity of 1,500 t/d (495,000 t/a) and will be based on Urea 2000Plus® pool reactor technology from Maire Tecnimont subsidiary Stamicarbon. The finishing section will also be based on Stamicarbon's fluid bed granulation technology. Both parties have begun basic engineering and project documentation works in collaboration with Russian Design Institute JSC NIIC.

Project finance will be arranged via foreign banks with the support of the Italian Export Credit Agency, SACE. In addition, SIMEST is also evaluating the possibility of sharing MET Development's investment in the project. Following financial closure

EuroChem looking to Chinese joint venture

EuroChem Group says that it has signed a memorandum of intent with the China National Chemical Corporation (ChemChina), China's largest chemical corporation, for a potential partnership to jointly produce industrial products in Russia. Both companies have agreed to form a working group to further evaluate the technical feasibility of the project and will carry out due diligence. At the moment the agreement covers production of propylene oxide (propylene glycol and polyols) and isocyanate (MDI/TDI) at EuroChem's Novomoskovskiy Azot plant in the Tula region of Russia. EuroChem says that this will allow it to optimise production capacity at this facility by integrating the additional downstream production process, which will use existing by-products and infrastructure. ChemChina will supply production technology for the estimated \$500 million project.

Dmitry Strezhev, EuroChem CEO, commented: "We are looking forward to working with ChemChina. This mutually beneficial partnership would complement our existing production, whilst also ending Russia's reliance on imports through this first production facility of propylene oxide and MDI/TDI in the country."

KBR to revamp ammonia plant

KBR has secured a contract to provide licensing and basic engineering design for a revamp of Dorogobuzh JSC's ammo-

nia plant in Russia's Smolensk region. The plant was originally constructed by KBR. Under the agreement, the company will license its ammonia technology to increase the capacity of the plant to 2,100 t/d (690,000 t/a) using KBR's Reforming Exchanger System (KRES) in conjunction with its True Cold Wall Add-on Converter. KBR says that this will enable Dorogobuzh to produce low-cost ammonia in a bid to meet increasing demand.

"The application of KRES in conjunction with KBR's reliable add-on ammonia converter provides low-cost revamping to Dorogobuzh and numerous similar plants in the region," KBR technology and consulting president John Derbyshire said: "The application of KRES in conjunction with KBR's reliable add-on ammonia converter provides low-cost revamping to Dorogobuzh and numerous similar plants in the region. We are pleased to contribute KBR's innovative technology solution for natural gas monetisation in Russia."

Dorogobuzh, part of the Acron Group company, anticipates spending \$75 million on the upgrade project, which is scheduled to be completed in the second half of 2019.

INDIA

FACT launches new ammonia barge

Fertilizers and Chemicals Travancore (FACT) has launched a 51 metre barge designed for the transport of liquid ammo-

and completion of FEED works – anticipated in Q2 2018 – the JV will enter into a lump sum turnkey (LSTK) EPC contract with Tecnimont for construction of the plant, with completion anticipated in 2021.

Aleksandr V. Gerasimenko, general director of KuibyshevAzot, commented: "A new urea plant project implemented in Togliatti together with our Italian partners will allow KuibyshevAzot to attract state-of-the-art technologies to the company's industrial site, increase output of the product demanded by consumers and strengthen the company's position in the mineral fertilizer market." Pierroberto Folgiere, Chief Executive Officer of Maire Tecnimont Group, commented: "This is a significant step forward in our industrial strategy, which confirms the reliable value proposition of an integrated company such as Maire Tecnimont, offering not only technology and EPC best in class expertise, but also project development capabilities. We are very keen to strategically cooperate with a prestigious industry leader such as KuibyshevAzot in one of our core business areas, fertilizers."

nia. The *FACT Pragatiyan* was built by the company's own FACT Engineering Works at its yard in Pallurthy, and was formally launched in June by Jissy Jacob, wife of FACT production coordination general manager Manuel Zacharias, by breaking a coconut across its bow. The vessel was designed by Cusat and has six ammonia storage bullets, each with a capacity of 32 metric tonnes. It is 9.6 m wide and is powered by twin 192 bhp engines. It will transport ammonia from FACT's storage facility at Willingdon Island and its Udyogamandal plant to its Ambalamedu production unit, and brings the company's combined barge capacity to 200 tonnes.

Safety features include twin safety valves on each bullet, a water spray system throughout the cargo hold area, trolley mounted breathing apparatus and single piece encapsulating full body chemical suits.

ISRAEL

Haifa Chemicals closes plant

Following a failure to secure alternative ammonia storage, and the forced closure of the company's 12,000 tonne ammonia storage tank at Haifa, Haifa Chemicals says that it will be shutting its Haifa Bay plant and firing the 400 workers there. Israel's Supreme Court has ruled that the tank must be emptied and shut down by September 18th, exhausting all appeals against the closure. The tank has been a bone of contention for years, with local

authorities claiming that the storage tank is a danger to life in the port in the case of accidental or deliberate breach, and withdrawing its operating permit, and with Lebanon-based Hezbollah repeatedly making threats against the facility.

Haifa Chemicals also operates a potassium nitrate plant in the south of the country, employing another 400 people, which will remain open. There has been discussion of constructing a small scale ammonia plant out in the desert to supply this unit, but so far no work has begun on construction. Other alternatives include bringing ammonia in by truck, and using a floating ammonia storage vessel moored offshore. The Environmental Protection Ministry has said that it was considering approving a 2,500 tonne offshore vessel, but this too has become caught up in wrangling with local authorities and environmental groups.

UNITED STATES

Stamicarbon signs supply agreement with Nooter

Urea technology developer and licensor Stamicarbon, part of the Maire Tecnimont Group, has signed a general framework agreement with the St Louis-based Nooter Construction Company, a provider of welding and mechanical services with application in urea processes. The agreement covers the supply of services regarding on-site welding of *Safurex*® material as well as relining and repair services for high pressure equipment made of austenitic materials for Stamicarbon and its North American customers. Both companies have worked closely together for many years in designing, engineering, manufacturing and delivering high pressure equipment, in particular in reactors and heat exchangers made of stainless steels operable under pressures prevailing in the synthesis section of urea plants. The companies say that this cooperation agreement is "a next step in the long-term partnership between the companies to better serve and assist Stamicarbon's customers in North America."

New ammonia barge for Mosaic

Vigor has launched the first liquefied ammonia transport barge built in the United States for Jones Act (internal US) trade since 1982. The *Harvest* is 156 m long and 29 m wide and took 23 months, 9,000 tonnes of steel and a million man-

hours to construct at Vigor's facilities in Oregon and Washington State. It will be operated by a subsidiary of Savage Companies on behalf of fertilizer manufacturer Mosaic, and will be part of an articulated tug and barge (ATB) unit. The accompanying tug, the S-182 *Vision*, is being built at Nichols Brothers Boat Builders, Freeland, Washington, and will be powered by twin 4,000 hp diesel engines.

"This is a significant project for both the maritime industry and our community," said Frank Foti, Vigor's CEO. "We've worked hard to bring together a family of companies designed to take our complex fabrication capabilities to the next level and bring new work to the Pacific Northwest."

Careful integration of various complex systems was required to support the Harvest's state-of-the-art, onboard re-liquefaction plant that keeps cargo cooled to -27 degrees Fahrenheit.

"It was an honour to partner with Savage and Mosaic on this project," added Joe Corvelli, Vigor SVP and program manager. "It was their close collaboration throughout the vessel construction that helped maintain an efficient and well-executed build."

Training course to follow ANNA conference

Following the conclusion of this year's Ammonium Nitrate/Nitric Acid (ANNA) Producers' Meeting in Austin, Texas from 1st-6th October, software developer ProSim is organising a free one-day event to introduce delegates to its ProSimPlus HNO3, the only software dedicated to the simulation and optimisation of nitric acid plants and nitrous vapour absorption. This course will enable participants to master the fundamental principles of simulation and optimisation of nitric acid plants and nitrous vapour absorption. Trainees will study issues related to nitric acid-specific physico-chemical properties, unit operation modelling, mass-energy balances and numerical aspects. At the end of the day, participants will have the required elements to assess the benefits of simulation and to analyse rigorously a nitric acid production process with flowsheeting software.

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ation in Europe, Middle East, and North Africa. "SNC-Lavalin is pleased to enter into this next phase of the Salalah ammonia project, which is of great importance for SMC and the Sultanate of Oman," added José J Suárez, president of mining and metallurgy at SNC-Lavalin.

INDIA

India revises sales tax on fertilizers

On July 1st this year, India introduced its new Goods and Services Tax (GST), a landmark tax reform which has been touted as the 'largest single tax reform since independence'. The GST will replace over a dozen central and state taxes on goods and services into one national rate for sales tax, and is hoped to boost GDP by up to 2%. All goods, except for precious stones and gold, fall into one of five sales tax brackets: 0%, 5%, 12%, 18% and 28%. Fertilizers were initially pegged to be in the 12% bracket, but hours before the launch of GST the government changed its mind and lowered the tax rate on fertilizers to the 5% band to ensure that fertilizer prices do not rise unduly and farmers' interests are protected. The government wants to double the income of farmers by 2022.

"Regarding the 12% rate of GST (on fertilisers), some felt it may increase the burden on farmers, so the consensus within the Council was to bring the rate down to 5%", explained Union finance minister Arun Jaitley, who chairs the GST Council. Pre-existing fertilizer sales taxes were in the range of 0-6%, and there worry was that a 12% GST rate could have seen retail prices increasing by Rs30-120 per 50kg bag of urea, diammonium phosphate (DAP) and potash in states like Punjab, Haryana and Andhra Pradesh, where there is currently no sales tax on soil nutrients.

UNITED KINGDOM

Johnson Matthey celebrates 200 years

Johnson Matthey reached its 200th anniversary in July this year. The date was celebrated with 70 parties taking place around the world, beginning with sunrise in Japan and lasting until sunset in the United States, and bringing together the company's 13,000 employees around the world, including scientists, engineers, sales teams and other staff across a wide range of disciplines. Tree planting, beach



The MV Cheshire on fire off Gran Canaria.

clean-ups and other community or charity focused activities also featured alongside traditional parties.

The company which became Johnson Matthey was founded in 1817 by Percival Norton Johnson, an assayer of gold and other precious metals. He went into partnership with stock broker George Matthey in 1851. Since then, although the company has moved from its traditional base of precious metals, it retains interests in that area via businesses such as production of platinum catalysts for catalytic converters and nitric acid manufacture. It has also been involved in the development of new battery technologies, advanced materials used in medical devices and hydrogen fuel cells, and had annual sales of more than £3 billion in 2016.

Robert MacLeod, Chief Executive of Johnson Matthey said: "This is a very special year for all of us at Johnson Matthey as we celebrate our 200th anniversary. We continue to make significant progress and our refreshed brand identity reflects the breadth of our science and the positive impact it has on our customers, and on the lives of millions of people around the world every day."

SPAIN

AN ship evacuated after fire on board

The MV Cheshire, carrying a cargo of ammonium nitrate, had to be abandoned on August 12th after a fire broke out in a cargo hold. All 24 crew were winched to safety by the Spanish coastguard – the vessel was off the coast of Gran Canaria in the Canary Islands at the time. The order to abandon ship was given after a

steep rise in the temperature of the ammonium nitrate on board made it dangerous for the seafarers to remain, both because of the explosion risk and the toxic gases released. The ship is owned by UK-based Bibby Line, and had been travelling from Norway to Thailand, and due to make a stop at Las Palmas on Gran Canaria. The ship was steered into open water once the fire broke out.

In a statement Bibby Line Ltd said that the vessel had reached a safe position 45 miles south of Gran Canaria on Sunday after the "thermal decomposition" of the ammonium nitrate, and that local authorities were offering assistance. The crew members, Ukrainian, Russian and Filipino, were said to be "well and in a good condition". The statement added: "The temperature in hold 4 rose to a high enough level to cause damage to the hatch cover and the plan is to use the assistance of a specialist salvage company to provide resources to cool the affected cargo and consequently, professional salvors. Resolve Marine have been appointed to attend the vessel and deal with the cargo problem. The crew of Cheshire have been evacuated and taken ashore for their own safety as wind conditions were so light that the fumes were not being carried away from the ship. A standby tug from the Spanish Maritime Safety Agency will stay with the vessel and an access ladder has been rigged. The vessel is drifting in a southerly direction away from any landmass."

MV Cheshire has been under Bibby Line ownership since March 2012 when she was delivered to the company from a shipyard in China.



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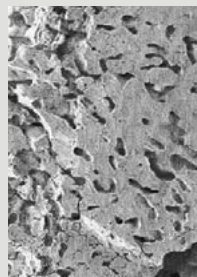
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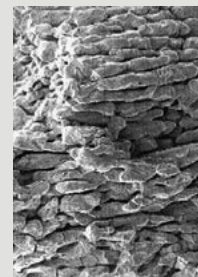
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INDIA

India to move ahead with methanol vehicles and ships

In a bid to promote transportation using alternative fuels such as methanol and ethanol, India's Road Transport, Highways and Shipping Minister Nitin Gadkari says that major automobile producers Volkswagen and Volvo have shown an interest in supplying flexible fuel vehicles to be run on a pilot basis in the country, beginning with 50 Volvo buses designed to run on methanol in the cities of Pune and Mumbai. Other fuels such as ethanol and compressed natural gas (CNG) are also under consideration. The minister said that talks were also ongoing with taxi operator Ola about running taxis on ethanol. "Volkswagen is ready to supply flexible fuel engines for 100 vehicles if an ethanol supply is ensured," he said. The aim is to cut India's huge \$110 billion crude oil import bill. Gadkari said that ethanol could be produced in large quantities in Uttar Pradesh and Bihar, and taxis running on it could save Rs4.5/km (\$0.07) in fuel charges, while Gujarat State Fertiliser Corporation could produce methanol. India is also considering the promotion of alternative fuels in the shipping sector, and vessels could be run on methanol, which the minister said was cheaper than LNG, adding that 500 barges were brought for Ganga which could be run on these if all goes well. Gadkari is also lobbying to lower the sales tax rates on hybrid vehicles and other automobiles that run on alternate fuels. Under the new Goods and Services Tax (GST) rates, hybrid cars will attract a 15% tax rate over and above the peak rate of 28%, the same as those of large luxury cars and SUVs. The automobile industry has expressed concern at the high tax rate on hybrids, arguing

it would dampen government plans to promote green vehicles. A recent report by Niti Aayog report has suggested that India should explore the options of setting up a methanol/dimethyl ether (DME) plant in Iran or Qatar, as these countries have huge reserves of natural gas at low prices. The report, titled 'India's Leapfrog to the Methanol Economy' said methanol/DME produced abroad can be imported to India for its direct application or for further conversion to chemicals like olefins. Although India may end up importing methanol, it is likely to be economically advantageous rather than importing crude, the report said. The report, however, also stressed that India should set up a mega coal-based complex for production of power, methanol and fertilizer in an integrated manner, which would significantly reduce the cost of various commodities produced. According to the report, a task force has been constituted which would work towards the development of overall framework of methanol production, distribution and utilisation in the country. "The government is likely to go ahead with a target of 15% blending by methanol/DME in gasoline/diesel by 2022 which, if achieved, could result in savings of around \$8 billion (based on the 2016-17 price of India basket of crude oil, which is likely to rise in the medium and long term) by 2022," it said.

Coal India Limited (CIL) has unveiled a plan to set up a coal-based methanol plant in West Bengal and has invited bids from licensors of coal gasification technology to set up a coal-to-methanol plant.

UNITED STATES

Methanol Institute establishes strategic partnership with SOFI

SOFI, the US Solar Fuels Institute, has signed a memorandum of understanding with the Methanol Institute (MI), to collaborate on areas of mutual interest. SOFI says that it is "excited" about the partnership, which aims to focus on the development of renewable methanol and small-scale production technologies. In a statement, the organisation said that: "the expertise of MI and its members provide significant partnership opportunities for SOFI and our members to enter the methanol and methanol-derived chemicals markets. SOFI's extended global network of experts and researchers, in return, can offer MI and its members promising technologies that are ready for further development."

Founded in 1989 and headquartered in Singapore, with offices in Washington, Brussels and Beijing, the Methanol Institute represents methanol producers, distributors, and technology providers around the globe in a 75 million t/a, \$25 billion

per year industry. SOFI attended the recent Washington Methanol Policy Forum on June 13th to discuss SOFI's efforts in assembling an international public-private network to accelerate the progress towards renewable fuels. However, the organisation stresses that the partnership "does not mean SOFI is betting the house on methanol. Methanol is one of the many pathways to a carbon-neutral solar fuel (potentially a major one in our opinion), and we are motivated to support and accelerate the commercialisation of all solar fuels and carbon utilisation technologies."

CHINA

Clariant opens new office in Yinchuan

Clariant has opened a new office for its Catalysts business in Yinchuan city, capital of the Ningxia Hui autonomous region in China. Clariant says that this makes it the first multinational catalyst provider in northwest China, a region that is home to a majority of the nation's coal-to-chemicals producers as well as numerous syngas and polyolefin companies. The local team in the new office will be supported

by Clariant specialists both within China and from around the world, as part of what Clariant describes as its "China Insider" strategy, which stresses the importance of local expertise for technologies specific to China, combined with full access to global innovations.

The Yinchuan office is Clariant's fifth branch in China. Clariant has two catalyst production facilities in China, one in the Jinshan district of Shanghai, and the other in Panjin city, Liaoning province. The facilities develop and produce a variety of catalysts specifically for China's market requirements and conditions.

NIGERIA

Air Liquide to license refinery hydrogen plant

Air Liquide Engineering & Construction has been selected to supply two hydrogen production steam methane reformer (SMR) units to Dangote Group, the largest manufacturing conglomerate in West Africa. The SMR units will be core to a new hydrogen generation complex producing 200,000 Nm³/h of hydrogen and high quality steam

for Dangote's new refinery located in the Lekki free trade zone in Nigeria. The new refinery is part of the largest industrial complex that is currently under development in Africa and will produce Euro V-compliant low sulphur fuels, among other products. The equipment supply agreement follows a first agreement related to technology license and process design which Air Liquide signed with Dangote in 2015.

Domenico D'Elia, vice president and chairman of Air Liquide Engineering & Construction, said: "We are proud to deliver our technology and the steam reformer packages for the large hydrogen production unit as part of this ambitious refinery project. Dangote's confidence in selecting Air Liquide's hydrogen and steam reforming technology reaffirms our leading position in this market segment, with more than 130 references around the globe "

AZERBAIJAN

SOCAR production still well below capacity

Azerbaijan's state oil company SOCAR has produced more than 140,000 tonnes of methanol from its methanol production facility from the beginning of 2017, according to operating unit SOCAR Methanol LLC. The plant, which has a capacity of 720,000 t/a, began operations in January 2014, but has not produced at capacity due to constraints on demand. The company had previously indicated that it plans to produce 250,000 t/a of methanol in the full year 2017, but that it has a strategic goal to bring production up to 500,000 t/a due to stable and growing demand in global markets. The methanol plant, which was built by AzMeCo LLC at a cost of \$520 million, is one of the biggest investments in Azerbaijan's non-oil sector.

UZBEKISTAN

Amec Foster awarded reformer contract for GTL plant

Amec Foster Wheeler has been awarded a contract by primary EPC contractor Hyundai Engineering to deliver hydrogen plant technology and design and terrace wall steam reformer technologies for Uzbekneftegaz's Oltin Yo'l GTL (gas-to-liquid) plant in the Kashkadarya region of Uzbekistan. Under the contract, Amec will offer materials supply for a terrace wall steam reformer heater at the plant, which is currently under construction. Work to be taken up also comprises the basic design engineering for the 37,100 normal cubic metres per hour (Nm³/h) hydrogen unit. In addition, Amec will provide engineering and procurement for the steam reformer heater, as well as

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cient storage and transportation of hydrogen. Among other benefits, this will greatly facilitate the conservation of surplus energy from wind and solar sources. Clariant has successfully developed dedicated catalysts for LOHC hydrogenation (EleMax® H101) and dehydrogenation (EleMax D101) and the company says that it "continues to support the research with catalyst products and expertise to further increase the efficiency of this technology".

The Power-to-X project is sponsored for an initial period of three years. During this time, the team intends to advance LOHC technology from research and development to industrial maturity.

JAPAN

More hydrogen supply planned for Japan

A consortium of Japanese companies plans to launch a hydrogen supply chain demonstration project by 2020 as part of the country's goal of becoming a "hydrogen society". Chiyoda Corp, Mitsubishi Corp, Mitsui & Co and Nippon Yusen KK said they would take part in the project to build a plant to generate hydrogen in Brunei, where it would be liquefied for onward transport to a hydrogen regasification plant in Kawasaki, near Tokyo. The demonstration project will only ship 210 t/a, the companies say, with the hydrogen being used as feedstock at a gas-fired power plant operated by a unit of oil refiner Showa Shell Sekiyu KK. The project is expected to cost several billion yen, or tens of millions of dollars. Two thirds of the cost will be subsidised by the Japanese government.

SWEDEN

Linde to provide hydrogen plant for refinery

Linde has been awarded the engineering and supply contract for a modular hydrogen plant for the St1 Refinery in Gothenburg. Linde's Engineering Division will start the engineering and construction of the unit immediately, with the completion estimated for the end of 2018.

John van der Velden, member of the board of directors of Linde Engineering Division said: "Linde's ability to deliver unique single-sourcing supplies provides St1 with a cost-effective solution supporting their advanced biofuel production." ■

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People

CF Industries Holdings' board has elected **John W. Eaves**, chief executive officer of Arch Coal Inc., and **Michael J. Toelle**, owner of T&T Farms and former board chairman of CHS, Inc., as independent directors of the company. The elections bring membership of the board of directors to 12. They are expected to stand for re-election by stockholders at the company's 2018 annual meeting.

Eaves, 59, has served as chief executive officer of Arch Coal, a top coal producer for the global steel and power generation industries, since 2012 and has been a member of its board of directors since 2006. He has more than 30 years of experience in the coal industry. During his tenure with Arch Coal, he has held positions of president and chief operating officer; senior vice president of marketing; and vice president of marketing and president of Arch Coal Sales, the company's marketing subsidiary. Mr. Eaves joined Arch Coal in 1987 after serving in various marketing-related positions at Diamond Shamrock Coal Company and Natomas Coal Company. He serves on the boards of the National Association of Manufacturers and the National Mining Association. He is a graduate of the Advanced Management Program at Harvard University, and he holds a bachelor's degree from the University of Kentucky.

Toelle, 55, is the owner of T & T Farms, a diversified farming company, and has been a member of Nationwide Mutual Insurance Company's board of directors since 2013. He is a former board chairman and longtime board member of CHS, Inc., the largest cooperative business in the United States. He also served as a board member for Cenex, Inc., before it merged with Harvest States Cooperatives to create CHS in 1998. Toelle is past chairman of the CHS Foundation and previously served as a director for the Agricultural Council of America and Country Partners Cooperative. He is a member of the National Association of Corporate Directors.

"We are pleased to welcome John and Michael to the CF Industries' Board," said Stephen A. Furbacher, chairman of the board, CF Industries Holdings, Inc. "John's extensive knowledge of the global coal industry and Michael's deep agricultural experience, along with their demonstrated leadership and understanding of commodity cycles, will benefit the Board and our management team greatly. We look forward to their insights and perspectives as we work together to create long-term value for our stockholders."

Nexant, Inc., has appointed **John Gustafson** as the company's new CEO. Mr. Gustafson comes to Nexant from Cognizant Technology Solutions where he served as

vice president and head of the utility and energy business unit. He has held similar leadership roles at other leading service and solution providers including Wipro, IBM, and others. Mr. Gustafson has experience leading all aspects of strategy, operations, delivery, and sales and marketing and, says the company, brings a strong customer-centred focus to creating growth and developing and delivering solutions to meet industry needs. He has worked closely with companies such as Oracle, SAP, Microsoft, IBM, GE, ABB, and other leading solution providers as well as leading advisors and analysts such as ISG, Everest, HFS, Avasant, IDC, Gartner, Ovum, and others.

"We are delighted to welcome John on board as CEO to scale Nexant to the next level. The utility and energy industries are going through transformative changes which present Nexant with an extraordinary market opportunity," said Arjun Gupta, executive chairman. John's experience will be invaluable to bring our new solutions to market."

"I am very excited to join a great company like Nexant with amazing talent and an outstanding reputation. I look forward to working with the Nexant team to build on the foundation that has been created, and scale Nexant to the next level as a market leader in providing new and innovative solutions in the utility and energy industries," Gustafson said. ■

Calendar 2017/18

SEPTEMBER

10-14

62nd AIChE Annual Safety in Ammonia Plants and Related Facilities Symposium, NEW YORK, USA.
Contact: AIChE Customer Service
Tel: +1 800 242 4363/+1 212 591 8100
Fax: +1 212 591 8888
Email: xpress@aiche.org

30

35th World Methanol Conference, BERLIN, Germany
Contact: Lynn Urban, IHS Markit
Tel: +1 303 397 2801
Email: Lynn.urban@ihsmarkit.com

OCTOBER

1-6

Ammonium Nitrate/Nitric Acid Conference, AUSTIN, Texas, USA
Contact: Hans Reuvers, BASF /

Karl Hohenwarter, Borealis
Email: johannes.reuvers@basf.com / karl.hohenwarter@borealisgroup.com

10-12

30th AFA International Fertilizer Technology Conference, AMMAN, Jordan
Contact: Arab Fertilizer Association
Tel: +202 23054467/64
Email: afa@arabfertilizer.org
Web: www.arabfertilizer.org

15-18

Gasification and Syngas Technologies Meeting, COLORADO SPRINGS, USA
Contact: Gasification and Syngas Technologies Council, 3030 Clarendon Blvd. Suite 330 Arlington, VA 22201 USA.
Tel: +1 703 276 0110
Fax: +1 703 276 0141
Email: info@gasification-syngas.org
Web: www.gasification-syngas.org

24-26

IFA Crossroads Asia-Pacific Conference, SHANGHAI, China
Contact: IFA Conference Service, 28 rue Marbeuf, 75008 Paris, France.
Tel: +33 1 53 93 05 00
Email: ifa@fertilizer.org

NOVEMBER

25-26

Iranian Nitrogen+Syngas Conference, TEHRAN, Iran
Contact: CRU Events
Tel: +44 (0) 20 7903 2444
Fax: +44 (0) 20 7903 2172
Email: conferences@crugroup.com

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26-March 1

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Plant Manager+

Problem No. 44 Rise in HP CO₂ stripper outlet temperature

In the high-pressure CO₂ stripper of a urea plant, which is essentially a shell-and-tube heat exchanger, the incoming carbon dioxide feed flows counter-current to the urea solution leaving the reactor.

On the shell side, the high-pressure stripper is heated with steam. The off-gas of the high-pressure stripper, containing

the feed carbon dioxide along with additional carbon dioxide and ammonia from the dissociated carbamate, is then fed into the carbamate condenser or pool reactor. This round table discussion is about the high-pressure CO₂ stripper and how to solve the problem of a high temperature in the bottom of the stripper.

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India initiates this round table discussion:

In a CO₂ stripping process, what could the reason be for an increase of the stripper liquid outlet temperature from 176°C to 182°C?

The stripper off gas temperature has reduced from 186°C to 183°C. The stripper vent on the shell side is more open than normal so that inerts can be vented and heat transfer will be more efficient. There is no fluctuation in the stripper level or steam drum level.

Majid Mohammadian from OCI Company, The Netherlands replies and asks for some further information:

There are several reasons for high temperature in the stripper bottom: high level at the stripper bottom, poor CO₂ distribution, poor liquid distribution, too high steam pressure, poor reactor composition etc.

In your case when the delta temperature between the liquid inlet and liquid outlet is low, about 1°C, it indicates poor liquid distribution in the top of the stripper, resulting in poor stripping. If you check the lab results you will find poor stripper efficiency as well. Please provide the lab results of the stripper inlet and outlet.

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India provides more information:

When the stripper outlet temperature was 182°C, the reactor outlet composition was 30.5 wt-% ammonia and 24.9 wt-% CO₂ and the stripper outlet composition was 10.1 wt-% ammonia and 11.2 wt-% CO₂. Urea is 33.3 wt-% at the inlet of the stripper and 54.1 wt-% at the stripper outlet.

The level in the stripper bottom is measured by RADAR indicator and is controlled by a level control valve. The level control is found to be accurate. The steam drum pressure is maintained at the normal value of 18 kg/cm²g at a low load.

Majid Mohammadian from OCI Company, The Netherlands returns to the discussion:

The CO₂ percentage in the reactor outlet (24.9 wt-%) looks wrong, it is normally between 17 and 19 wt-%.

The stripper as per the lab data is working with 75% efficiency (design efficiency is 80%).

Poor liquid distribution at the top of stripper is the reason for high temperature at the bottom of the stripper (less delta T around the stripper) and consequently poor efficiency.

My advice is to stop the plant, open the stripper and check the liquid dividers, holding plates, Teflon rings and fix the problem.

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India asks further questions:

The stripper outlet temperature is found to be dropping day by day (it has dropped to 178°C which is almost back to the normal value) as the steam pressure of the stripper increases above 20 kg/cm². What is the reason for the normal bottom temperature after increasing the stripper pressure? Has anyone faced the same kind of problem? What actually happened inside the stripper?

Maksud Alam from KAFCO Company, Bangladesh asks for more information:

Could you please tell us about the following points:

1. Throughout the process of increasing steam pressure was the plant load the same?
2. Did you notice any change in the low-pressure section and process condensate treatment load before and after?
3. Do you have a condensate level control on the steam side (not the saturator level but the stripper shell side level).

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India replies:

1. At a higher plant load the steam pressure used to be 19.20 kg/cm² but now for the same plant load the steam pressure is 20.30 kg/cm². If the steam pressure is dropped then the low-pressure system load increases.
2. Yes, the low-pressure system is increasing and more gas is venting from the low-pressure system to the scrubbing system.
3. There is no condensate level control on the stripper shell side. Level control is there only for the saturator.

Mark Brouwer from UreaKnowHow.com, The Netherlands provides more information:

I am also thinking that there may be issues with the liquid distribution:

1. Have you ever experienced fouling in the top of the HP stripper?
2. What is the age/condition of the liquid dividers?
3. Did these problems occur shortly after a turnaround?

Sandesh Patil from Mangalore Chemicals and Fertilizers Company replies to Mark's questions:

1. We have never experienced fouling in the new stripper.
2. The top ferrules were replaced last year and the stripper was replaced in 2006.

3. This problem occurred after starting the plant after a short shutdown of 6 days. The turnaround took place in May 2016 and after starting the plant in June 2016, the plant has been running smoothly for the last four months.

Maksud Alam from KAFCO, Bangladesh replies with more information and some words of caution:

As described before, poor liquid distribution leading to poor stripping efficiency may be the root cause of your problem, which you have overcome by increasing the steam pressure. But please bear in mind that some tubes may have a thicker layer of film and some may be thinner/dry. In either case the higher steam temperature will contribute to higher corrosion rates.

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India replies and raises new questions:

At present the stripper shell side pressure is maintained higher compared to the normal pressure at the full load. If there is a problem in the liquid distributor, how many days can the plant be kept in operation at this higher stripper pressure? Tubes whose liquid distributor is not working properly may be overheating and there will be higher corrosion in those tubes. (Stripper outlet analysis is carried out weekly for Ni, Cr and Fe and it is within the design limit). Is there any other way we can identify the higher corrosion rate in those tubes where the liquid distributor may not be working properly so that the plant can be shut down before the stripper tubes reach final breakdown?

Mark Brouwer from UreaKnowHow.com, The Netherlands asks a new question regarding the material of construction:

What is the material of the stripper tubes?

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India replies:

The stripper tubes are made of Safurex.

Bob Edmonson from a contracting company in Canada gives his point of view:

I agree with the others who have concerns regarding the liquid distribution. Your plant is quite new so it's hard to imagine why, but if one or more bolts that secure the hold down plates are loose, due to corrosion, a plate can lift and the liquid divider tubes can rise up somewhat. If that happens, you will see stains on both sides of the Teflon rings when you are in the stripper next time. The risk is that these tubes can (will) corrode more rapidly if 25-22-2 so it might be worth checking the ppb nickel in the stripper outlet liquor if this happens again.

Ali Ancaza from Igsas, Turkey raises new questions:

- How long has the stripper steam pressure been like that?
- How is the level in the stripper steam drum?
- How long has the stripper steam temperature been like that?

This series of discussions is compiled from a selection of round table topics discussed on the UreaKnowHow.com website. UreaKnowHow.com promotes the exchange of technical information to improve the performance and safety of urea plants. A wide range of round table discussions take place in the field of process design, operations, mechanical issues, maintenance, inspection, safety, environmental concerns, and product quality for urea, ammonia, nitric acid and other fertilizers.

Mark Brouwer from UreaKnowHow.com, The Netherlands shares his experience:

Although with Safurex tubes there is no risk of active corrosion, it makes sense to check the liquid distributor system as soon as possible.

Ali Ancaza from Igsas, Turkey shares experience from their plant:

We have experienced the same problem in our system in 2002. In order to help you, I need to know the following values:

1. What is the CO₂ input temperature to stripper?
2. What is the output temperature of stripper urea?
3. What is stripper vapour pressure and steam temperature?
4. What is the steam drum level of the stripper?
5. What is the low pressure recirculation pressure?

Salam Malih from North Fertilizer Company, Iraq offer his advice:

1. Check the CO₂ content in the stripper outlet and check the performance of the level measurement of the stripper as there is the possibility of CO₂ slippage.
2. Check the ratio of O₂ in CO₂.
3. Check the HP steam consumption of the stripper.
4. Check the synthesis pressure.

Sandesh Patil from Mangalore Chemicals and Fertilizers Company, India answers Ali's questions:

1. CO₂ input temperature is maintained at 135°C.
2. Stripper urea outlet temperature is 179°C.
3. Steam temperature of shell side of the stripper is 215°C and the steam drum pressure is maintained at 20.30 kg/cm²g (which vapour pressure do you need?)
4. Level in the steam drum is maintained at 45%.
5. Which recirculation pressure do you need – the carbamate recirculation?

... and responds to Salam's questions as follows:

1. At the stripper outlet the CO₂ is 10.82 wt-% which is close to the design value of 10.50 wt-%.
2. Oxygen is maintained at 0.55 vol-%. Oxygen analysis at the outlet of the H₂ converter is carried out daily.
3. HP steam consumption for stripper for full load is normal (almost 44-46 t/h)
4. Synthesis pressure is maintained at 144.5 kg/cm²g.

Ali Ancaza from Igsas company, Turkey provides some suggestions to reduce the stripper liquid outlet temperature:

Gradually increase the stripper steam level of the drum up to 55-60%. According to the recirculation pressure, increase the pressure of the drum by 21.5-22 kg/cm². The outlet temperature of the stripper solution will decrease.

Kashif Naseem from SABIC, Saudi Arabia shares his experience:

Check your stripper liquid distribution system which is the main cause for it.

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Trans-Pacific methanol: a viable solution?



Chinese-backed Northwest Innovations Works (NWIW) is trying to build two huge methanol complexes on the Columbia River in the US Pacific northwest to make methanol from cheap US shale gas and ship it to China. Is this scheme practical or over-ambitious in a lower oil price environment?

As we noted in our article in the previous issue (*Nitrogen+Syngas* 348, Jul/Aug 2017, pp20-23) Chinese demand for methanol has come to dominate the market for that product, with more than 50% of all global use, initially via fuel uses such as direct methanol blending into gasoline and production of dimethyl ether for propylene blending. However, the current surge of demand is being led by petrochemical derivatives, specifically methanol to olefins (MTO) or methanol to propylene (MTP) production. This has been a boon for China because it has allowed it to use domestic coal for chemical and petrochemical production, rather than having to import even more expensive oil from the Middle East. The huge push to develop new MTO capacity means that there is now 25 million t/a of capacity (in terms of methanol demand) as of 2016, with as much again planned or under construction.

One interesting development has been the amount of non-integrated MTO capacity that has been or is being built. This relies on 'merchant' methanol – bought on the open market. Much of this is produced domestically within China – China produced 43 million t/a of methanol in 2016, according to the National Statistics Bureau, and production continues to grow at around 6% year on year (7% in 2016), with a further 5.4 million t/a of new or re-started capacity due to come into the market during 2017. Even so, China's rapidly growing demand means that the gap between domestic methanol production and consumption continues to grow. Although the country has a large amount of 'unused' methanol capacity, much of this is either small scale and too expensive to operate or otherwise not likely to be able to compete with imported methanol based on cheap natural gas. Total methanol

imports in 2016 were around 8.8 million t/a, up nearly 60% on 2015. The ability of the international market to supply all of the methanol required is by no means a given. Chinese olefins demand was 40 million t/a in 2015, and this is expected to double to 80 million t/a by 2023, with MTO representing 40% of that demand, or 20 million t/a, equivalent to over 40 million t/a of methanol required.

The answers to this conundrum have been twofold. One has been to try and develop integrated methanol/MTO capacity elsewhere. In April 2017, Mingyuan Holdings Group Company announced plans to set up a major MTO plant alongside a giant greenfield methanol scheme at the China-Oman Industrial Park, in the Sultanate. A capacity of 1.8 million t/a of methanol is being proposed in the first phase. However, another has been to try to set up dedicated methanol supply elsewhere and ship this to China to be converted into olefins there. This has been the rationale for a major series of projects which have been proposed by Northwest Innovation Works, to develop large-scale methanol plants in Washington State on the west coast of the United States, to allow cheap US shale gas to feed 10,000 t/d of methanol production for export to China for olefins production.

Northwest Innovation Works

Northwest Innovation Works (NWIW) is the company carrying out the methanol projects in the Pacific Northwest. NWIW is a subsidiary of Pan-Pacific Energy Corp, which is itself owned by Shanghai Bi Ke Clean Energy Technology Company Ltd. (also known as Clean Energy Commercialisation Company, CECC). Shanghai-based CECC, founded in 2008, is majority owned by the Chinese Academy of Science Holdings, together with Double Green Bridge Hong Kong (a investment consortium including senior members of CECC management) and Johnson Matthey plc. CECC says that its "primary focus is methanol production and downstream conversion businesses," aiming to pioneer "new technical and commercial innovations in both China and North America to create a platform for delivering methanol into the industrial processes where it is most needed." This will be achieved by using abundant US natural gas resources to produce methanol using low emission technology and shipping it to China to be used in methanol to olefins and other downstream production.

The projects

The three original project proposals were for Tacoma and Kalama in Washington State and the port of St Helens at Port Westward in Oregon, with 20,000 t/d of methanol capacity at Tacoma and 10,000 t/d each at the Kalama and St Helens projects – these two sites are both on the Columbia River, north of Portland Oregon, but on opposite banks (the river forms the state boundary), about 10km from each other, and could be thought of as the Tacoma project split over two sites. Both 10,000 t/d sites will be comprised of two 5,000 t/d (1.7 million t/a) methanol trains. In April 2016, NWIW said that it was putting the \$3.6 billion Tacoma project on hold for the time being, after strong local opposition and a lukewarm response from local government officials. However, the two \$1.8 billion projects on the Columbia River are still both proceeding. Land was leased to NWIW at both sites in 2014, and the permitting process began.

So far the Kalama project has been the one which the company has pursued most vigorously. As with Tacoma, there has been a rocky road as regards local opposition, but following a revision to the proposals to make the project more environmentally friendly, key milestones have been achieved; in June

2017 an air quality permit and local shorelines use permit were both granted by local authorities, and site clearing has been completed. Natural gas supply for the project has been agreed, and will come from a new spur pipeline which is being constructed to bring gas to the site. The Kalama Lateral Project is a 3.1 mile, 24" diameter pipe connecting the Northwest Pipeline to the port of Kalama, and was authorised in 2016 by the Federal Energy Regulatory Commission (FERC). Williams, which owns the Northwest Pipeline, says that it is targeting 2019 for the pipeline spur construction, assuming that the NWIW plant goes ahead.

Innovative technology

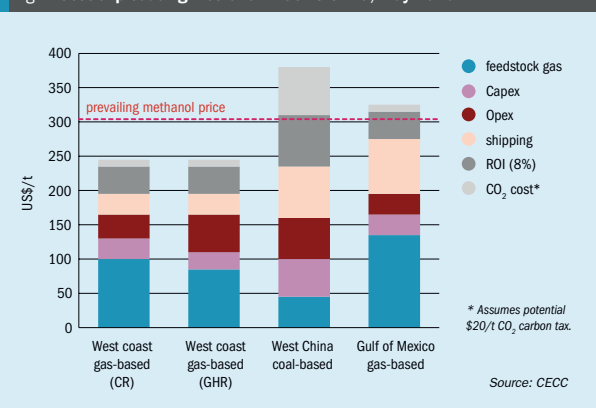
In a bid to pass the stringent scrutiny of US domestic environmental groups, NWIW has had to look towards innovative ways of producing the methanol to achieve the lowest possible emissions. CECC estimates that the overall greenhouse gas equivalent of producing olefins from coal-based methanol in China is 3.84 tonnes CO₂ per tonne of olefins, whereas producing it in the US from natural gas and shipping it to China to make olefins represents only 0.74 tonnes CO₂ per tonne, with the shipping representing only 3% of this.

However, via their technical partners Johnson Matthey, they have examined ways of further reducing the emissions from the plant inside battery limits (ISBL) in order to meet exacting local targets for GHG emissions. As discussed in our article on low carbon syngas last year (*Nitrogen+Syngas* 339, Jan/Feb 2016, pp20-25), this can be done

by maximising the use of electrically operated pumps, motor-driven (as opposed to steam turbine driven) syngas compressors, and electrically-powered cooling. This so-called ultra-low emissions (ULE) technology can reduce the greenhouse gas generation ISBL from around 32.9 g CO₂/MJ methanol for a conventional gas heated reformer plus autothermal reformer combination to around 15 g CO₂/MJ for the maximum electrical import option – a reduction of almost 60%. Of course, this merely displaces the energy generation requirement from the plant to the local electricity generator, and if electricity is generated by renewable means then this is also the way of generating the minimum CO₂ emissions over the whole system. As it is, the approximately 200MW of electricity that will be required by the facility will in fact come approximately 50% from an on-site gas-based electrical generation facility and 50% from local supply. Nevertheless, this helps the state of Washington maintain its overall greenhouse gas emissions limit, and helped the plant to secure its environmental permit.

Waste water is also a concern for the facility. The methanol plant will use 3,000 gallons per minute of water, about 90% of that for cooling and 8% as steam in the reforming process. Most of the water will evaporate to atmosphere at the cooling towers, which will also discharge non-contact cooling water to a pond where it will be cooled with incoming raw water, treated, and ultimately sent to an existing outfall. However, NWIW says that it wants to maintain 'zero liquid discharge' from the plant to recycle and re-use waste water within the facility.

Fig 1: Cost of producing methanol in US vs China, May 2015



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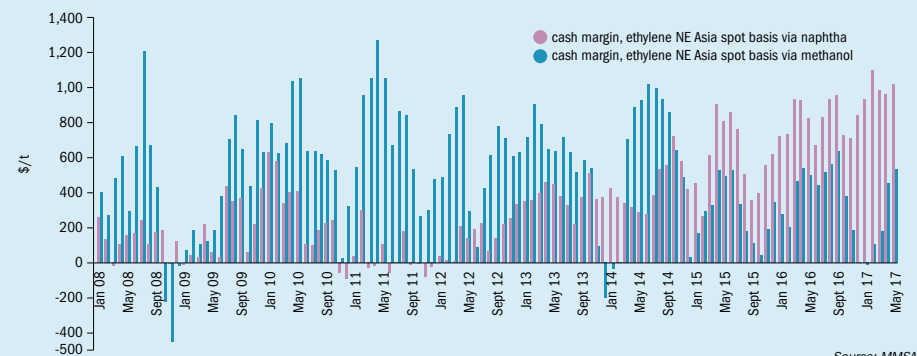
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Fig 2: Cash margins on ethylene, MTO vs naphtha



Source: MMSA

CMTX

In addition to the methanol production technology, CECC also have their own proprietary MTO conversion technology which they plant to employ at the Chinese end of the project. This technology, called CMTX, converts methanol to propylene and butylene (mainly) and has been developed in a collaborative research partnership between CECC, the Shaanxi Coal and Chemical Industry Group Research Institute, and Shanghai Hoto Engineering Inc.

The process is described in full in *Nitrogen+Syngas* 344, Nov/Dec 2016, pp50-52, but briefly, methanol is injected into a fluidised bed reactor after vaporisation and heating, where it is catalytically converted to a high carbon olefin-rich product gas. The reaction heat is recovered while the product gas is chilled and distilled to form the final product mix of propylene, butylene, and mixed aromatics. The catalyst is regenerated at 360-400°C in a regeneration column. CECC says that single pass conversion of methanol is 99.8%, with propylene single pass selectivity of >38%, and butylene single pass selectivity of >24%. The company says that the catalyst has a good resistance to carbon deposition and a low coking rate (0.06%), while the reactor has a high space-velocity, low water to methanol ratio, and relatively small size.

Economics

So, permitting aside, the key question for the NWIW project is: do the economics stack up? Figure 1 shows the cost of pro-

ducing methanol and shipping it to a methanol to olefins producer on the Chinese coast for four potential routes – from a natural gas based plant on the US west coast (with both a combined reforming – CR and gas-heated reforming – GHR – production route), from a plant in the Gulf of Mexico, and from a coal-based plant in the coal-rich west of China. A theoretical carbon tax is also shown to illustrate the environmental advantage of gas-based production over coal-based production. As can be seen, the US west coast option appears to have a distinct competitive advantage over production elsewhere in the US and domestic production within China, even when the cost of shipping large volumes of methanol across the Pacific Ocean are taken into account.

However, the downstream MTO plant must not only reckon on the relative cost of importing methanol from different sources, it must also compete with other routes to produce olefins, and the key to project economics is the relative price of producing olefins from methanol as compared to via more conventional routes such as from naphtha cracking. Figure 2 shows the margins for production of ethylene from naphtha as compared with MTO for a hypothetical production site in NW Asia over the past decade. The effect of lower oil prices over the past couple of years can be clearly seen, with the red bars for naphtha production far more positive towards the right hand side of the graph. However, the cash margin for MTO production is generally higher than for naphtha, and even though this has not been the case since early 2015, MTO nevertheless retains a positive cash margin.

Furthermore, CNCC would argue that here the CMTX route to olefins has some advantages over other potential MTO routes. The margins for ethylene production are often not as good as the margins for propylene production – the rise of ethane-based ethylene production in the Middle East over naphtha and increased demand for polypropylene has meant that there is a shortage of propylene from naphtha crackers and hence a growing amount of on-purpose propylene production. Butylene is likewise in increased demand as a polyethylene co-monomer, and China is the largest and fastest growing area of demand for this. As the CMTX process is skewed towards propylene and butylene production, it is argued that it thus has a further advantage over an MTO process that focuses on ethylene and propylene.

Summing up

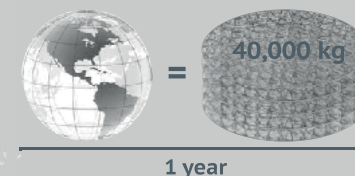
There has been some scepticism about the feasibility of the NWIW plants, both in terms of the difficulty of getting permits for greenfield chemical sites in the US, and on the grounds of the economics of production, especially in a lower oil price environment. However, with a dedicated off-take MTO facility which will also be licensed by the owner and developer of the methanol plant, there appears to be some ground for optimism on the economics front, and the move to a low GHG production technology inside battery limits and total water recycle seems to have answered the demands of US environmentalists. At the moment, there seems no reason to believe that this ambitious project will not be realised. ■

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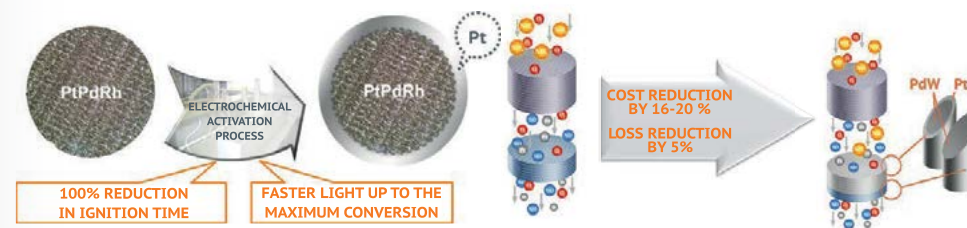
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The 14th International Methanol Technology Operators' Forum (IMTOF) was held in London in July, this year coinciding with Johnson Matthey's celebration of its 200th anniversary.

Johnson Matthey (JM) hosted the 14th International Methanol Technology Operators' Forum (IMTOF) at the Hilton Paddington in London from July 9th-12th. With JM now holding both the Davy Process Technology and the old ICI Methanol portfolios, the meeting remains one of the few held globally which are dedicated to methanol technology.

Methanol has certainly come of age over the past decade or two. JM methanol market director Peter Roberts, welcoming delegates, highlighted methanol's spectacular demand growth over the past few years, achieving rates of 6-7% year on year, as new derivatives like methanol to olefins (MTO) and methanol to gasoline (MTG) make themselves felt, as well as the large opportunity presented by methanol's fuel uses, most recently in the marine sphere (the subject of our lead article in *Nitrogen+Syngas* 347, May/June 2017, pp18-20). And although China's bottleneck industrial growth has slowed, Peter said, JM has licensed three more new methanol plants there this year with a combined capacity of 6.2 million t/a, the most recent Shenhua Yulin, for which the contract was signed on June 13th, with a capacity of 6,120 t/d of coal-based methanol, making it the largest coal-based plant in the world. Elsewhere, the lifting of sanctions on Iran

has provided further options for gas monetisation there, possibly including downstream derivatives, although financing remains the key issue, especially as regards continuing US sanctions. North America faces construction risk and environmental hurdles, and although there is interest in renewable or biomethanol, smaller capacities tend to need subsidies to work economically. However, opportunities for using the JM gas heated reformer while importing renewable power make for a potential low carbon and water use option. There are also Chinese-backed projects in the FSU and potential for development in North Africa, where in spite of political issues in Algeria, Libya and Egypt there are positive signs for gas developments. However, methanol prices continue to be held back by low oil prices. OPEC has been trying to tighten supply, but US production continues to increase.

Turning from the present to the future, Mark Sutton, JM's technical director, presented JM's vision of the way ahead. Innovation has been at the heart of the company's longevity, he said, with 12% of the company's workforce engaged in research and development. But now, factors such as population growth, urbanisation and increasing wealth, health and nutrition, resource constraints, and environmental factors such as climate

change and emissions, particularly carbon dioxide, are all driving the move towards industries having to become more sustainable, he said. The company's aim is therefore to apply cutting edge science and technology to provide customised solutions and next generation products, with the aim of becoming a trusted and admired speciality chemical company, via partnerships with scientific centres of excellence to enhance innovation such as Diamond Light Source at Harwell (the UK's former Atomic Energy Authority laboratory) with its state of the art materials characterisation facility, allowing the observation of catalyst activity at a molecular level.

Methanol market

Mark Berggren of Methanol Market Services Asia (MMSA) gave a round-up of the global methanol market. Methanol's versatility has driven its growth, with MTO becoming increasingly important to the market. Methanol demand in 2016 was 85 million t/a, but by 2022 this will have risen to 122 million t/a, with MTO representing one third of this, possibly even overtaking formaldehyde as the major methanol derivative. Prices have fallen from the peaks of the first quarter of 2017, he said. Chinese demand is estimated to rise 3% this year over 2016, taking it to 55 million t/a, but China's demand growth comes in large, uneven steps. Nevertheless, MTO is still alive and capable of being a major consumer even at modest price levels. Supply growth has begun to slow, and there is little Chinese merchant methanol production in the pipeline, outside of two major projects in the US which will impact over the next five years. New Chinese capacity is plateauing and demand still rising rapidly, implying operating rates rising (are the technical limits to these), and China will also need to grow its imports of methanol to sustain coastal MTO demand. Iranian projects are competing for gas supply, and there is little certain investment outside these over the next five years. US demand stands at 6.9 million t/a. There has been no net increase in capacity this year so far, but more is expected at the end of the year. Uncertainty persists over US projects, and funding can be difficult without market access, but capacity is nevertheless seemingly rising towards 10 million t/a by 2022, with net exports from 2020.

Methanol pricing continues to be set – on the floor side, by Chinese coal-based capacity, at around \$200-250/t, and on the ceiling by the price that non-integrated Chinese MTO producers are willing to pay,

around \$400/t. Asian methanol prices tend to be below Atlantic basin rates, as 'swing' export capacity is mainly in the Middle East. Low oil prices advantage the naphtha route to olefins, and Chinese MTO capacity is currently not as competitive, but cash margins remain positive and MTO operating rates are still above 70%. Likewise methanol continues to be priced below the gasoline equivalent blend value, and this is consuming 5-10 million t/a.

Looking far ahead into the future, out towards 2040, Mark saw growth rates for methanol demand falling towards 2.6% per year, but this still implied a methanol market of 195 million t/a by that time, another 103 million t/a of new capacity (assuming an 85% operating rate), or an extra 49 world scale methanol plants – more than 2 per year, with only 19 million t/a of likely new methanol capacity currently on the books.

Safety

Greg Dolan of the Methanol Institute noted that while China still does not have a national M15 standard for methanol blending in fuels, the national focus is now moving towards M100 – complete replacement of gasoline with methanol as a fuel, with 100,000 vehicles now capable of running on methanol, representing 5,000 bbl/d of demand. He also highlighted MI's work on product stewardship, with safe handling of methanol by those potentially unfamiliar with it key to its continued expansion. To this end, MI has developed a new edition of its methanol safe handling manual which includes a new 220 page section on fire safety and risk-based process safety management, a crisis communications guidebook, a methanol safe handling video, technical bulletins, and a safety newsletter.

Chris Loveridge of ABB considered best practise process safety and emissions technology for fired equipment. Reformers are expensive and have long equipment lifetimes, but emissions standards and instrumented protective systems do change over time, and may not have moved on sufficiently in terms of adequate integrity of fuel gas isolation, lack of flame failure detection or control over air/fuel ratio, burner ignition systems, safety function logic and ignition source control. Additionally, emission regulations have been tightened in Europe, and for reformers the problem with more stringent emissions expectations is NOx. Optimising the burner set-up and combustion air will often reduce NOx and is the starting point

for further improvements. Very low NOx burners are not suitable for reformers, but other options include selective catalytic reduction (which can be expensive as a retrofit) and selective non-catalytic reduction, which can often achieve the required target at lower cost, as well as combined techniques.

In other safety-related papers, Cristiano Azavedo and Jalila al-Zadjali of the Oman Methanol Company described their company's approach to process safety management, including lessons learned from a handful of minor accidents at the site; while Michael Eltink of Protomation explained the advantages of using a high fidelity operator training simulator to take operators through a variety of scenarios and enhance both safety and plant reliability. Finally, JM's Paul Cassidy looked at safety issues resulting from a pipe failure at a coal-based plant in China, and attempts to diagnose and rectify the root cause.

Sustainability

On the sustainability front, Dr Sean Axon described JM's sustainability strategy. In 2007 the company set out a 10 year strategy covering two key areas; using fewer resources, and making more efficient and durable products for customers. To the so-called 'triple bottom line' of social, environmental and financial sustainability, the company has also added governance and health and safety. With 10 years now elapsed, he reported on progress with the six targets which were set in 2007. The company aimed to halve the carbon intensity of its activities – something which has been achieved. It also aimed to double earnings per share, also delivered. On the target of halving key resources consumed per unit of output, this is 96% achieved, with water, gas and electricity all beyond target. The company's aim of sending zero waste to landfill is 58% achieved, and its targets of zero occupational injuries/illness has been 76% achieved. A new strategy, setting targets for the next eight years to 2025, is now under development.

Other sustainability papers tackled boosting methanol production using carbon capture from reformer flue gas using Fluor's *Econamine FG Plus* technology.

Operating experiences

As usual several plant operators presented their own methanol plant operating experiences. Jan Coetzee of Methanex New

Zealand described an incident with hot banding and carbon deposition in reformer tubes at the Motonui methanol plant, which turned out to be caused by dimethyl sulphide slip from the desulphurisation unit. Jamal Ali Shawoosh of GPIC in Bahrain described how a cathodic protection system enhancement has helped his company's management of corrosion in methanol storage tanks.

China Coal Shaanxi Yulin Energy gave an account of operating their mega methanol plant in Yulin City, China. The plant has a nameplate capacity of 1.8 million t/a, although actual production can reach 1.95 million t/a. Downstream there is 600,000 t/a of polyolefin production, although again this can in practise run higher; in 2015 it produced 683,000 tonnes, in 2016 709,000 t/a, and this year it is expected to produce 690,000 t/a, equivalent to around 2 million t/a of methanol consumption. Commissioned in 2014, it began operations in early 2015.

Methanol technology

Two new technologies formed the 'meat' of the methanol technology section. Charles Yiu of JM presented the new Davy Advanced Series Methanol Synthesis Loop (ASL). This uses an axial steam raising converter on one side in a higher carbon loop and a radial steam raising converter on the other side in a higher hydrogen loop – an attempt to benefit from the high heat transfer of one and the low pressure drop of the other. This gives better feed and energy efficiency for the same catalyst volume and allows the process to operate at lower pressure than other competing low recycle loops, for greater cost savings per tonne of methanol.

JM's Simon Early next described a new design for a radial steam raising converter. The original design used around 100,000 2mm holes drilled in a 25mm steam plate which was then rolled into a cylinder and welded. This had drawbacks to do with fabrication, potential uneven gas flow from the weld seam, and the small holes could be plugged by debris and difficult to clear. The new distributor design uses C-channels with fewer, larger holes which can be punched rather than drilled, saving time and effort and avoiding the other issues.

Elsewhere, Tess Cakebread-Brown ran through the various options for enhancing methanol plant performance throughout the life cycle of the plant, from major equipment item-led revamps to optimised catalyst loadings. ■

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Syngas project listing 2017

A round-up of current and proposed projects involving non-nitrogen synthesis gas derivatives, including methanol, hydrogen, synthetic/substitute natural gas (SNG) and gas- and coal to liquids (GTL/CTL) plants.

Contractor	Licensor	Company	Location	Product	mt/d	Status	Start-up date
CANADA							
n.a.	Sasol	Sasol	Strathcona, AB	GTL	7,400	P	n.a.
CHINA							
n.a.	JM (DAVY™)	Shenhua Yulin Energy	Yulin, Shaanxi	Methanol	6,120	CA	2020
n.a.	JM (DAVY™)	Shaanxi Yanchang	Shaanxi	Methanol	5,760	CA	2020
n.a.	JM (DAVY™)	Suxin Energy Hefeng	Tacheng, Xinjiang	SNG	2 bcm/a	BE	2019
n.a.	JM (DAVY™)	Xinwen Mining	Yili, Xinjiang	SNG	2 bcm/a	C	2017
n.a.	JM (DAVY™)	Jinmei Tianqing	Qinyang, Henan	SNG	0.3 bcm/a	C	2016
n.a.	JM (DAVY™)	Lianhe Energy	Baotou, Mongolia	SNG	0.5 bcm/a	BE	2019
n.a.	JM (DAVY™)	Shaanxi Yanchang	Shaanxi	Methanol	5,500	UC	2018
n.a.	JM (DAVY™)	China Coal Shaanxi Yulin	Shaanxi	Methanol	5,500	CA	2019
n.a.	JM (DAVY™)	Qinghai Provincial Mining	Qinghai	Methanol	5,800	BE	2019
Casale	Casale	Shandong Jinmei	Zhangqiu, Shandong	Methanol	2,000	C	2017
Casale	Casale	Xinneng Energy Ltd	Erdos, Mongolia	Methanol	2,000	UC	2017
Casale	Casale	Precious Tyrone Qitaihe	Qitaihe, Heilongjiang	Methanol	2,000	UC	2017
n.a.	n.a.	Gansu Pingliang	Pingliang, Gansu	Methanol/MTO	5,500	UC	2017
INDIA							
Engineers India Ltd	Haldor Topsoe	Assam Petrochemicals	Namrup	Methanol	525	UC	2019
n.a.	Haldor Topsoe	Bharat Petroleum Co	Mumbai	Hydrogen	148	RE	2017
n.a.	Haldor Topsoe	HPCL Mittal Energi	Bhatinda	Hydrogen	2 x 131	RE	2019
n.a.	JM (DAVY™)	Reliance Industries	Jamnagar	SNG	1.4 bcm/a	UC	2017
INDONESIA							
n.a.	n.a.	Petronas	Bintulu, Sarawak	Methanol	n.a.	FS	n.a.
IRAN							
Namvaran	Haldor Topsoe	Marjan Petrochemical	Marjan	Methanol	5,000	UC	2018
TCC	Haldor Topsoe	MEKPCo	Pars	Methanol	5,000	UC	2018
Namvaran	Haldor Topsoe	Badre-Shargh Pet Co	Chabahar	Methanol	5,000	BE	2019
PIDEC	Casale	Kaveh Methanol	Bander Dayyer	Methanol	7,000	UC	2018
PIDEC	Casale	Apadana Methanol	Assaluyeh	Methanol	5,000	UC	2018
n.a.	Casale	Bushehr Pet Co	Assaluyeh	Methanol	5,000	UC	2017
KAZAKHSTAN							
Fluor	Compact GTL	Kazoil	Aktobe	GTL	350	DE	Delayed
NETHERLANDS							
n.a.	Air Liquide	Akzo/Enerkem/Port of Rotterdam	Rotterdam	Methanol	n.a.	P	n.a.
NIGERIA							
n.a.	Haldor Topsoe	Brass Fert & Petchem	Brass Island	Methanol	5,000	DE	2020
PAPUA NEW GUINEA							
LG Industries	n.a.	Sojitz/NPCP	West Papua	Methanol	3,000	CA	2021
PHILIPPINES							
n.a.	Haldor Topsoe	Petron	Bataan	Hydrogen	2 x 69	CA	2018
RUSSIA							
n.a.	Haldor Topsoe	Shchekinoazot	Shchekino	Methanol	1,350	UC	2019
Hyundai	Toyo Engineering	National Chemical	Nakhodka	Methanol	3,000	CA	2021

Contractor	Licensor	Company	Location	Product	mt/d	Status	Start-up date
TRINIDAD AND TOBAGO							
MHI	MGC	Caribbean Gas Chemical	La Brea	Methanol	3,000	UC	2019
MHI	MGC	Caribbean Gas Chemical	La Brea	DME	300	UC	2019
TURKMENISTAN							
Hyundai/LG/Tochu	Haldor Topsoe	Turkmen GTL	Ashgabat	GTL	2,200	UC	2018
Hyundai	Haldor Topsoe	Turkmenbashi Refinery	Turkmenbashi	Hydrogen	100	BE	2019
UNITED STATES							
Fluor	JM (DAVY™)	South Louisiana Methanol	St James Parish, LA	Methanol	5,000	DE	2021
OCI	Air Liquide	Natgasoline LLC	Beaumont, TX	Methanol	5,000	UC	2017
Proman	JM (DAVY™)	Big Lake Fuels	Lake Charles, LA	Methanol/MTG	4,200	DE	2020
n.a.	JM (DAVY™)	Northwest Innovation	Clatskanie, OR	Methanol	2 x 5,000	P	2020
n.a.	JM (DAVY™)	Northwest Innovation	Kalama, WA	Methanol	2 x 5,000	DE	2019
n.a.	Primus Green Energy	Primus Green Energy	Proctor, WV	Methanol	160	UC	2018
n.a.	Technip	Air Products	Baytown, TX	Hydrogen	2,500	DE	2018
Fluor	n.a.	Lake Charles Methanol	Lake Charles, LA	Methanol	3,000	P	2020
n.a.	Relocated plant	US Methanol	Charleston, WV	Methanol	480	BE	2018
UZBEKISTAN							
Hyundai	Haldor Topsoe/Sasol	Oltin Yo'l GTL	Shurtan	GTL	5,000	UC	2020
KEY							
BE: Basic engineering		DE: Design engineering		P: Planned/proposed		Conversion:	
C: Completed/commissioning		FS: Feasibility study		RE: Revamp		1 t/d of hydrogen = 464 Nm³/h	
CA: Contract awarded		n.a.: Information not available		UC: Under construction		1 t/d of natural gas = 1,400 Nm³/d	



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Environmental improvements in UAN production

Elisa Puci and Joey Dobrée of Stamicarbon describe ways of integrating emission treatment technologies in urea and UAN plants.

The liquid fertilizer UAN has been around for decades in the North American market. The product is mainly popular in areas where agriculture is intensive and highly mechanised, but the suitability of UAN for application via irrigation systems also facilitates the expansion of its market towards areas in need of efficient water management in e.g. Latin America and Asia.

UAN is typically produced by blending liquid urea and ammonium nitrate, the latter coming from the neutralisation of nitric acid with ammonia. The most sophisticated way of producing UAN is via integration of urea synthesis with neutralisation and blending to UAN. This design optimises the capital investment and limits the amount of equipment typically required in a urea plant.

Over the past few years several new developments have taken place in the field of UAN production in response to more stringent environmental requirements for nitrogen fertilizer plants and the automotive industry. The focus has

been on addressing particulate emissions as acidic aerosols and has required new technology developments to reduce the environmental footprint of the UAN plant. The integration of Stamicarbon's MMV scrubber in the urea/UAN plant design has further optimised the UAN

production, turning a disadvantage into an advantage.

Another major development is the ongoing trend in the automotive industry for compulsory application of diesel exhaust fluid (DEF) in diesel engines worldwide. Both environmental developments have created a unique opportunity for UAN producers to diversify their product range, mitigating seasonal variations and product price fluctuations.

Integrated emissions control

Urea dust and ammonia emissions from a finishing section are often reduced by wet scrubbing with water. However, in spite of its apparent high physical solubility, ammonia is not well absorbed at the low concentrations at which it is present in the plant exhaust. That means that, if ammonia is to be reduced efficiently to low levels, acidic scrubbing has to be used. Nitric acid (HNO_3) is one of the most commonly used acids, so ammonium nitrate salt solution is generated. The ammonium nitrate can be fed to a urea ammonium nitrate (UAN) or ammonium nitrate (AN) plant. With optimal integration between these plants, no waste streams are generated. Combining the AN solution from the acidic scrubbing with urea to form liquid UAN is especially attractive if

Fig. 1: Main emission sources in gas and liquid phase

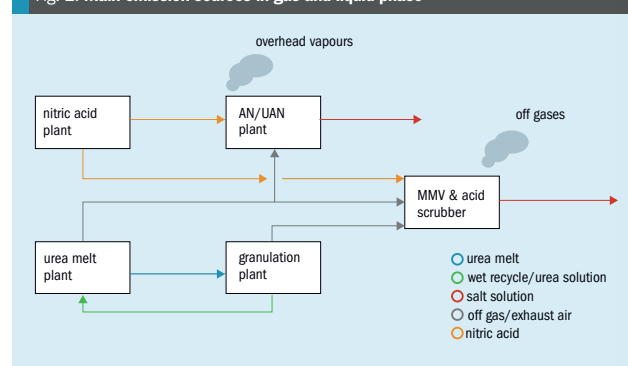


Table 1: Emission sources

Emission source	From urea melt plant	From AN/UAN Plant	From granulation plant
Gaseous	Off-gases (containing ammonia, carbon dioxide, water and traces of N_2 and O_2).	Overhead vapours resulting from neutralisation of nitric acid with ammonia.	Exhaust air containing urea dust and ammonia vapour.
Liquids		Acidic condensate from the off-gas treatment section.	Salt solution from acidic scrubbing.

Fig. 2: Liquid waste stream integration

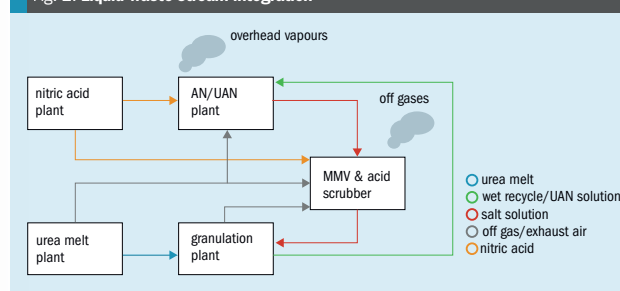
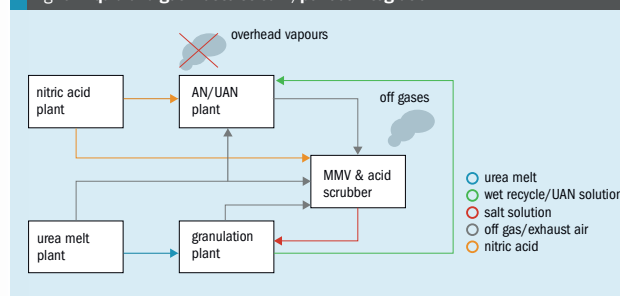


Fig. 3: Liquid and gas waste stream, perfect integration



the client already has a UAN plant on site. Proper system design of all waste streams of an industrial park can lead to a high level of integration between the plants, with further reworking of the salt within the site including emissions reduction. This type of integrated approach avoids waste streams outside the site limits.

A typical industrial park can include a nitric acid plant, a urea melt plant, an AN/UAN plant and urea granulation plant. From every plant it is possible to identify several emission sources in the gas and liquid phase (Table 1 and Figure 1). Three options, depending on how the liquid and gas waste streams are integrated within the plants will be considered.

Liquid waste stream integration

In this concept, acidic salty condensate from the off-gas treatment section of the AN/UAN plant is used in the scrubbing system as make-up water, reducing the amount of process condensate needed. Conversely, a recycle (containing 43 wt% urea, 55 wt% water and 2 wt% AN) from the granulation is routed into the AN/UAN plant. In that way, no liquid waste streams

are sent to battery limits (see Figure 2). The benefits of the integrated liquid waste stream option include:

- lower, optimised capex with less equipment in an integrated solution
- no additional capex involved for equipment if a revamp is considered
- liquid waste streams are completely reprocessed within the battery limits of the site
- lower total amount of process condensate needed.
- in case of Stamicarbon nitric acid technology, no acidic condensate waste stream needs to be recycled to the nitric acid plant any more.

However, in this configuration one emission point in the AN/UAN plant, emitting overhead vapours, is still not integrated.

Gas and liquid waste stream integration

In the second concept, further integration is introduced which completely eliminates emissions coming from the AN/UAN plant. This concept routes the ammonia off-gas coming from the AN/UAN synthesis section directly to the scrubbing

system of the granulation plant equipped with MMV technology (Figure 3). This MMV technology is capable to achieve emissions levels below 5 mg/Nm³ for particulate matter and 17 mg/Nm³ for NH_3 . This solution leads to a significant reduction in the amount of equipment required, as all the equipment items present in the off-gas treatment section of the AN/UAN plant can be eliminated.

Normally the off-gases from an AN/UAN plant are washed by a circulating acidic liquid solution in a packed-bed column in order to reduce the amount of ammonia released to the atmosphere. The gases leaving the packed-bed column still contain entrained nitric acid and ammonium nitrate, which are captured in an acidic condenser. The generated acidic condensate is partly recirculated to the AN/UAN plant (some into the neutraliser and some into the UAN mixer) and the rest is sent to battery limit.

The excess of acidic salty condensate sent to battery limit represents the liquid waste stream of a typical Stamicarbon AN/UAN plant. If the off-gases from the neutraliser are directly routed to the scrubbing system of the granulation unit, no acidic salty condensate is produced and the complete off-gas treatment section from the AN/UAN plant is eliminated. No ammonia is emitted to the atmosphere. This further simplifies the AN/UAN plant without affecting the size of the scrubbing system of the granulation section or its emission values. Removing the off-gas treatment gives a further size reduction in the synthesis section of the AN/UAN plant: no recycles go back into the synthesis, and nothing apart from the final product is sent to battery limit.

Based on this configuration the following advantages can be obtained:

- a capex reduction of approximately 40% in equipment costs in the AN/UAN plant.
- one single emission source in the total industrial park with emissions levels below 5 mg/Nm³ for particulate matter and 17 mg/Nm³ for NH_3 .
- no emissions from the UAN plant.
- elimination of the emission point in the AN/UAN plant, optimising the dispersion model required as input for the air permit.

UAN production from acidic scrubbing – no by-product

Another possible option is the production of UAN directly from the scrubbing system of

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Fig. 4: UAN production from scrubbing system

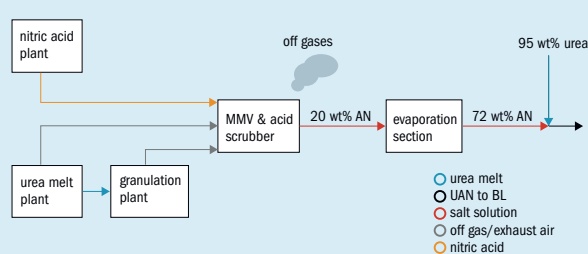
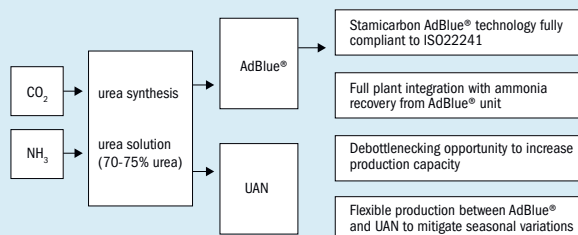


Fig. 5: Advantages of producing AdBlue® in combination with UAN



the finishing section. The by-product (AN salt solution) is routed to a dedicated evaporation section to be further concentrated and mixed with urea solution (Figure 4). This allows for emissions abatement and UAN production in one step. The generated salt from the scrubber is used for UAN production. In this way no liquid waste streams are sent to battery limits and UAN production is performed without the need for an AN/UAN plant. Only one single emission source is left in the total site and no extra ammonia emissions are generated, whilst the feedstock requirement for acidic scrubbing is made redundant by smart integration.

Co-production of UAN and DEF

A major recent development in the field of urea has been the rise of DEF, an aqueous urea solution of high-purity urea and deionised water. DEF is the common name for various trademarks and regional indications such as AdBlue®, Air1®, Arla-32, AUS-32 and NOx®. The function of DEF is to reduce NOx emissions in diesel engine exhaust using selective catalytic reduction

(SCR) technology. The application range varies from diesel cars to trucks, trains, ships and power stations. The market for DEF has grown rapidly over the past few years and is expected to grow substantially over the coming decade. More stringent environmental requirements for diesel engines, global endorsement of the concept and a wider application range towards private cars, ships and trains have boosted demand and DEF has become a welcome high margin diversified product for fertilizer manufacturers.

The synergy between production of UAN and DEF lies in the product specification of both products, as the urea feedstock is similar. The similarity allows for simultaneous continuous production of both UAN and DEF from the same urea synthesis plant, with the possibility to divert urea solution directly to both outlets. The key to the solution lies in the strict specification of DEF within the ISO-22241 specification with respect to the maximum allowable ammonia content. As the ammonia content within UAN is less relevant for the fertilizer application, removing and recovering the excess

ammonia is the solution to produce the purified urea solution. Controlling the ammonia content can be achieved by optimising the operating settings of the existing synthesis and recirculation process, or by the use of steam stripping.

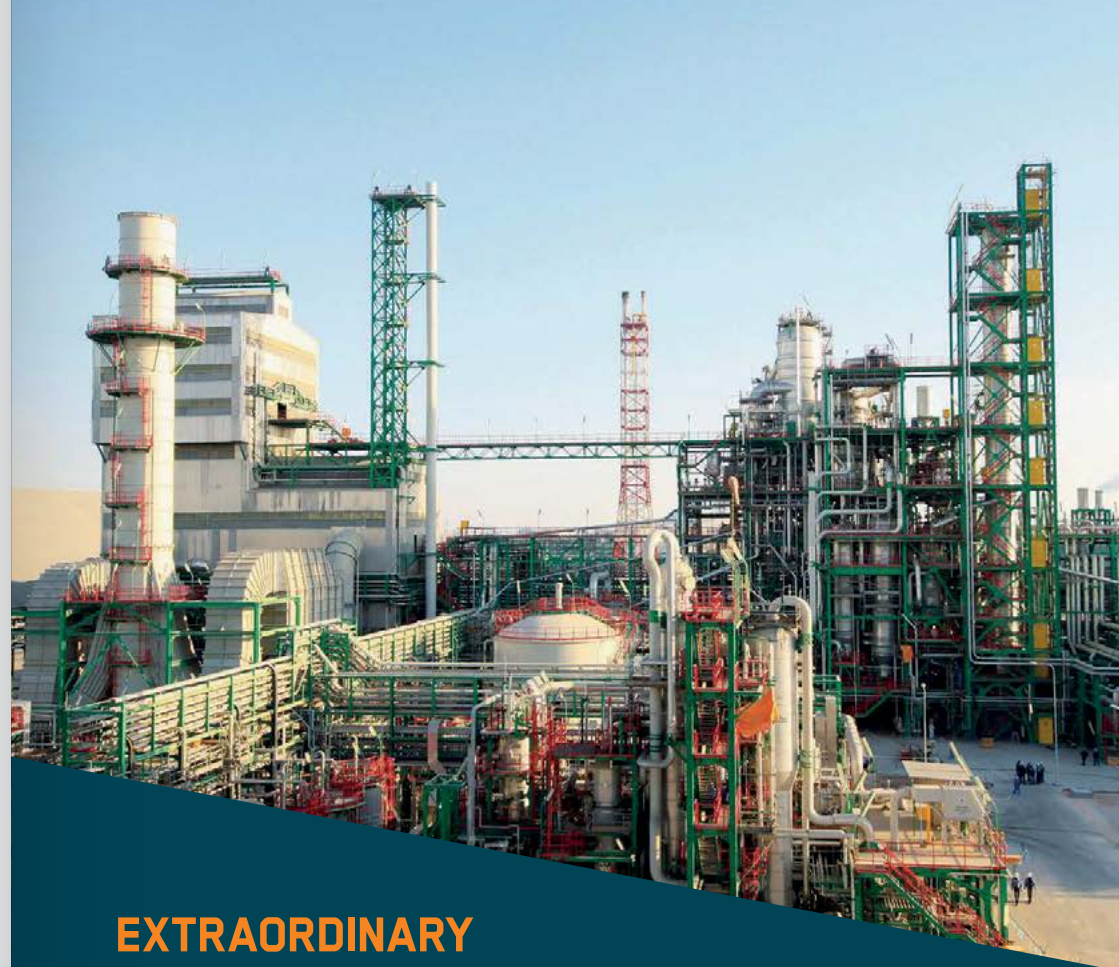
This new feature of Stamicarbon's technology creates the possibility to adjust any regular UAN plant to a dual purpose plant for the production of DEF as automotive, marine and intermediate grade, fully compliant with the ISO-22241 standard. In the recent years, Stamicarbon has conducted various plant upgrades of this kind in order to enable producers to diversify their product portfolio.

Conclusion

On plant level stringent emission requirements have instigated new innovations in the urea granulation plant with Stamicarbon's MMV scrubber, capable of achieving performance figures below 5 mg/Nm³ on particulate matter and 17 mg/Nm³ for NH₃. The combination of the MMV technology in an existing or new UAN plant eliminates emissions from the AN/UAN plant by integrating the emission point within the industrial complex. Consequently it improves the dispersion model required as input for the air permit, while using the ammonium nitrate as feedstock for the acidic scrubber. It is the type of technological synergy that can justify investments in environmental measures.

New stringent environmental regulations in the car industry have resulted in new technological developments and opportunities for UAN plants. The growing demand for DEF has become an attractive business opportunity for UAN producers, as the existing UAN plant can be adjusted for the production thereof with relatively minor investment, thereby creating access to a seasonal independent premium niche market.

Reduction of emissions from fertilizer plants should be considered for the overall site overlooking the individual process plants. New technological developments allow smart and efficient integration thereof with the purpose to reduce the environmental footprint, while maintaining a positive business case. With increasing interest of UAN in new markets, the integration of UAN production with existing urea plants could result in an interesting portfolio diversification for producers. ■



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Nitric acid technology trends



PHOTO: tkIS

L. Nielandt, J. Mathiak and M. Groves of thyssenkrupp Industrial Solutions AG provide a brief history of nitric acid production technology and discuss current challenges to reduce plant emissions. Recent abatement achievements during start-up and partial load are presented.

A brief history

The oxidation of ammonia over a platinum catalyst to nitrogen oxides, and their absorption in water to form nitric acid was first carried out in 1838 by C.F. Kuhlmann. However, this discovery was not commercialised at that time as ammonia was too expensive compared to the Chile saltpeter used to manufacture nitric acid in those days.

The history of the modern nitric acid process really began in 1901 when Prof. Wilhelm Ostwald established the ammonia oxidation conditions necessary for high nitrogen oxide yields. The first plants using the Ostwald process were started up in the first decade of the 20th century.

With more than 60 plants built since 1980 thyssenkrupp Industrial Solutions

(formerly known as Uhde) is nowadays the leading licensor for nitric acid technology worldwide.

As early as 1905, Dr. Friedrich Uhde, the founder of Uhde GmbH, designed and constructed a nitric acid pilot plant in co-operation with Ostwald.

Since the company's foundation in 1921, the number of plants for the production of weak and concentrated nitric acid as well as ammonia combustion units for nitrogen oxide production and absorption plants for nitrous waste gases, which have been designed, constructed and commissioned by thyssenkrupp under a variety of environmental conditions, total some 200. The company builds its knowledge and expertise on more than 100 years of experience and is committed to ongoing optimisation and improvements of its processes.

Market and uses of nitric acid

Nowadays 60 million tonnes of nitric acid of various concentrations are produced worldwide yearly. Approximately 85% of this is used as an intermediate in the production of nitrogenous mineral fertilizers, primarily ammonium nitrate and its derivatives such as calcium nitrate and urea ammonium nitrate solution. The remainder goes into the production of porous prilled ammonium nitrate for the mining industry, and as an intermediate for the production of special chemicals such as adipic acid. Nitric acid is not a traded commodity, it is typically processed where it is produced. Plant capacities are therefore defined by the downstream markets.

The market for nitric acid plants is quite limited with approximately three nitric acid

Above: New nitric acid plant at Nitrogénművek in Hungary.

Fig. 1: Nitric acid plant capacities since 1990

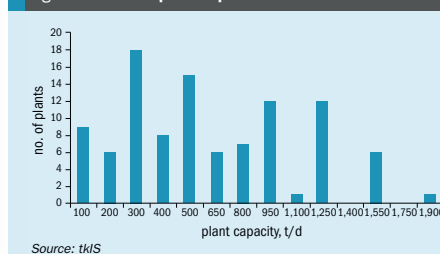
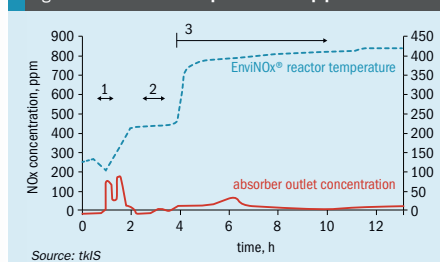


Fig. 3: Brownish plumes during start-up.

Fig. 5: Emissions with improved start-up procedure



plants built every year worldwide.

Historically plant capacities were rather small due to limitations of process pressure or manufacture of equipment. With the development of the modern dual pressure process higher capacities became more feasible. Fig. 1 shows the result of a survey of nitric acid plant nameplate capacities. More than 100 plants built after 1990 are included.

Small plant sizes (up to 500 t/d) are primarily required for the production of porous ammonium nitrate for mines close to the nitric acid plant. Special purpose nitric acid plants (e.g. azeotropic acid, i.e.

68-69% acid concentration) are also typically small scale plants.

Nitric acid plants in fertilizer complexes are typically designed for larger nameplate capacities, typically in the range of 900 to 1,200 t/d. These sizes utilise economies of scale in terms of investment costs as well as for operation costs.

There is a third group of nitric acid plant capacities, around 1,500 t/d for liquid urea ammonium nitrate (UAN). This capacity is required for state-of-the-art ammonia and urea complexes. The largest nitric acid plant has a capacity of 1,830 t/d and is also a thyssenkrupp reference.



PHOTO: tkIS

Fig. 2: New nitric acid plant at Nitrogénművek in Hungary.

Fig. 4: Typical emissions during start-up of a nitric acid plant

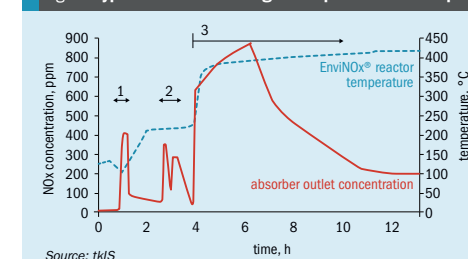
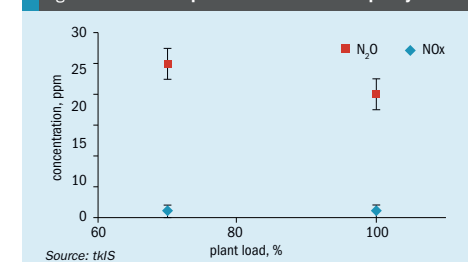


Fig. 6: Emissions at partial load and rated capacity



Above capacities of 500 t/d the vast majority of nitric acid plants are dual pressure plants.

Current nitric acid projects

Recently thyssenkrupp has commissioned two nitric acid plants. At the beginning of 2017, a 1,150 t/d nitric acid plant with 60 wt% concentration came on stream at Nitrogénművek in Hungary (Fig. 2). In the spring of 2017, a 1,500 t/d nitric acid plant with the same concentration was started up for Iowa Fertilizer Company (IFCo) in Wever, USA. The start-up data of

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these plants and others are utilised for the evaluation shown below.

Preparations for the commissioning of a 1,200 t/d nitric acid plant at Grodno Azot, Belarus are ongoing on and in the summer of 2017, the basic engineering for Pulawy in Poland was started.

Current challenges

Nitric acid technology has reached a high degree of maturity already. Since the beginning of this century the focus has been on emission abatement for NOx as well as for N₂O. thyssenkrupp has successfully introduced its EnviNOx® technology for emission reduction which sets the standards according to best available technique (BAT)¹. During rated load operation N₂O removal rates of 99% and above and NOx outlet concentrations below 1 ppm can be achieved.

Current challenges are in the dynamic behaviour of nitric acid plants during start-up and also during partial load operation. There are two drivers behind this market force – public interest of neighbours who are concerned about brownish plumes during start-up (see Fig. 3) and some local authorities who have recently changed the emission limits, not only to low concentrations during normal operation, but also to a certain quantity of CO₂ equivalent that may be emitted to the environment over a year.

First theoretical considerations on start-up were presented at the 2016 Nitrogen+Syngas conference². Today, data from industrial scale plants are available for the evaluation of the effects.

The emissions during start-up (see Fig. 4) are typically as follows: Initially the compressor train is started and emissions occur which depend on the situation after the previous shutdown. Nitrogen oxides in the plant are blown out of the plant stack by the flow of air. This is the first peak (1) during start-up. Filling the absorption tower leads to the next peaks (2). Immediately after ignition NOx is generated in the ammonia burner and most of it is converted to nitric acid in the absorption tower. However, the acid concentration profile is not completely established in the column and the operating pressure is lower. This leads to a higher NOx concentration at the absorption tower outlet (3) and to roughly the same emissions from the stack as a consequence. Afterwards NOx concentrations approach the steady

state level of approximately 200 ppm at the outlet of the absorption column. Once the tail gas reactor is put into operation the stack outlet concentration is reduced to minimum figures.

There are several independent possibilities to reduce the peaks. The first peak (1) can be minimised if the compressor train is restarted and the plant purged with air shortly after the ammonia burner shut down. If the abatement system is still hot efficient NOx removal can be ensured. A reduction of the emissions from about 400 ppm down to 20 to 100 ppm can be achieved depending on the shutdown duration.

The following emission peaks (2) are observed if the absorber is filled with concentrated, unbleached acid in the bottom part, diluted acid in the middle section and water in the top area. The flow of air through the unbleached acid strips the dissolved NOx and leads to emissions.

Reduction of the peaks can be achieved if the middle and top trays of the absorption column are filled with water and the lower section is filled with bleached acid. The absorption tower outlet peak concentration can be reduced from approximately 400 ppm down to approximately 200 ppm NOx.

It is apparent that the emissions and the temperatures after ignition show a very steep slope. Every minute of advanced abatement system operation would reduce markedly the emission peak value. The first approach is to start ammonia supply to the abatement system at lower temperatures. A detailed analysis of ammonia slip and temperatures in the expander must be made. This option can reduce the peak value of NOx emissions from roughly 800 ppm down to 60 to 100 ppm.

Early ammonia supply might already be possible during the absorption tower filling process if the minimum temperatures are met at the DeNOx-unit. This would reduce the emission peak (2) significantly.

Fig. 5 shows the emissions at the stack with advanced air blowing, improved absorption tower filling and early ammonia supply.

Besides the reduced emission peaks, the total amount of NOx released to atmosphere can be reduced significantly. However, it is required to ensure no ammonia leaves the DeNOx system if the tail gas expander is not protected against ammonium nitrate formation. Integration of the emissions over time gives approximately

“2,000 ppm x hours” as per the typical start-up and only roughly “300 ppm x hours” applying the improved start-up procedure.

To ensure that ammonium nitrate generation had been avoided, the expander and adjacent piping were visually checked after operation. No trace of ammonium nitrate was found in this investigation. This is an important verification that ammonia can be introduced to the EnviNOx® reactor earlier, if special care is taken with observation of ammonia slip and the temperature profile in the tail gas system.

Additional reduction of emissions during this phase of start-up is possible by advanced supply of ammonia at even lower temperatures. This requires even more sophisticated observation of ammonia slip and temperatures downstream of the abatement system.

During partial load the temperature and the flow in the tail gas reactor are lower. From a catalytic point of view these have opposite effects on the conversion in the reactor. On the one hand, reduced flow leads to higher conversion, but on the other hand, the lower temperature leads to lower conversion. As a consequence, the reactor needs to be designed and cross-checked for all operating conditions. Fig. 6 indicates the performance of the EnviNOx® reactor in a nitric acid plant at partial load compared to rated load. The emissions can be tuned by adjustment of ammonia and natural gas supply to fulfill authority regulations.

Installation of a heat source in the tail gas heating section of a nitric acid plant would optimise not only the start-up sequence but also the energy efficiency during normal operation and the efficiency during partial load operation. Steam injection, indirect heating or direct heating by a burner would provide technical options to increase the flexibility and efficiency of the plant.

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“Oxyboost Technology” increasing efficiency

A new concept for improving the absorption efficiency of nitric acid plants has been developed by Messer Group. This concept uses pure oxygen for the oxidation of nitrous acid (HNO₂) and nitrous gases mainly in the liquid phase. Benefits include: reduced NO_x emissions, less down time, up to 10% additional production, low investment costs and less ammonia needed for DeNO_x.

The majority of the global industrial production of nitric acid utilises catalytic oxidation of ammonia, followed by nitric oxide oxidation and absorption in water. Although the process was developed more than 100 years ago, the production process has been improved and developments continue to take place to improve raw material and energy efficiencies.

To date, Messer has not been greatly involved with application processes for the chemical industry. To find process technology entry into the chemical industry, nitric acid production and its optimisation with oxygen was chosen.

Messer thus started an R&D project some years ago to investigate the possibility of oxygen applications to increase the absorption efficiency in nitric acid production. After several smaller trials at different nitric acid plants throughout Europe, the concept was applied at a large industrial application at Azomures S.A. in Targu Mures, Romania which operates three nitric acid lines.

Challenges for Azomures Line II

Due to strict new regulations from the state authorities with regard to NO_x emissions Azomures faced a difficult situation with their nitric acid plant, especially with Line II. Azomures even thought about shutting down this facility as they did not see an economically feasible way of operating the plant whilst adhering to the new restrictions. At this point in time Messer proposed their technology and Azomures decided to perform a test run to verify if the problem could be solved using Messer's recently patented “Oxyboost

Technology”. The alternative would have been a new DeNO_x catalyst which would also have required a major investment or even total replacement of the existing DeNO_x plant.

Industrial production of nitric acid

Industrial production of nitric acid takes place via the catalytic oxidation of ammonia (Ostwald process). In the first step of this process, ammonia is combusted with air. Ammonia and atmospheric oxygen are essentially converted to nitric oxide and water vapour in a reactor with a gauze catalyst usually consisting of precious metals:



Atmospheric oxygen is used in this reaction. The use of pure oxygen was tried several times in the past, with the result that an approx. 10% increase in capacity could briefly be achieved. However, this application was then stopped due to technical and economic difficulties, the main one being the reduced lifetime of the catalysts.

In general, the use of oxygen in the secondary air is the preferred method of increasing capacity. In this case, the additional capacity comes from the secondary air thus saved which can be used to provide additional primary air.

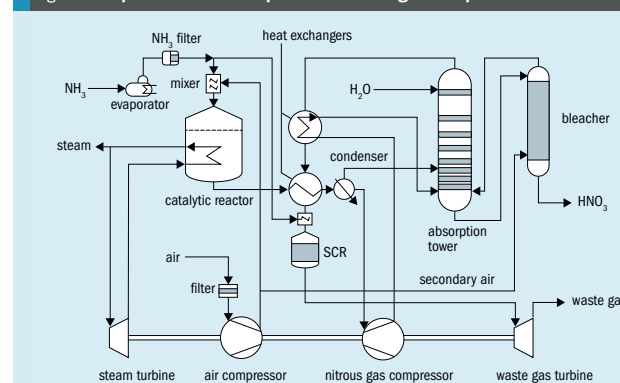
The product mixture resulting from reaction (1) is then cooled in several steps. The NO content is thereby oxidised to NO₂ with further atmospheric oxygen. Some plants use their own “oxidation columns” for this, as can be seen in the photograph of the Azomures plant in Targu Mures (Fig. 1).

Subsequently the process gas is cooled down in a condenser to a temperature at which water condenses. A part of the nitrogen dioxide formed from nitric oxide and oxygen reacts with water forming an aqueous nitric acid solution containing, moreover, nitrous acid (HNO₂) and nitrogen oxides in dissolved form. The remaining, undissolved gas mixture is fed to one or more absorption columns. Here, a portion of the gaseous nitrogen monoxide is oxidised with oxygen to nitrogen dioxide and its dimer dinitrogen tetra oxide, and then reacted with water to form nitric acid:



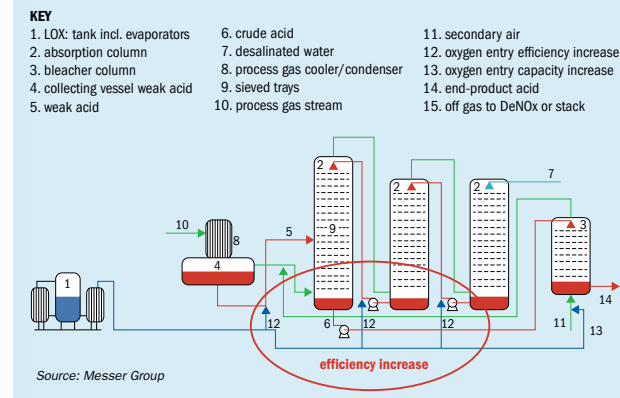
Fig. 1: Nitric acid plant Azomures Line II: Oxidation towers (2 brownish coloured columns at the front) for cooling and oxidation of NO to NO₂; and absorption columns (2 columns at the back, of which just one is visible) for absorption of NO_x.

Fig. 2: Dual pressure nitric acid process with a single absorption column



Source: Messer Group

Fig. 3: Schematic process flow chart of nitric acid production with four possible introduction points for the Oxyboost Technology

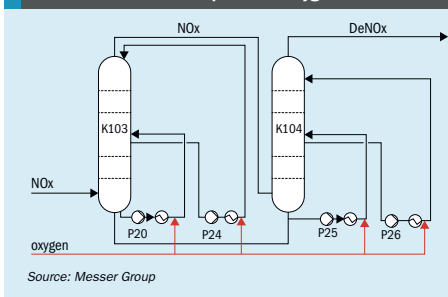


Source: Messer Group

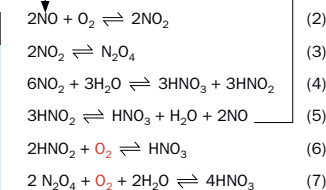


Fig. 4: Heat exchangers (left) and buffer vessels for the pumps (right) which are filled with weak acid from the absorption process from nitric acid Line II.

Fig. 5: Schematic flow chart of the last two absorption columns at Line II with 4 points for oxygen introduction



Source: Messer Group



The water usually runs through the absorption tower countercurrent to the ascending gas stream. The nitric acid collects at the bottom of the absorption tower in an aqueous solution. The thus formed “crude acid” is bleached in a further process step, usually in an additional column. This operation is performed by stripping, using air which enters the bottom of the column and frees the crude acid of the nitrous gases it contains. Fig. 2 is a schematic diagram of the Ostwald process. It shows the so called “dual pressure process” which is the best available technology, especially when the objective is large production capacities.

However, despite dual pressure technology being the most important technology for the construction of new plants, there are still many mono pressure plants in operation. Some of these work at quite low pressures (0.4-5 barg) and therefore need more than one absorption column in order to achieve reasonable absorption efficiencies. Oxyboost technology aims, in the first step, to boost the absorption efficiency of these mono pressure plants with low pressure at absorption.

Oxyboost Technology for nitric acid production

When using Oxyboost Technology, pure oxygen is added to the liquid nitric acid (weak acid). By adding pure oxygen, reactions (6) and (7) listed earlier occur in the

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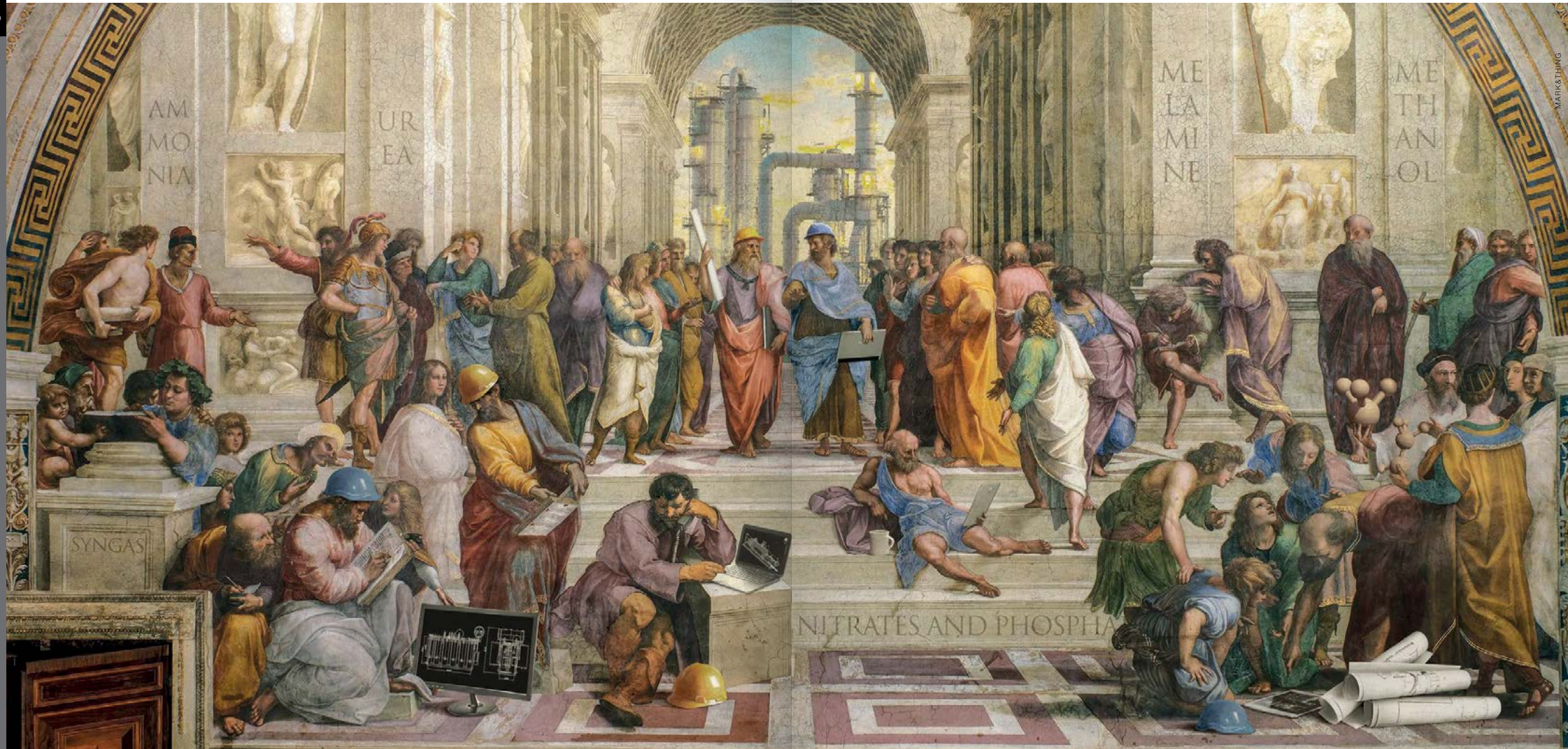
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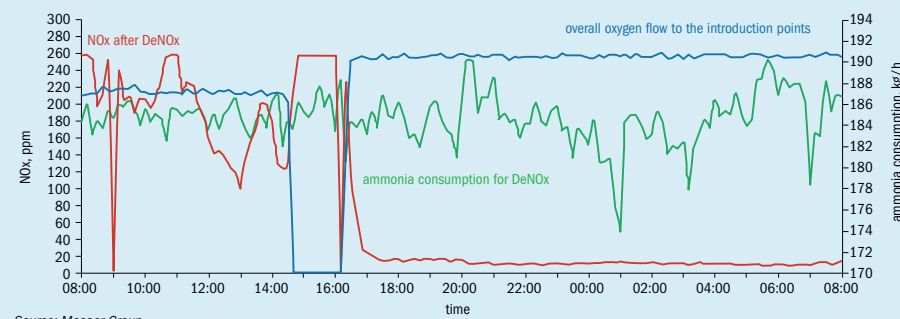
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Fig. 6: Trial runs at Azomures Line II



Source: Messer Group

liquid phase so that additional product acid is produced here. Without the addition of oxygen both of these reactions run, if at all, with a minimal reaction speed.

Oxyboost Technology aims to feed oxygen, through optimal entry devices, into the liquid phase of the evolving aqueous nitric acid.

Fig. 3 shows a schematic diagram for the use of this method. Here oxygen is introduced directly into the ascending line of the weak-acid-carrying pipelines, which connect the three absorption towers together.

Azomures process description

Nitric acid Line II at the Azomures site is the oldest of the three units operated by Azomures. It has a capacity of 730 t/d (100% nitric acid) and was commissioned back in 1968. It uses Stamicarbon mono pressure technology with an average pressure of approx. 4 barg. The absorption takes place in four consecutive columns of which the first two are oxidation columns. The technology uses trays with "Raschig" packing for the absorption columns instead of sieved trays, which is the usual case for absorption columns at nitric acid plants. A further difference is the cooling of the weak acid which takes place outside the columns in plate heat exchangers which are cooled by rinsing cooling water outside of the cooling plates.

This means that weak acid is extracted from each plate and flows to a buffer vessel from which it is pumped through the heat exchanger and back to the top of the packed tray where it was previously extracted.

Fig. 4 shows the heat exchangers and buffer vessels for the pumps which are filled with weak acid from the absorption process.

Oxygen introduction at Azomures

Due to the external cooling, large quantities of weak acid are pumped from each plate and sent back after external cooling to the top of the same plate. Fig. 5 shows a schematic flow chart of the last two absorption columns at Line II with the four points for oxygen introduction. The introduction took place by means of appropriate introduction devices and patented technology developed by Messer Group.

Results

The trial run lasted 4 weeks in the period between September and October 2015. Fig. 6 shows what was probably the most decisive day of the whole test period.

After this very positive result which occurred during the trial period, Azomures decided to invest in a fixed installation for the tested Oxyboost Technology (Fig. 7). Both partners signed a delivery contract for the equipment together with a long term oxygen supply contract.

Commissioning took place in September 2016, since when the equipment has been running continuously.

The following improvements have been achieved by introducing Oxyboost Technology on Line II in Azomures:

- Substantial reduction of NOx emissions, keeping to emission limits;
- Less down time due to keeping to limits
- Additional production (proportional to the improved absorption)



PHOTO: MESSER GROUP

Fig. 7: Nitric acid Line II at Azomures site with the final installation of Oxyboost Technology.

- Less ammonia needed for DeNOx
- Low investment costs
- Messer delivers a full custom-made package for fixed installations after checking viability
- Capacity increase of nitric acid plants (up to 10 % possible)

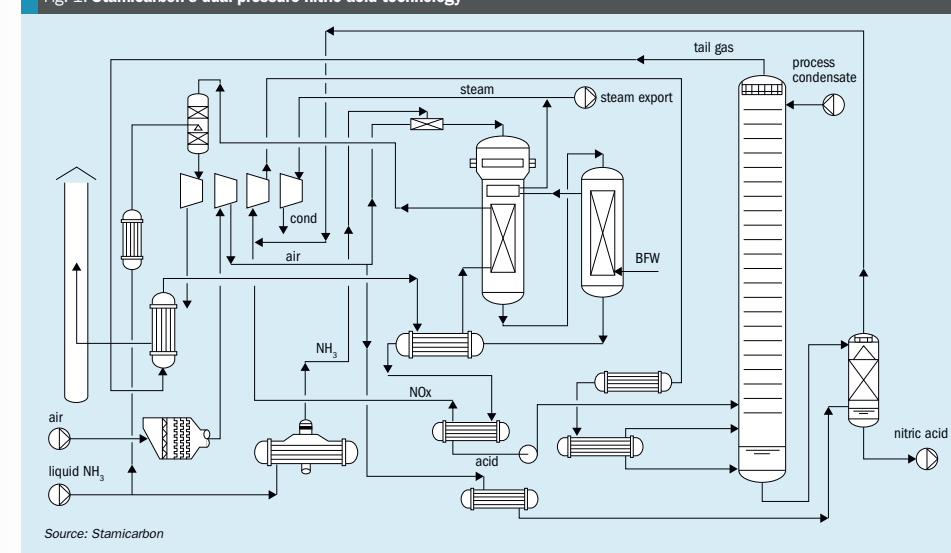
Acknowledgement

This article is based on the Nitrogen+Syngas 2017 paper "Oxyboost Technology" – Increasing efficiency of nitric acid plants through introduction of oxygen by Walter Bachleitner of Messer Group GmbH, Laurentiu Popovici of Messer Romania Gas and Csaba Boldizar of Azomures S.A., presented in London in March 2017.

High energy recovery with low emissions

Stamicarbon is launching its renewed dual pressure nitric acid technology, developed for highest energy recovery with low emissions in order to accommodate an increased customer need for lowest total cost of ownership of nitric acid plants.

Fig. 1: Stamicarbon's dual pressure nitric acid technology



Source: Stamicarbon

Stamicarbon has been licensing a range of nitrate technologies since the 1930s and has designed over 40 nitric acid plants in the past. By applying many decades of experience and retaining the reliability and operability from industrially proven elements that are appreciated by its customers, Stamicarbon is now able to offer a dual pressure nitric acid technology design for single train capacities of up to 1,600 t/d.

Stamicarbon nitric acid process

A dual pressure configuration allows the combination of medium oxidation pressure with high absorption pressure which

makes the nitric acid production more efficient. In the Stamicarbon process, the oxidation section operates at about 5 bar(a) whilst the absorption sections runs at a pressure of about 10 bar(a).

Typically the dual pressure process is used for capacities between 600 and 1,600 t/d.

Stamicarbon's dual pressure nitric acid process, as described here, will also be presented at the ANNA conference in Austin, Texas (1-6 October, 2017).

Process description

The total process flow diagram is shown in Fig. 1. Air is compressed by the air compressor to approximately 5 bar(a)

reaching a temperature of about 220°C. Prior to entering the air compressor, the air is preheated and filtered. Ammonia is received from battery limits in liquid state and evaporated by means of cooling water. Ammonia is then superheated and filtered.

Subsequently, air and NH₃ are mixed in a static mixer and sent to the ammonia burner. In the ammonia burner, NH₃ is oxidised to NO on Pt/Rh gauzes, small amounts of N₂ and N₂O are formed as side products. The temperature at the gauzes is controlled at about 885°C. This temperature is selected as it combines an optimal ammonia conversion with low Pt losses.

Homogenous distribution of the NH₃/air mixture over the Pt-gauzes is essential to

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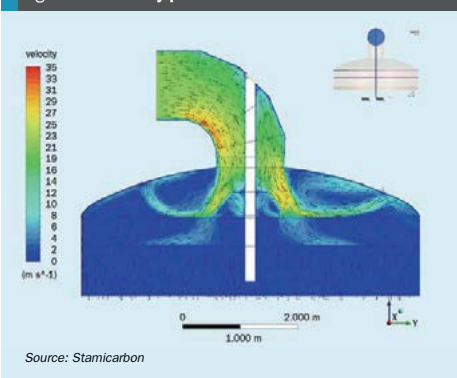
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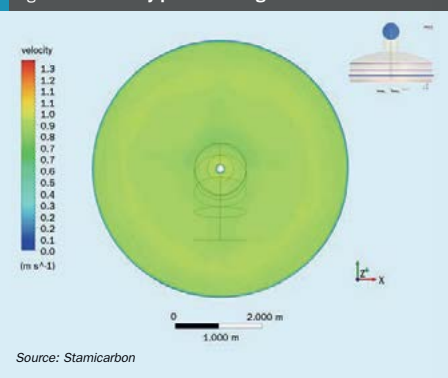
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Fig. 2: CFD velocity profile in the burner head



Source: Stamicarbon

Fig. 3: CFD velocity profile in the gauzes



Source: Stamicarbon

guarantee optimum operation. To achieve this, Stamicarbon has developed a distribution system with perforated plates that combines a uniform gas distribution with low pressure drop. In addition, this type of distribution system reduces notably the risk of pre-ignition when compared with other types of distributors like deflector plates.

Stamicarbon has performed CFD studies and the results are shown in Figs 2 and 3. The velocity profiles show a homogeneous distribution of gas over the gauzes.

The produced NO oxidises to NO₂ in the gas phase of the downstream sections which will lead to the formation of HNO₃ downstream. The heat released by the oxidation reaction is first used to produce high pressure steam and heats tail gas up to 480°C.

Downstream of the waste heat boiler, the nitrous gas is further cooled down in a gas/gas heat exchanger to 160°C before entering the low pressure cooler condenser. In the low pressure cooler condenser, the gas stream is cooled and a weak acid solution is condensed and sent to the oxidation/absorption column. The cooled nitrous gas is compressed to 10 bar(a) by the NOx compressor reaching a temperature of 180°C and is cooled down by means of a BFW preheater to about 115°C. The last cooling step takes place in the high pressure cooler condenser where the gas is cooled and acid solution is formed at about 50°C.

The remaining gas enters at the bottom of the oxidation/absorption column which consists of a series of sieve trays. On the trays, NO₂ is absorbed in water which is introduced at the top of the column; while NO is oxidized to NO₂ in the free space

between trays. To promote the absorption and oxidation reactions, the heat released by the absorption and oxidation reaction is removed by cooling water coils mounted on the trays. At the bottom of the oxidation/absorption column, the acid from the trays is mixed with the acid from the high pressure cooler condenser and the acid, at the desired concentration, is sent to the last purification step, the bleaching.

In the bleaching column, the last traces of dissolved NOx are eliminated by means of secondary air. This secondary air is about 10% of the total air compressor intake and is cooled down by demi-water before entering the bleaching column. After passing the bleaching column, the secondary air is injected in the suction line of the NOx compressor to provide the required O₂ for the oxidation of NO to NO₂ in the high pressure part of the plant. The bleached nitric acid is sent to the storage tank.

At the top of the oxidation/absorption column, a gas stream containing N₂ with traces of NOx, N₂O and O₂ is heated up in three steps. First, in a tail gas/tail gas heat exchanger to 95°C; secondly, in a process gas/tail gas heat exchanger to 200°C and finally, in a process gas/tail gas heat exchanger to 480°C.

The next step in the process consists of eliminating the traces of NOx and N₂O before sending the tail gas to the atmosphere. For that, a tertiary abatement system is used where N₂O and NOx are catalytically decomposed. The concentrations of these two components at the outlet of the abatement system is less than 20 ppmv.

The tail gas is expanded in the expansion turbine where the in-line compressor

train configuration allows the usage of the released power to drive the air and NOx compressor. Additional power needed in the compressor train is provided by a steam turbine.

After expansion, the heat content of the tail gas stream is still sufficient to allow another heat exchanging step. The tail gas is cooled down to 110°C before being released to the atmosphere.

Highlights of Stamicarbon process

Maximum energy recovery

The main feature of Stamicarbon nitric acid process is the position of its third tail gas heater. The location of this heat exchanger in the ammonia burner vessel, below the super heater and before the boiler, allows tail gas temperatures of 480°C, 50°C above conventional nitric acid processes.

This has several advantages:

- N₂O decomposition is favoured by high temperature and takes place without the addition of an external agent like natural gas, which is common practice in conventional nitric acid process;
- higher power generation in the expansion turbine;
- extra heat recovery step before releasing the tail gas to atmosphere;
- all this leads to lower steam consumption of the compressor train and higher steam export from the plant.

The Stamicarbon heat exchanger layout is the result of a pinch analysis which maximises the heat recovery of the system. The process allows an export of HP steam of more than 800 kg steam/t HNO₃, typically at a pressure of 45 bar(a) and a temperature of 450°C.

Low emissions

NOx and N₂O are considered harmful species for the environment contributing to acid rain, the greenhouse effect and ozone layer depletion. Stamicarbon's design has an optimum tail gas temperature for the decomposition of N₂O in a tertiary N₂O/NOx abatement system and allows the catalytic reduction of NOx and N₂O to concentrations below 20 ppm(v).

Commercially available iron zeolite-based catalyst is used in the first catalyst bed to decompose N₂O without the addition of an external reducing agent. This can only be accomplished in processes with a tail gas temperature of >430°C as in the Stamicarbon design. A similar catalyst is used in the second catalyst bed to reduce the NOx gas to N₂ and H₂O; in this case, the addition of a reducing agent (NH₃) is required.

The Stamicarbon N₂O/NOx abatement reactor is designed to guarantee maximum efficiencies with lowest pressure drop. The design of the process minimises the NH₃ slippage and ensures that no ammonium nitrate is formed during start-up and unsafe situations are prevented.

Corrosion prevention design

Stamicarbon nitric acid process does not require the use of "exotic" construction materials. By proper process design and cautious selection of process temperatures, the risk of acid condensation/re-evaporation is minimised. Stamicarbon's running reference plant shown in Fig. 4, has been in operation for more than 30 years with no major corrosion issues.

Safe to operate

During plant start-up, the risk of formation of ammonium nitrate and ammonium



Fig. 4: Dual pressure nitric acid technology plant in operation since 1989.

PHOTO: STAMICARBON

nitrite due to NH₃ slippage increases. Ammonium nitrite is formed when the environment becomes alkaline and undergoes an explosive decomposition when pH becomes acidic. The formation of ammonium nitrite is likely in the low pressure cooler condenser and in the first compartment of the oxidation/absorption column when a substantial amount of NH₃ may slip through. By continuous circulation of an acidic solution through these apparatus and additional start-up safety measures, Stamicarbon's design guarantees a safe start-up eliminating the risk of explosion.

Nitrogen fertilizer technology portfolio

Stamicarbon's current extended nitrogen fertilizer technology portfolio can support projects comprising an entire grass root urea ammonium nitrate (UAN) industrial complex including its own urea and nitric acid technology. For ANNA projects, including the production of CAN and/or AN (both LDAN and HDAN), Stamicarbon partners with INCRO from Spain, thereby combining its nitric acid technology with an established portfolio of nitrate finishing technologies.

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Boosting nitric acid plant capacity

Jean-François Granger of Casale discusses novel solutions to boost the capacity of nitric acid plants without modifications to the turbot set machine.

Debottlenecking a nitric acid plant, whatever the process, is always a question of introducing more air into the process and the capacity of the air compressor, expander and steam turbine to handle the increased load. In the case of a dual pressure process, a NOx compressor is added in order to operate the absorption part of the plant at a higher pressure than the reaction part. In so doing, both parts of the process are optimised with regard to opex and capex.

The most obvious way to debottleneck a nitric acid plant is therefore to re-work the turbo set machine. However, this also

brings some drawbacks with regard to certain aspects of operation:

- The turbo set machine is very costly, about 20% of the total price of the plant, and intervention of it is not without risks.
- There are not so many companies who can provide such equipment and very often it is only the original manufacturer that is able to perform the changes. This results in a lack of competition and in most cases a monopoly situation. Therefore, the cost of the intervention becomes completely out of control.
- The rework of these machines is specialised and requires highly experi-

enced people. The duration of the job is also quite long, usually one month or more. The plant stoppage for the debottlenecking will require at least another two weeks, resulting in a high loss of production. If we consider a 20% capacity increase and a 2 month stoppage of the plant for its implementation, the payback of the investment has to be increased by one year to compensate for this loss of production.

- A last concern is the turndown of the plant. Usually the plant is designed to be operated between 70 and 110% of the nameplate capacity. After a 20% increase

in capacity, this becomes 84 to 132% of original nameplate because of the limitations of the machine. In addition, the higher minimum capacity requires more imported steam for start-up.

For these reasons, Casale has always considered that debottlenecking a nitric acid plant should be done without touching the turbo set machine.

Let's now consider each component of the machine.

When checking the limits of the machine, the steam turbine has never been found to be limiting as its design limit is the start-up conditions when the expander is delivering much less power to the train than during normal operation. This does not apply in the case of an expansion of the full train as the start-up conditions would change to a higher flow meaning a higher demand on the steam turbine.

The expander has sometimes been found to limit the capacity, but there are a few options that can be considered to overcome this:

- directly vent the excess gas and thus accept the corresponding loss of energy;
- partly cool down the tail gas in order to recover some heat and reduce the volumetric flow to the expander down to its limit;
- add a parallel expander to run the new air compressor or to generate electricity.
- These three options all solve the issue of the limiting capacity with increasing capex and energy efficiency from top to bottom. An interesting point is that if in the first step the first option is chosen, subsequently one of the other two options can be easily implemented.

The air compressor is always the limiting factor. One should never forget that to produce nitric acid three raw materials are required: ammonia, air and demineralised water.

As up to 20% of the air is used to bleach the product, the addition of a dedicated air compressor allows all the capacity of the main compressor to be used to feed the burner. In this way, the capacity of the plant can be easily increased by 20%. The minimum capacity of the plant remains at 70% of original nameplate while the maximum rises to 132% making the range of the plant from 59 to 110% of the new nameplate capacity. In addition, as the start-up conditions remain the same, no additional imported steam is required.

In the case of a dual pressure plant, such a debottlenecking plan will overload the NOx compressor by 20%. In some cases, this is becoming an issue when it cannot compress the NOx gases produced in the LP part of the plant. The main option, developed and patented by Casale to overcome this is to divert part of the flow from upstream to downstream of the NOx compressor.

In dual pressure plants, the bleaching process is done in the LP part as the air is supplied by the air compressor at this pressure and then the gases leaving the

bleacher are mixed with the NOx gases from the LP cooler condenser upstream of the NOx compressor.

Casale's solution is to deliver the additional air by an air compressor supplying the bleaching air at a pressure slightly higher than the pressure at the bottom of the absorption tower. Then the gases leaving the bleaching tower are mixed with the HP NOx gases downstream of the NOx compressor. In this way the NOx compressor is relieved of its load by about 20% and thus the capacity of the plant can be increased by an equal amount.



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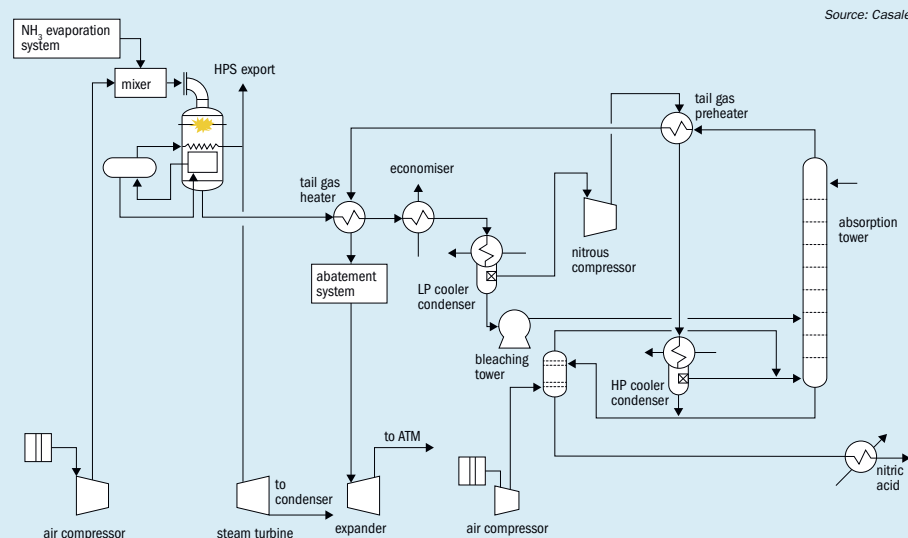
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Fig 1: Casale nitric acid debottlenecking with HP bleaching



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Catalysts for ammonia oxidation

A Bazhenov and **G Gushin** of Ekaterinburg Non-Ferrous Metals Processing Plant JSC (EZOCM) consider the design and technical characteristics of catalysts for ammonia oxidation in the manufacture of nitric and hydrocyanic acids, nitrogen fertilizers and hydroxylamine sulphate production.

EZOCM JSC has supplied catalyst gauzes for nitric and hydrocyanic acid production and hydroxylamine sulphate manufacture at various Russian companies for almost 70 years. Over the past 20 years intensive works have been performed at the plant aimed at the improvement and diversification of these products. For example, in 2012, catalytic systems with activated catchment gauzes were introduced, while from 2013-2015 came the construction of a new complex for catalyst gauze scrap processing, alongside modernisation of the production process for catalyst gauzes. In 2016 EZOCM catalytic systems were successfully tested together with a secondary bulk catalyst in Europe, and thus year there have been production tests on new catalytic systems with gauzes made from a high-strength alloy.

Gauzes

A modern catalyst system for ammonia oxidation in nitric acid and hydroxylamine production is represented by a set of gauzes, namely:

- catalyst gauzes;
- catchment gauzes;
- separating gauzes (used for catchment gauze separation and prevention of them sticking together during operation).

EZOCM JSC develops and produces the catalyst systems with due account for the specific features of a plant's operation and the customer's individual requirements. For this we use gauzes made of precious metals that differ according to alloy composition (see Table 1); weaving and knitting density (for woven and knitted gauzes); wire diameter; and mechanical characteristics.

Over the past few years the technology for producing catalyst gauzes has been changed. A pyrometallurgical process for processing of scrap catalyst gauzes has been implemented, instead of the previous hydrometallurgical and electrochemical technology (see Figure 1).

The previous method included four cycles: intake melting, bar forging for the electrolysis unit, scrap refining using electrolysis, then another melting step and testing of the cathode deposits produced. It should be noted that during electrolysis anodic residues are formed, which are melted once more. The new pyrometallurgical metal refining technology allows us to decrease precious metals losses by almost 1.5 times; to decrease power consumption; to increase the yield ratio by



Fig. 1: New flow process for scrap processing

Table 1: Chemical compositions of alloys for catalyst systems

No.	Alloy	Chemical composition, %					
		Pt	Pd	Rh	Ru	W	Y
1.	92.5%Pt, 4%Pd, 3.5%Rh	92.5±0.3	4.0±0.2	3.5±0.2			
2.	81%Pt, 15%Pd, 3.5%Rh, 0.5%Ru	81.0±0.7	15.0±0.5	3.5±0.4	0.5±0.3		
3.	92.5%Pt, 7.5%Rh	92.5±0.3		7.5±0.2			
4.	95%Pt, 5%Rh	95.0±0.3		5.0±0.3			
5.	92%Pt, 8%Rh	92.0±0.2		8.0±0.2			
6.	90%Pt, 10%Rh	90.0±0.3		10.0±0.3			
7.	90%Pt, 5%Pd, 5%Rh	90.0±0.3	5.0±0.2	5.0±0.2			
8.	Pd-5%W		94.0-95.5			4.30-5.97	0.03-0.20

Source: EZOCM



Fig. 2: Equipment for wire drawing and annealing



Fig. 3: Weaving machine



Fig. 4: The new section for the gauzes electrochemical activation

Fig. 5: Operational characteristics without bulk catalyst

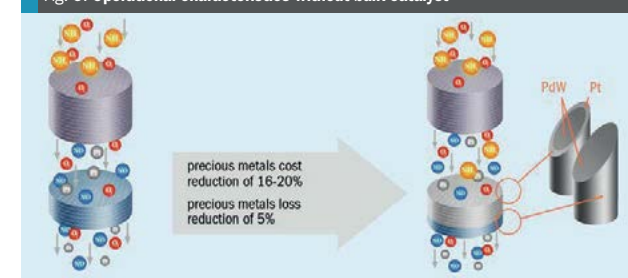


Table 2: Operational characteristics of systems with active catcher installed in UKL-7 unit

System modification	Precious metals weight in system, kg					Life, hrs	Output, t HNO ₃	Ammonia conversion, %	Loss of precious metals, mg/t 100% HNO ₃	
	Total	Pt	Pd	Rh	Ru				Pt + Rh	Total
UKL-7 unit, Ø 1,700 mm										
Standard	19.5	11.1	7.8	0.5	0.05	4,490	58,100	93.5	43	97
Active catcher	16.3	7.7	8.2	0.3	0.05	4,700	53,100	93.6	25	85
UKL-7 unit, Ø 1,930 mm										
Standard	18.2	9.5	8.2	0.4	0.05	5,393	74,800	92.4	38	92
Active catcher	16.8	8.5	7.9	0.4	0.05	5,500	76,800	93.0	25-30	80-85
AK-72 unit										
Standard	60.0	35.9	22.6	1.4	0.08	5,570	131,700	96.5	34	78
Active catcher	55.7	24.5	28.8	1.0	0.15	up to 6,800	150,000	95.5-96.0	15-25	60-70

Source: EZOCM

7.5%; and to decrease the time taken by a factor of ten.

In addition, the wire drawing section has been re-equipped with higher capacity wire drawing equipment, and a conveyor furnace for wire annealing has been commissioned which has resulted in the improvement of the wire's mechanical properties and an increase in performance (see Figure 2). New knitting and weaving machines with programma-

ble control have also been purchased, which makes it possible to complete orders quickly (Figure 3), and a new section for the electrochemical activation of catalyst grids and platinum plating of catchment gauzes has been built (Figure 4).

Finishing of catalyst gauzes by electrochemical activation was implemented in 1995 and includes the electrochemical deposition of dispersed platinum on

a specially cleaned gauze's surface. As a consequence the gauze's active surface is increased by ten or twenty times, and due to the increased platinum concentration the initial ammonia oxidation temperature is substantially decreased. Such gauze processing provides acceleration of catalyst gauze start-up as the production unit comes into operation and allows the system to rapidly reach its maximum shift and output values. It also eliminates the

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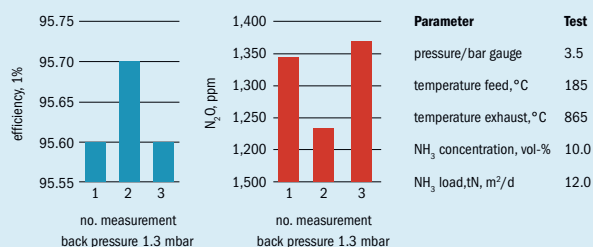
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Fig. 6: Operational characteristics without bulk catalyst



Source: EZOCM

necessity of additional grid processing before its incorporation into the unit, and allows the possibility of installation into the system catalyst gauzes with reduced (up to 81%) platinum concentration, even in areas which first meet the gas flow, while creating no problems connected with start-up.

Catchment systems

In 2012 we developed and implemented an electrochemical activation process for catchment gauzes made from the palladium-based alloy of Pd-5%W grade containing 5% tungsten. As in the case of the catalyst gauzes, a platinum layer 0.1 µm in thickness is formed on the surface of palladium-based catchment gauzes, due

to which their catalytic activity becomes no lower than the catalytic activity of platinum-based grids. This allowed the development of catalyst systems in which part of the catalytic platinum-based gauzes have been replaced by activated palladium-based gauzes which have a dual purpose: both ammonia oxidation and platinum catchment. Notably, as opposed to regular catchment gauzes, the activity of the activated palladium-based gauzes comparable to the activity of platinum-based gauzes from the moment of being put into operation.

The advantages of a system with an active catcher are illustrated in Figure 5. Systems with an active catcher have been tested in the UKL-7 and K-72 plants, and operational data are presented in Table 2.

Secondary catalyst

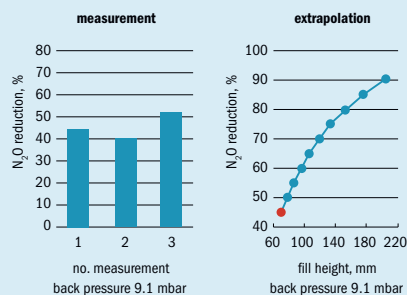
In view of the favourable effect that oxide catalysts have on catalyst systems operation and the flow process as a whole, in 2015 a new oxide block-type catalyst, YK-42-9 type, was developed and tested jointly with the Borekov Catalysis Institute in Novosibirsk. All investigations of the YK-42-9 catalyst effectiveness were performed at Borekov using a pilot flow reactor with an internal diameter of 11.2 mm and a reaction mixture feed speed of 880-890 l/hour at pressure of 3.6 atmospheres, which corresponds in operational terms to the K-72 unit. During the experiments it was established that the YK-42-9 catalyst increases the shift of the catalyst system by 1.0-1.5 % through temperature and gas flow equalisation along the whole reactor area, as well as via increasing the temperature of the front gauzes, while nitrous oxide emissions are decreased by 76%.

In 2016, testing of EZOCM catalyst systems with bulk catalyst were carried out. The purpose of these tests was to produce a high degree of control of nitrous oxide emissions into the atmosphere. Tests have been carried out on a standard catalyst system with different types of catalysts and different heights of filling layer. As can be seen from the graphs in Figures 6 and 7, use of secondary catalyst allows the reduction of nitrous oxide emissions by 70-90%.

Fig. 7: N₂O emission versus height and composition of the bulk catalyst

PGM secondary catalyst

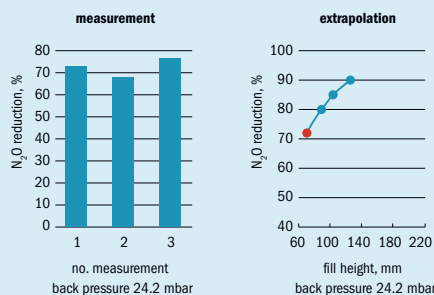
45% reduction @ 70mm → 90% reduction @ 205mm
Low back pressure @ 70mm (~9.1mbar)



Source: EZOCM

NPGM secondary catalyst

72% reduction @ 70mm → 90% reduction @ 125mm
Intermediate back pressure @ 70mm (~24.2mbar)



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Inspired by Vincent Van Gogh, *Paysage sous un ciel mouvementé*,
April 1889, oil on canvas, 60.5x73.7 cm.

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The future size of methanol plants

More and more, the size of a methanol plant is determined by construction considerations (modularisation, shipping limits on size or weight due to tunnels and bridges) or product sales (how often will a 50,000-tonne vessel need to be loaded to export your weekly production rate?) or availability of financing. Even though most projects are limited by such considerations, not all of them are. Clients almost always start with the question: "How big can you go?". In this article **Simon Early** of Johnson Matthey, considers the answer to that question.

A methanol plant consists of several sections. For the purposes of this article we will look at four major blocks: syngas generation, syngas compression, methanol synthesis loop and product distillation.

There are already ten methanol plants of around 5,000 t/d or larger¹ in operation, and the world's first 7,000 t/d plant is due to start up later this year². These large plants typically feature parallel units for heat exchangers, and layout considerations for parallel streams are important to consider when determining the cost of a plant. While some items of equipment do benefit from the economies of scale, an ever-increasing proportion of the equipment is sufficiently large that the vendor costs are linear with size, and the cost savings are instead

associated with the on-site construction costs. Many projects are expected to make extensive use of modularisation, where parallel modules are an attractive option.

A brief history of size

In the early years of syngas (1960s and 1970s) plants would double in size every few years. First there were the 300 t/d plants, then 600 t/d, and then 1,200 t/d. In the late 1970s the first 2,000 t/d plants were designed, starting up in the early 1980s. After that, things were steady for 20 years, slowly edging up to 3,000 t/d.

In 2001 two projects at 5,000 t/d were announced. The first to be commissioned, in 2004, was the 5,000 t/d Atlas plant in Trinidad^{3,13}, followed in 2005 by the 5,400

t/d M5000 plant, also in Trinidad¹³. These provided proof-of-concept for the new scale of plants – both for combined reforming (Atlas) and for conventional steam methane reforming (M5000).

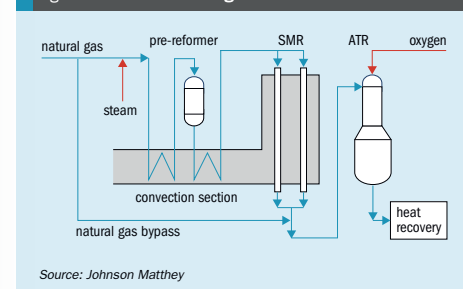
The biggest plants stayed around this capacity for a decade. The 7,000 t/d Kaveh plant in Iran is due to start-up towards the end of 2017. Industry observers and clients are again asking "how big can you go?". The answer is "probably larger than you first thought".

Syngas generation

Coal-based plants

Coal-based plants already use multiple streams, with each gasifier able to produce the syngas needed for 1,000 to 1,200 t/d

Fig. 3: Combined reforming



Source: Johnson Matthey

of methanol production, with one or two gasifiers as installed spares. While gasifier sizes will slowly increase, coal-based methanol plants are going to continue to use multiple parallel gasifiers.

Steam methane reforming (SMR)

The largest SMR-based plant is the M5000 project in Trinidad, with 864 tubes. The SMR at M5000 is based on a modular design that can, in theory, be extended further. The conservatism of the methanol industry has rightly been characterised as "the rush to be second", and no clients have decided to proceed with plants based on an SMR larger than 864 tubes. The largest of the current crop of SMR-based plants in design are based on the same 864 tube design used for M5000.

The flowsheet sketch in Fig. 1, shows a pre-reformer upstream of the SMR. The duty of the pre-heat coil (between the pre-reformer and the SMR) recovers heat from the convection section and therefore reduces the duty of the main SMR radiant box, allowing more syngas to be produced from each tube.

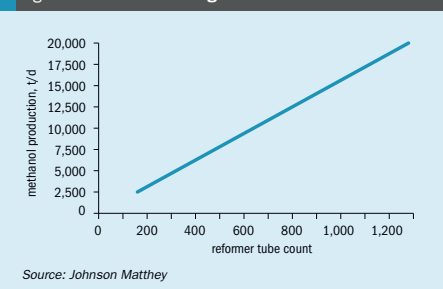
The syngas production from any reforming process, including an SMR, can be conveniently expressed as the Nm^3/h of " H_2+CO ". The approximate production is:

- 864 tube SMR with pre-reformer: 530,000 Nm^3/h " H_2+CO "
- 864 tube SMR: 450,000 Nm^3/h " H_2+CO "

The theoretical limit for methanol that can be made from syngas is $(\text{H}_2+\text{CO})/3$, and relies on adding CO_2 to increase the stoichiometric ratio, $R = (\text{H}_2-\text{CO}_2) / (\text{CO}+\text{CO}_2)$, to 2.0.

For syngas with $R=2$ or more the equations simplify to: Theoretical limit for methanol $= (\text{H}_2+\text{CO}) / (R+1)$

Fig. 4: Combined reforming case



Source: Johnson Matthey

Practical limits on loop conversion efficiency, even with hydrogen recovery and recycle, and distillation losses, stop you reaching the theoretical limit of 6,060 t/d for 530,000 Nm^3/h of $\text{H}_2 + \text{CO}$, but with a bit of ingenuity we can get pretty close to the theoretical limit in a modern flowsheet.

In Fig. 2 the lower section of the line is for a typical SMR plant where CO_2 is added to reduce the R value down to 2.15. The impact of CO_2 is the same for CO_2 added to the SMR feed or for CO_2 added direct to the syngas compressor or to the loop (or any mixture of all three options). Adding CO_2 to the SMR feed produces more process condensate and less water in the crude methanol than adding CO_2 downstream of the SMR. The upper section of the line shows the increase in production associated with adding even more CO_2 , to get to an R value of 2.0, which will require the recovery of hydrogen from the synthesis loop purge.

The new generation of foil-based catalyst⁴, such as CATACEL[®]SSR[™] will increase the production of $\text{H}_2 + \text{CO}$ per tube by at least 10%.

SMR + ATR (combined reforming, CR)

Around 30 to 50% of the natural gas feedstock is bypassed around the SMR to react with oxygen in the autothermal reformer (ATR). The ATR increases the total H_2+CO production, but also has a lower R value, typically around 2.15, so there is little further to be gained by adding CO_2 . By increasing the SMR bypass and increasing the oxygen flow to the ATR you can get to an R value of 2.0, which requires the recovery of hydrogen from the synthesis purge gas to keep the stoichiometry in the right region for the converter catalyst. Fig. 3 shows a flowsheet for combined reforming.

The CR plants currently in operation have much smaller reformers than 864 tubes, and usually do not include a pre-reformer. If we consider our theoretical limit to be (multiple) ATRs downstream of an 864-tube SMR with a pre-reformer then we could make a lot of H_2+CO ... enough for 13,500 t/d of methanol.

How many ATRs you need depends on the design of the ATR. Currently the largest ATR is on the 7,000 t/d Kaveh plant in Iran^{2,5}.

Fig. 4 shows the methanol production possible when running at $R=2.0$ exit the ATR. Smaller plants are more likely to operate at around $R=2.15$ to avoid the need for hydrogen recovery from the loop purge gas, which needs a slightly larger SMR than shown.

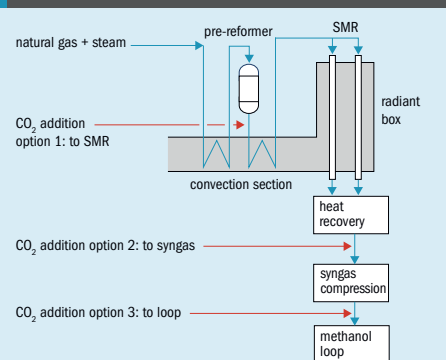
Advanced combined reforming (ACR)

The ACR flowsheet (Fig. 5) uses the gas heater reformer (GHR) and ATR for syngas generation. The smaller footprint of the GHR (compared to an SMR) allows it to be shipped ready-made to site. Johnson Matthey's ACR flowsheet won the 2016 IChemE Global Award for Outstanding Achievement in Chemical and Process Engineering⁶ because it allows a huge reduction in CO_2 emissions, compared to conventional reforming technologies. In regions where renewable electricity is plentiful, the ACR would use parallel streams of GHR+ATR, and so can be scaled to any size plant and still be around 10% lower total installed cost than conventional CR⁷.

Parallel reforming (PR)

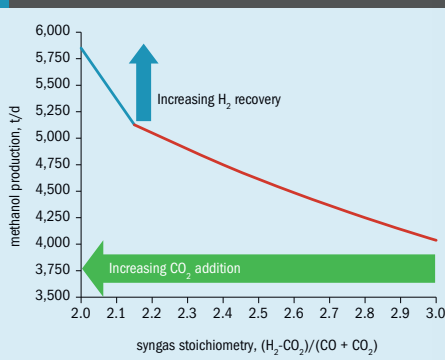
The concept of parallel reforming has been around for many decades – have an SMR to produce a hydrogen-rich syngas (around $R=2.95$) and combine it with a carbon-rich syngas from coal gasification, or POX or

Fig. 1: Pre-reformer upstream of SMR



Source: Johnson Matthey

Fig. 2: SMR case



Source: Johnson Matthey

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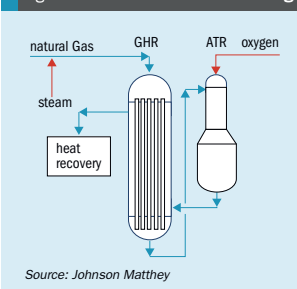
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Fig. 5: Advanced combined reforming

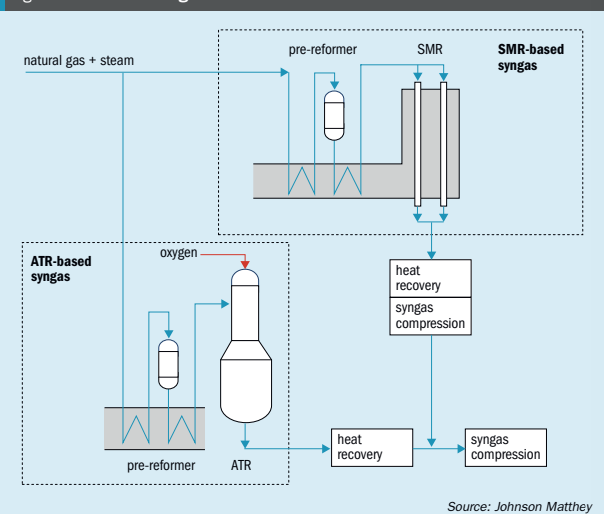


an ATR ($R=1.6$ or less). The two sources of syngas are blended to give a mixture (around $R=2.15$), but if you include hydrogen recovery and recycle on the synthesis loop purge gas, you can push this to the theoretical production limit at $R=2.0$.

The PR concept is already in operation for two 5,500 t/d projects in China, where half of the syngas is provided by an SMR and half is provided by coal gasification. Unlike other coal-based plants, very little CO_2 needs to be removed from the coal-based syngas to adjust the stoichiometry, due to the surplus hydrogen in the syngas from the SMR. In this article, the focus is on using an ATR for the source of syngas with an R of less than 2.

The flowsheet sketch in Fig 6 shows a pre-reformer upstream of the SMR and a pre-reformer upstream of the ATR. The duty of the ATR pre-heat coil (between the pre-reformer and the ATR) may come from the SMR convection section, but is more likely to come from a separate fired heater. The ATR operates at higher pressure than

Fig. 6: Parallel reforming



the SMR, so the syngas compression is split into two stages, with the ATR syngas added only for the second stage of syngas compression.

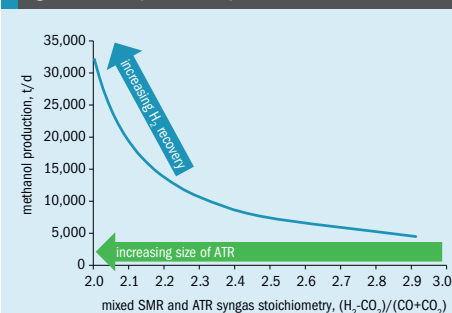
One of the advantages of PR is that hydrogen from the methanol loop, either recovered from the purge gas or using untreated purge gas, can be added to the ATR so that a pre-reformer stage is not required to avoid soot formation in the ATR. However, the maximum production is for an 864-tube SMR with pre-reformer, plus large ATR(s) with the highest possible R value, which is associated with

using a pre-reformer for each ATR to increase the inlet temperature, as shown in the flowsheet sketch. CO_2 addition reduces the maximum theoretical production in the parallel reforming flowsheet, so no CO_2 has been added when calculating Fig. 7.

The maximum production in Fig. 7 is at $R=2.00$ in the mixed syngas, when the H_2+CO from the ATR(s) is $4.5 \times$ the $530,000 \text{ Nm}^3/\text{h}$ H_2+CO from the SMR.

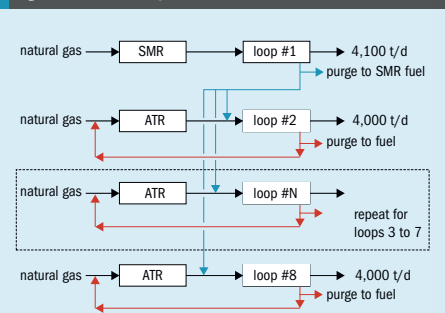
A single 32,000 t/d plant would probably cause major fluctuations in the spot market if it were off-line. Such marketing

Fig. 7: PR case (SMR + ATR)



Source: Johnson Matthey

Fig. 8: Parallel ATR plants



Source: Johnson Matthey

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H_2/CO
Ratio

Steam methane reforming (SMR)

- 130 references
- Ideal for hydrogen plants
- Proven reference for CO production
- Top fired – better scale up, compact and efficient units

Autothermal reforming (ATR)

- 30 references
- Ideal for methanol and FT
- Economy of scale, H_2/CO flexibility as stand alone or with combined reforming

Gas POX

- Ideal for oxo gas and CO production
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- Enabler of low emission configurations

Capacity



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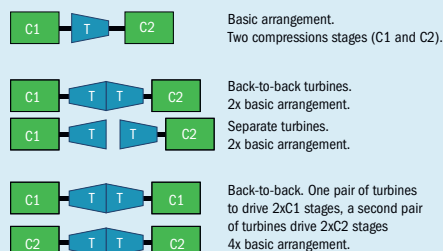
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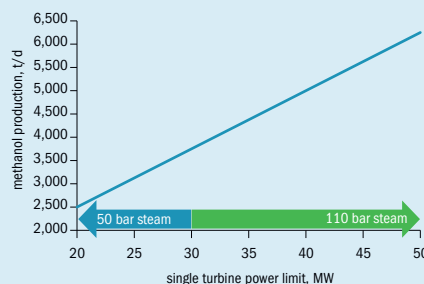
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Fig. 9: Syngas compressor and turbine drive arrangement



Source: Johnson Matthey

Fig. 10: Syngas turbine power limits (basic arrangement for SMR-based plants)



Source: Johnson Matthey

considerations are already apparent for the 5,000 t/d plants that sell product on the spot market (rather than providing feed to an adjacent MTO plant), and it is unlikely that any client will want to take the PR concept to the limit of 32,000 t/d.

Parallel ATR plants (PA)

First you start with an SMR-based plant, with no CO₂ addition. This has a lot of left-over hydrogen in the purge gas which can be combined with the syngas produced in an ATR. Each ATR-based syngas generation will look like the equivalent section of the parallel ATR flowsheet, discussed above.

This arrangement allows multiple identical ATR-based plants, each with their own methanol synthesis loop, to be added one-by-one. When the potential reduction in repeat engineering, construction cost, and schedule for each phase was studied to see the impact of the offset in start-of-construction dates for each phase, this concept showed noticeably better overall project economics than "super sizing" conventional technology. The later stages of the project will be able to call on financing backed-up by the already-operating earlier phases.

In Fig. 8, purge gas from loop #1 is initially used as fuel for the SMR. As later ATR-based plants are added, about 15% of the purge gas is diverted to each new plant to supplement the ATR syngas and increase the R value of the mixed syngas to just above R=2. Eventually, after seven ATR-based plants have been built, all purge gas from loop #1 is used in the ATR-based plants, and the purge gas from each of the ATR-based plants is used as fuel, either in the SMR or for the fired heater on the ATR-based plant.

In a paper by Charles Yiu⁹, the production for the ATR-based plant was 2.75 x the production of the base SMR plant. That was seen as big enough (about 16,000 t/d from a single 864-tube SMR), but by recycling the majority of the purge gas from the loops on the ATR-based plants you can go even further. By sharing the purge gas among the ATR-based plants, and a bit of ingenuity, you can keep the ATR-based plants on-line at only slightly reduced production when the SMR plant is shutdown.

For example, a single 864-tube SMR forms the basis for a 32,100 t/d complex of eight plants, which is about the practical maximum. 32,100 t/d is approximately the same peak production as the parallel reforming option (which is what we would expect, because both are determined by the same underlying H₂+CO mass balance) but in practice the number of ATR-based plants is likely to be less than seven, depending on the operating conditions of the ATR, use of pre-reformers, hydrogen recovery and so on. But, when answering the question "How big can you go?", the same assumptions have been used that are behind the PR case.

Syngas compression

The limits on turbine power for the high-speed/variable-speed drive of a syngas compressor has long been one of the key limits to assess on an SMR-based plant. For an SMR plant the syngas needs two stages of compression, C1 and C2, driven by a single steam turbine, T, as shown in the basic arrangement in Fig. 9. For the CR and ACR flowsheet only a single compression stage

is needed, so the compression limits are roughly twice those of an SMR plant. The pressure of coal-based syngas is broadly similar to combined reforming, but most coal-based projects raise steam at 50 bar or less, rather than at 110 bar, and so the plant size limits (per turbine) are similar to those of an SMR. High pressure coal-gasification technology, as used on the 6,000 t/d plant for Pu Cheng Clean Energy Chemical Company Ltd (PCEC), does not require any syngas compression at all¹⁴.

Fig. 10 provides only a broad indication, as the specific breakpoint between 50 bar steam and 110 bar steam will vary slightly depending on the chosen vendor. There is also an impact of using vacuum exhaust from the turbine (the default choice until 2015) and the increasingly popular option in recent years of using LP exhaust from the big syngas turbine, on projects where LP steam is needed for the distillation reboilers and the deaeration of demineralised water.

The first upgrade, using back-to-back turbines or separate turbines for stages C1 and C2, is sufficient to get to around 10,000 t/d for an SMR or coal-based flowsheet, and to about 20,000 t/d for combined reforming flowsheets.

Methanol synthesis

The capacity of the synthesis loop depends chiefly on the size of the methanol converter. There are several popular choices for the methanol converter, but the radial-flow steam-raising converter (R-SRC) is one example of a converter that is ideal for a very large methanol plant. There are over 20 R-SRC in operation around the world, most of them in China. The low-pressure



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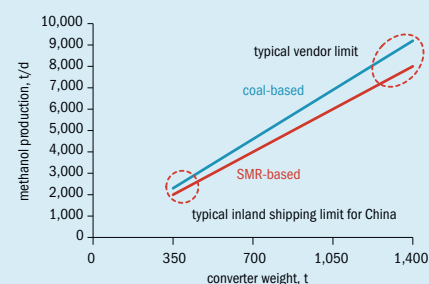
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Fig. 11: Radial-flow steam raising converter sizes



Source: Johnson Matthey

drop associated with radial flow (typically less than 0.5 bar per converter, nozzle-to-nozzle) means that there is no longer a limit on the tan-tan dimension of the R-SRC. Earlier designs, in particular the tube-cooled converter (TCC) and quench-cooled converter (QCC), were limited to about 5 m bed depth, but designs like the R-SRC completely remove this limit.

Fig. 11 shows the methanol production possible per converter.

For plants in inland China, which is most of the coal-based plants, the shipping weight limit is around 350 to 400 tonnes. This is why all of the 5,500 t/d plants designed by Johnson Matthey have two R-SRC installed in a series loop. By shipping the pressure shell and the internal bundle separately, and completing the closing weld on site, the maximum converter weight is increased for these plants to around 500 to 600 tonnes. But vendors can build them much bigger.

Vendor capabilities have been increasing steadily every year. In 2001, when the converters for the M5000 project were built, a typical crane limit at the vendor's workshop was 800 tonnes per converter. Now all the major vendors can lift converters weighing 1,200 to 1,500 tonnes, and the limits are slowly rising. Installing a converter weighing 1,500 tonnes at site is not a trivial exercise, as the cranes required for such a heavy lift cost hundreds of thousand dollars per day to hire in addition to mobilisation charges. The greater availability, and lower cost, of smaller cranes means that multiple converters are still likely to be chosen on projects that perform a detailed evaluation of their construction logistics and costs.

R-SRC is not the only design suitable for very large plants. The methanol converter for the 7,000 t/d Kaveh plant in Iran is designed by Casale SA and weighs 1,250 tonnes^{2,8}.

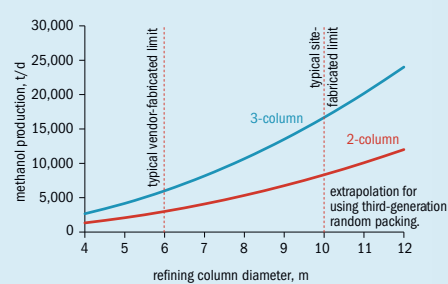
The size of the methanol converter is not the only factor to consider. Heat exchangers in a 5,000 t/d plant are often at the maximum size, with some split into multiple parallel shells. For high-efficiency loops, which are associated with high recycle ratios, the main loop piping is also at its practical limit for 6,000 t/d plants. Due to the existing use of parallel exchangers, the use of parallel piping does not have a significant impact on the construction cost, but it does highlight once again why traditional economies of scale apply to fewer and fewer parts of the plant as the size increases.

Product distillation

The plants in China mostly produce MTO-grade methanol for their adjacent MTO plant, and have relatively modest distillation column sizes. Plants that produce Federal Grade AA product, or the very similar IMPCA specification, require relatively large refining columns.

Since 2001 almost all methanol refining columns have used directed flow sieve trays, such as the UOP slotted sieve tray¹⁰ or fixed valve trays from other vendors¹¹. These trays use a small momentum transfer from the gas to the liquid to eliminate the hydraulic gradient that causes vapour cross-flow channelling (VCC) on large diameter trays. Now that the problem of VCC no longer limits the size of the refining column, more conventional fabrications limits come into play.

Fig. 12: Refining column sizes



Source: Johnson Matthey

A refining column built at a vendor workshop and shipped to site is limited to about 6 m in diameter. Site fabrication is significantly more expensive, but the maximum diameter is also greater. Site fabrication techniques are improving all the time, but keeping the column cylindrical becomes more and more difficult as the diameter increases, and if the column is not close to perfectly cylindrical then the trays are difficult to install. Currently, the realistic maximum refining column diameter for use with trays is about 10 m.

So-called "third generation" random packing, such as the Raschig Super-Ring¹², offers performance comparable with structured packing, with the installation convenience of random packing – which does not require the column to be a near-perfect cylinder. Again, the conservatism of "the rush to be second" in the methanol business means that no clients have decided to proceed with product distillation based on third generation random packing.

In Fig. 12 there are two lines. The upper line is for the so-called 3-column distillation design: a topping column followed by an HP refining column at about 8 bara pressure, which provides the reboil heat for the subsequent LP refining column, operating at about 1.5 bara pressure. The lower line is for 2-column distillation: a topping column followed by an LP refining column at about 1.5 bara pressure. Generally, 3-column distillation is used with combined reforming flowsheets, while 2-column distillation is typically used with SMR flowsheets.

The 5,000 t/d Atlas plant uses combined reforming and 3-column distillation,

with a single topping column, an HP refining column and an LP refining column.

The 5,400 t/d M5000 plant uses SMR and 2-column distillation, with a single topping column and two LP refining columns in parallel.

The benefits of 3-column distillation is not in the total number (and cost) of columns, but in the lower energy consumption (for the reboilers) and the associated lower reflux condenser duty for the LP refining column.

Conclusions

10,000 t/d of methanol is not a bad estimate for the maximum size for the next generation of methanol plants. At 10,000 t/d everything is a reasonable extension of what has been done before.

Above 10,000 t/d the plant is not really single stream, as almost every major item of equipment will be multiple parallel units. As plants get bigger, traditional economies of scale apply to fewer and fewer sections of the plant, and a single big plant has the same installed cost as two plants each of half the size. In this

situation, the author expects that building multiple identical plants will be the most attractive option:

- construction can be phased;
- phase 2+ will have a shorter schedule and lower construction cost than phase 1;
- phase timing will suit vendor fabrication times;
- re-use engineering;
- project financing for later phases will be based on income from phase 1+.

The first project to use multiple identical plants will be watched with great interest. If the cost savings and project financing benefits are as great as the author expects, then we will probably see the maximum size of each train being reduced. If that does come to pass, then future plants will be multiple trains of 3,000 to 4,000 t/d, with only a few plants at 5,000 t/d or larger.

The big plants are coming, but there are likely to be only a few 7,000 to 10,000 t/d plants before the superior economics of carefully-phased multi-stream plants are demonstrated and then start to dominate the new plant market.

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Methanol plants reach 7,000+ tonnes per day

The world's largest methanol plant currently stands at 7,000 t/d. In this article we report on the latest methanol technology and design concepts for large scale methanol plants from Casale, thyssenkrupp Industrial Solutions and Topsoe.

Casale big methanol plants

Casale designed the world's largest methanol plant with a single train capacity of 7,000 t/d. Fig. 1 is an aerial view of the construction site of that plant, which is presently under commissioning and will be on stream in 2017. The size of all critical items (steam reformer, ASU, machinery, reactors and columns) in Casale's design is such that they are commercially available and can be supplied for a capacity of 10,000 t/d.

In addition, Casale design has also been selected for three 5,000 t/d methanol plants currently under construction.

The process design of Casale methanol plants is based on combined steam reforming plus ATR technology for the production of the synthesis gas, proprietary IMC technology for the synthesis section and a three column design or a proprietary four column design for the distillation section (Fig. 2).

Process description

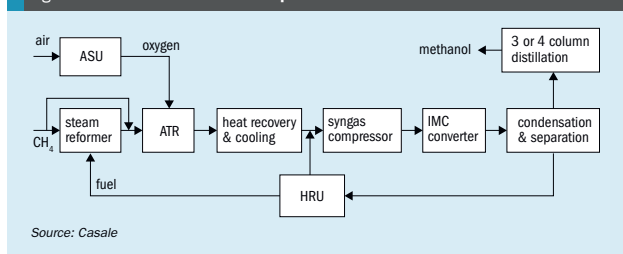
Natural gas feed treatment and saturation
Natural gas is pre-heated and fed to the hydrodesulphurisation reactor, where all sulphur is removed via the hydrogenation of organic compounds containing S to H₂S and the consequent absorption of H₂S on a ZnO bed.

A saturation unit is included to optimise the low temperature heat recovery, which is one of the main goals to be considered in the design of a methanol plant. All the process condensates separated in the front end section are mixed and fed to the saturator; the water exiting the bottom of the column is recycled to the top of the saturator after being heated by the steam produced in the synthesis reactor plates. The addition of the saturator provides the following benefits:



Fig. 1: Construction of world's largest methanol plant, 7,000 t/d with Casale technology.

Fig. 2: Casale advanced methanol process



Source: Casale

- a reduction of the overall energy consumption
- a reduction of the liquid effluents and their pollutants; the main portion of the process condensate is stripped by natural gas and is largely recovered as a reactant for the reforming reaction; the remaining portion of process condensate is stripped by MP steam making it suitable for its re-use in the steam system after treatment in the resins unit.

Reforming

The reforming section comprises the steam reformer, the pre-reformer and the autothermal reformer. Before being fed to the pre-reformer, the saturated natural gas is mixed with medium pressure (MP) steam in order to reach the desired steam to carbon (S/C) ratio and is pre heated in the convection section of the primary reformer.

The heat content of the flue gases exiting the radiant box is recovered in the

convection coils and in the combustion air pre-heater. The flue gas duct heat recovery has been designed using pinch technology for maximum efficiency and minimum capital cost. Thanks to this advanced design the flue gases are released at the stack at about 150°C with a very low content of nitrogen and sulphur oxides.

The reformed gas leaves the primary reformer and is fed to the autothermal reformer (ATR).

The oxygen required for the autothermal reforming reaction is provided by means of air separation units (ASUs). The oxygen coming from the ASUs is heated by means of steam up to 225°C and then sent to the ATR burners. The burners of the autothermal reformer are a special Casale design allowing for low pressure drop and smooth control of carbon formation.

The ATR outlet stream is sent to the high pressure steam boiler, process superheater and then to the BFW heaters, successively the low temperature heat is recovered by the reboilers of the topping and refining columns and in the demi-water pre-heating. No exchanger bypasses are planned in this section for very easy and stable operation; any variation in the heat recovery due to fouling of the units has a negligible consequence on the operation of the plant and can be easily managed by the control of the steam system. The final cooling of the process gas is performed by a trim water cooler.

The synthesis gas is compressed in a single body syngas compressor; the discharged gas is mixed with the recycle gas and finally discharged into the synthesis loop.

Synthesis section

The core of the plant is the synthesis section. The two main parameters to be considered in the design of the synthesis loop are the conversion efficiency and the heat recovery which need to be optimised versus the capital cost. Conversion efficiency is measured by the so-called carbon efficiency, the amount of carbon oxides converted into the crude methanol, while the heat recovery is measured by the amount of reaction heat recovered per tonne of product methanol.

Casale draws upon its long experience of optimising the synthesis section of various existing plants with different original configurations to offer a unique design where the carbon efficiency and the specific reaction heat recovery are maximised.



Fig. 3: Methanol converter installation in 7,000 t/d plant.

The adoption of the Casale patented IMC technology, already successfully applied in various plants, is a key feature of the synthesis section (Fig. 3). The IMC reactor comprises two axial radial catalytic beds: in both beds the cooling plates are immersed to directly remove the heat of reaction by raising MP steam: the reaction heat is exploited directly by heating up the saturator circulating water and, consequently, increasing the steam content in the NG feed to the reformer.

In the steam raising bed the inlet temperature is controlled automatically by the bypass of the feed-effluent exchanger. The inlet temperature of the second bed is maintained by the bypass of the first bed on the syngas side. This temperature control system allows the heat removal to be adjusted and, accordingly, the temperature profile along the catalyst.

The crude methanol condensation is achieved by air cooling and final trim cooling with water; the final water cooler is designed with a suitable oversize to allow for occasional failure or underperformance of the air cooler and to meet the target of the maximum plant reliability and productivity.

Distillation

The crude methanol coming from the synthesis section is purified in the distillation section. Casale has designed and patented an efficient arrangement for this

section, adopting a topping column and three refining columns.

The crude methanol coming from the loop is fed to the topping column, where the lighter components (e.g. dimethyl ether and methyl formate) are separated overhead. The topping column vapours are cooled down through a cooling water condenser.

The bottom stream is pumped and sent to the first refining column, operating at 19 bar abs, where the methanol is distilled and separated on the top of the column. The high purity methanol produced (AA grade) is sent to storage.

The first refining column bottoms consisting of crude methanol is fed to the second refining column (operating at about 8 bar abs), where the methanol is distilled and separated on the top of the column. The high purity methanol produced (AA grade) is sent to storage.

The second refining column bottoms is fed to the third refining column (operating slightly above atmospheric pressure) where AA grade methanol is distilled from the top and water with a residual methanol content of a few ppm is separated from the bottom. This water is partially re-used in the crude methanol let-down separator.

The heavier components (especially heavier alcohols) are separated as a side stream in the middle of the refining column and recycled back to the front-end where they are used as fuel. Casale design of the refining column features alternative trays for the fusel oil purge, in order to guarantee a maximum operating flexibility.

The required heat for the topping and refining columns is supplied by the process.

Thanks to its innovative design, the Casale advanced methanol process from natural gas (M7000) has a total energy consumption of less than 7 Gcal/t of refined methanol including the ASU. The efficiency of the plant is also reflected in the minimum consumption and size of the utilities plant.

Because of their similarity, the process schemes of Casale's M7000 methanol plant and A6000 ammonia plant can be easily combined to form an integrated plant producing both chemicals. In this case the plant features a syngas generation hub producing the feed for the two back ends. Other integrations are also possible, e.g., utilising the excess H₂ of the methanol plant in the ammonia plant or the excess CO₂ of the ammonia plant in the methanol plant.

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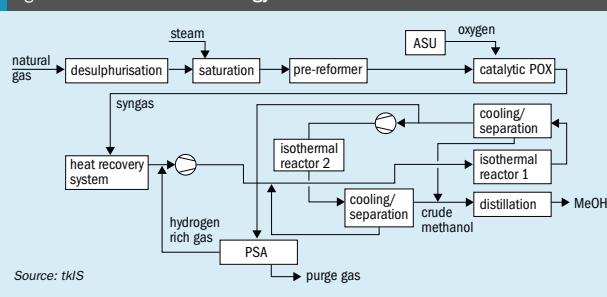
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Table 1: AdWinMethanol® improvements

	Conventional methanol process	AdWinMethanol®	Improvements
Reforming	combined reforming (steam reformer + ATR)	catalytic partial oxidation (ATR)	Steam reformer not required • lower energy consumption • smaller ATR reactor due to higher pressure
Syngas compressor	high ΔP	low ΔP	• significantly smaller compressor • more reliable operation
Reaction section	2-stage reactor system with 2 different reactor types	2-stage reactor system with one reactor concept and design	• simplified design (state-of-the-art isothermal reactor for both stages)
MeOH separation	no separation between the 2 reaction stages	methanol separation between the 2 reactor stages	• higher COx conversion • lower recycle flow rate • lower catalyst volume in 2nd stage
Recycle compressor	Location after second reactor stage	Location after separation, but before 2nd reactor stage	Due to higher pressure, 2nd reactor stage becomes smaller and more efficient Recycle ratio reduced • smaller recycle compressor

Source: tk/S

Fig. 4: AdWinMethanol® technology



Source: tk/S

thyssenkrupp's AdWinMethanol® technology

thyssenkrupp has extensive experience and a large number of references for the design and implementation of gasification plants and plants for the conversion of syngas into valuable products. Based on this comprehensive knowledge and expertise, thyssenkrupp has advanced its own methanol technology, AdWinMethanol® – an advanced integrated methanol technology.

AdWinMethanol®, developed together with GasConTec, a technology oriented consulting and engineering company, is a further development of conventional methanol technologies available on the market. The process is based on proven equipment (sizes comparable to those of 5,000 t/d combined reforming plants which are in operation today), but utilises an advanced process arrangement (see Fig 4), especially in the methanol synthesis section, and

optimised process parameters (Table 1).

AdWinMethanol® technology is suitable for large scale methanol plants with a capacity starting at 3,000 t/d up to 10,000 t/d in a single train. The basic concept is to process conventional and/or non-conventional short-chain hydrocarbon feedstock into syngas via catalytic partial oxidation (ATR) with a high flexibility of the feed stock. Conversion of the produced syngas to methanol takes place in two sequential isothermal reactor stages and is followed by conventional methanol distillation.

AdWinMethanol® technology stands out from other methanol technologies by its reduced synthesis gas volumes, significantly simplified compressors and elimination of the steam reformer.

The main advantages of AdWinMethanol® technology for the owner/operator are:

- more reliable;
- less EPC cost, also resulting in less total installed cost (TIC);

- high feedstock flexibility;
- reduced natural gas consumption by approximately 3%;
- significantly reduced catalyst cost;
- reduced maintenance cost.

Topsoe ATR syngas technology

Topsoe has invested significant efforts in developing specialised technologies for syngas generation and methanol synthesis, including the implementation of stand-alone autothermal reforming (ATR) for syngas generation. With its low steam-to-carbon ratio, the unique stand-alone ATR technology lowers both the initial overall methanol plant cost and the running operating cost significantly.

Stand-alone ATR is advantageous because it does not require the supply or dissipation of thermal energy to or from the reaction. This feature makes ATR useful for many applications, and it has been proven industrially to provide quality syngas for large-scale plants. A major feature of Topsoe ATR reforming technology is the ability to directly produce syngas with an H₂/CO ratio of approximately 2.0. A lower steam requirement implies that the syngas unit can be reduced in size, thus saving on capital expenditures (capex). Low S/C also enables larger single-train capacity. Furthermore, the water-gas shift equilibrium dictates that under low S/C conditions, the syngas is characterised by low H₂/CO ratios and high CO/CO₂ ratios. An advantage of high CO/CO₂ ratios is reduction in size of the methanol synthesis section.



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Fig. 5: Topsoe ATR reforming process for methanol production

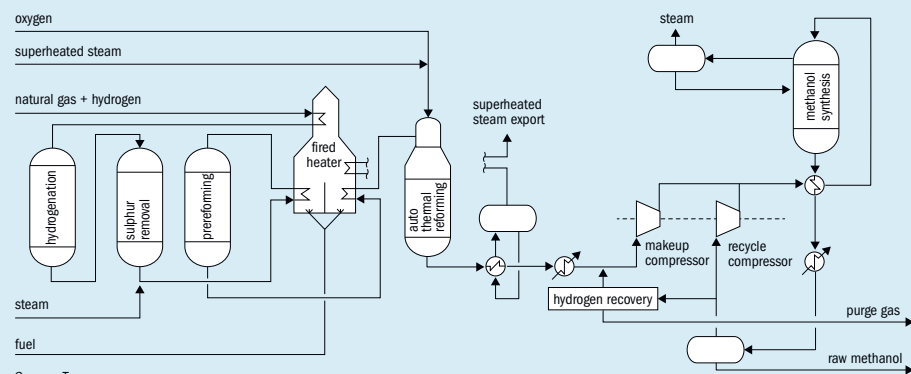
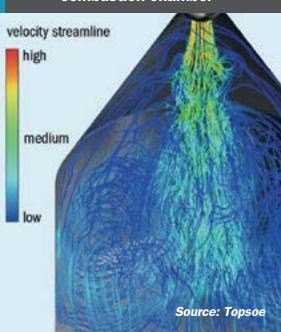


Fig. 6: Topsoe ATR reformer



Fig. 7: Visualisation of mixing in the combustion chamber



conditions such as lower S/C ratios at increased capacities.

In recent years, Topsoe has introduced a new generation of CTS burner for operation in high severity ATR reformers. A main feature of the new generation CTS burner is the flow-optimised burner nozzles (Fig. 7), the use of alloys with high temperature stability, and the use of an enhanced nozzle cooling. These features make the Topsoe CTS burner suitable for very large single-line capacities for Topsoe ATR operation at S/C ratios of 0.6 and below, providing high reliability and long lifetimes. The new generation CTS burner has been in operation for more than six years in large-scale industrial ATR reformers converting in excess of 180,000 Nm³ natural gas per hour.

New catalyst bed support

The current trend of building increasingly larger ATR reactors, necessitates a catalyst bed support with increased mechanical strength. For decades Topsoe has successfully used the pyramid type bed as catalyst support. Some degree of maintenance is generally reported for individual pieces of the ceramic elements.

A new Topsoe catalyst bed support has therefore been developed using finite element analysis in order to improve reliability and reduce maintenance. Fundamental examination and a thorough understanding of the forces acting on the catalyst bed support and the individual brick shapes has facilitated the new design taking into account the weight of catalyst above, fluid dynamic pressure drop and thermal loads.

vidual elements is critical for overall good performance of the ATR.

CTS burner and combustion chamber

Topsoe continuously improves and develops refractory lining solutions based on material science and exploration of novel and sophisticated refractory materials which improve the refractory design solution.

The ATR reformer burner is a crucial piece of equipment. Topsoe CTS (Cool Tip Swirl) burners have been in operation in ATR reformers for decades with high reliability and outstanding safety records. The industrial Topsoe ATR reformers have accumulated more than 70 years of safe operation at a low S/C ratio of 0.6. The design is undergoing continuous improvement in order to meet a wider range of

Together with the right selection of process conditions and catalysts, ATR technology can offer methanol producers high availability factors and reliability of operation.

Topsoe ATR syngas technology is proven to be reliable and efficient for large-scale GTL plants and can be applied in a similar layout for large-scale methanol production (Fig. 5). ATR reformers in large-scale GTL operation currently cover an operating range equivalent to 7,000 t/d methanol.

The reactor design for the Topsoe ATR consists of a burner, a combustion chamber, target tiles, a fixed catalyst bed, a catalyst bed support, a refractory lining, and a reactor pressure shell as illustrated in Fig. 6. Good performance of these indi-

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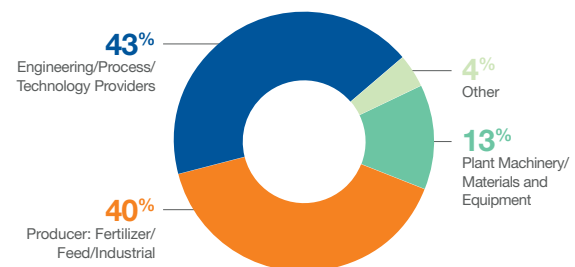


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Higher single train ammonia capacity

The trend to higher ammonia plant capacities continues. In this article, thyssenkrupp presents a concept for a 4,700 t/d single train ammonia plant based on proven technology and Casale presents its concept for a single train ammonia process suitable for capacities of up to 7,000 t/d.

One of the main drivers in the ammonia industry has been the financial benefit associated with the economy of scale offered by larger plants. Improvements in plant design have delivered this over time with some notable step changes in increased capacity. In the 1960s, the development of centrifugal compressor technology allowed a marked increase in capacity from a few hundred t/d to over 1,000 t/d by 1970.

There followed a period of steady capacity increases with plant capacities increasing to 2,000+ t/d by 2003.

The next major increase in capacity was achieved in 2006 when the first 3,300 t/d ammonia plant went on stream at Saudi Arabian Fertilizer Company (SAFCO IV) using the Uhde dual pressure ammonia process, available from thyssenkrupp Industrial Solutions (tkIS). This was a remarkable step increase in capacity as at that time the largest plants had a nameplate capacity of only 2,200 t/d (see Fig. 1).

tkIS has gone on to design and build three more plants of the same capacity:

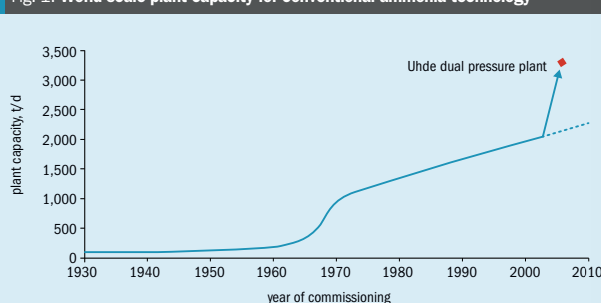
- Ma'aden 1, Saudi Arabia (start-up 2011)
- CF industries, USA (start-up 2016)
- Ma'aden 2, Saudi Arabia (start-up 2016).

All have surpassed their nameplate capacity of 3,300 t/d.

In addition, Ma'aden has another project for a new plant of this capacity underway and SAFCO IV is currently undergoing a revamp to increase capacity by 11%.

Other ammonia projects with capacities above 2,200 t/d include Kaltim-5 by PT Pupuk Kalimantan Timur, Indonesia with 2,500 t/d and EuroChem at Kingisepp, Russia with 2,700 t/d which are designed according to the KBR ammonia process, which is not a dual pressure process.

Fig. 1: World scale plant capacity for conventional ammonia technology



Source: tkIS

In the recent past there have also been projects where producers have demanded even higher capacities at a single site and built them in the form of two identical trains: Sorfert, Algeria (2 x 2,200 t/d, tkIS Uhde process) and QAFCO 5 and 6, Qatar (2 x 2,200 t/d, Topsoe process).

thyssenkrupp's ammonia plant scale-up

The two main considerations during the design of the first single train 3,300 t/d plant were to keep as much as possible from the existing referenced process where scale up is not critical and to develop a way to avoid the scale-up risk in those places where it is high.

Analysis revealed that the synthesis gas compression and the synthesis loop in particular are the areas with high scale-up risk. The solution was to insert a once-through ammonia synthesis upstream of the loop synthesis, operating at the outlet

pressure of the LP casing, then to separate the ammonia produced there and finally to send the remaining synthesis gas into the conventional synthesis loop (refer to Fig. 2). The desired effects are:

- The syngas compressor load is lowered and the same model as in a 2,200 t/d plant can be used.
- Synthesis loop equipment and piping sizes are identical to that of a conventional 2,200 t/d plant, thus avoiding dimensions for which no references exist.

Both contributed in lowering the scale-up risk to an acceptable level.

Since ammonia production takes place at two pressure levels (once through and loop), it is also referred to as the dual pressure process.

Despite its success, the dual pressure process has a weak point – for the sake of risk-free scale up, it adds equipment to the standard process, mostly another ammonia converter and the cooling train associated

Fig. 2: thyssenkrupp's dual pressure ammonia synthesis

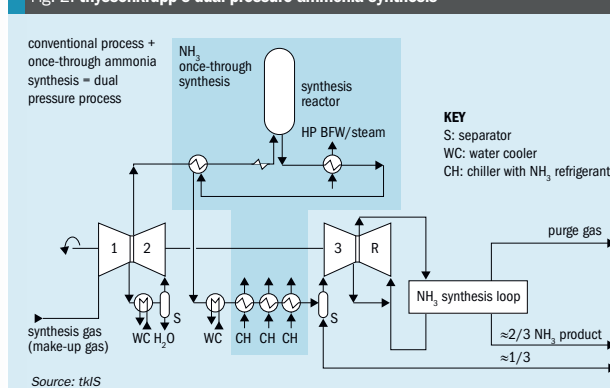
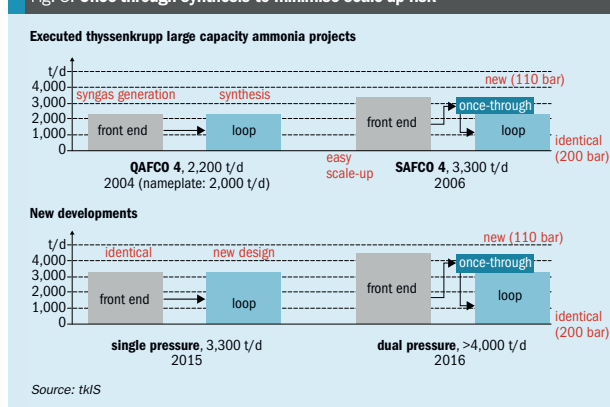


Fig. 3: Once-through synthesis to minimise scale-up risk



to it. Hence, it does not fully exploit the potential of the economy of scale. Further savings are possible by deleting the additional equipment and integrating its function into that of the standard process.

Therefore, thyssenkrupp has launched two R&D projects on this subject:

- Explore the cost saving potential by building a 3,300 t/d ammonia plant using the conventional process, however keeping the scale-up risk at an acceptable level.
- When a concept for a conventional 3,300 t/d plant is available, it is the logical consequence to investigate the possibilities to use its synthesis gas compression and synthesis loop and check to which capacities it can be expanded by adding a once-through synthesis.

These two developments are schematically shown in Fig. 3, bottom row.

Cost optimisation of 3,300 t/d plants

The cost optimisation of the 3,300 t/d plant mostly consists of deleting the once-through synthesis and increasing the capacity of the loop synthesis and the syngas compressor. Attention is focused on addressing measures to minimise shaft power of the syngas compressor, keeping it small in size to make use of existing reference installations:

- synthesis loop with high conversion and low circulation rate;
- suction chilling to 4°C at all stages;
- synthesis gas drying unit;
- cryogenic hydrogen recovery unit.

Ammonia synthesis

The biggest changes are located in the ammonia synthesis loop which has to be increased by approximately 50% compared to the reference plants.

The ammonia synthesis loop of a 2,200 t/d plant as well as that of a dual pressure 3,300 t/d plant has a pipe size of 20". For the conventional concept, this is being increased to 24", the next available standard size. Keeping the velocities and other parameters unchanged would allow for 3,170 t/d with this pipe size. More production is achieved by higher conversion per pass and by lower circulation rate.

Several configurations of ammonia converters and waste heat boilers have been investigated. The one found to be most suitable was to distribute the catalyst volume into four beds in two converters. Both ammonia converters have two catalyst beds and an internal heat exchanger for temperature control of the bed inlet temperatures. The synthesis loop of the dual pressure plant is already equipped with two ammonia converters (three beds in total) – that means no equipment is added, it is only designed in a different way. High catalyst volume and highly active catalyst KATALCO_{JM} 74-1 from Johnson Matthey is provided. This additionally promoted high activity catalyst has originally been developed for low-pressure applications, but it is well referenced and provides a performance enhancement at usual synthesis loop pressures of around 200 bar. The second converter houses more catalyst than the first one. The diameters of both pressure vessels are intentionally kept within a self-imposed limit of 3.2 m, which is the size of the largest converters designed by thyssenkrupp so far (see Table 7). Fabrication of larger diameter vessels is of course possible, but this procedure ensures there will be no fabrication and transport issues.

As in the reference plants, an HP steam generator is installed downstream of each ammonia converter. This ensures best recovery of the reaction enthalpy. The two waste heat boilers are similar in size and capacity.

The gas/gas heat exchanger downstream of the second waste heat boiler, preheating the converter inlet gas, can be built in two parallel units if highest energy recovery is desired. The remaining synthesis loop equipment is less critical and no major risk is expected with its size increase.

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Table 1: Syngas compressor data for different plant sizes

Plant capacity (t/d)	Shaft power in normal operation (kW)	Normal speed (rpm)
2,200 conventional, as built	25,864	9,575
3,300 dual pressure, as built	28,600	9,674
3,300 conventional	33,200	9,701
4,700 dual pressure	38,078	8,859

Source: tkfS

Syngas compression

For economic reasons, the general configuration of the syngas compression train, consisting of one turbine, two compressor casings, and no gearbox, shall be maintained. This means that the concepts with separate turbines for LP and HP compressor casing or with three compressor casings are not taken into consideration. It is essential that there is a suitable operating reference for the syngas compressor – anything else would not be welcomed by the market.

Changing a dual pressure plant (Fig. 2) to a single pressure (conventional) synthesis loop plant has the following implications on the synthesis gas compressor:

- shaft power increases because the full amount of synthesis gas has to be compressed up to the outlet of the third stage and the recycle gas flow rate increases;
- the higher flow rate present in the third compressor stage leads to higher distance between bearings, making the machine more susceptible to vibrations;
- the higher shaft power demand might require a larger turbine with lower speed. If the speed gets too low this has to be outbalanced by the compressor by either having larger impeller diameters or more impellers. In both cases this would lead to a different

compressor model, meaning that there might be no operating experience with it in an ammonia plant. This again would increase the risk involved, which is against the intention of the project.

The good news is that the same model of the syngas compressor of the existing 3,300 t/d plants can be used for the simplified plant. Data are compared in Table 1.

As seen from Table 2, changes in the LP casing against the reference plant are absolutely marginal and not critical. Impellers 5 to 8 even get slightly smaller (by 30 mm in diameter). The distance between bearings (bearing span) remains unchanged. The LP casing power in the new 3,300 t/d single-pressure plant is a little less than in the reference because the pressure profile is slightly different.

In the HP casing, the impeller sizes of the newly proposed plant are slightly larger (by 25 to 30 mm in diameter), except for the recycle wheel which remains unchanged. The main difference however is its bearing span, which, due to the higher flow rate passing through the HP casing, increases by 152 mm (plus 10%), but remains much smaller than in the LP casing. The approx. 50% increase in flow rate manifests itself in higher shaft power.

Other than in most of the conventional 2,200 t/d plants, a synthesis gas drying

unit has been added to the flowsheet. Such a unit is already part of the 3,300 t/d dual-pressure reference plants. (The plant can operate at lower rate in case the drying unit is not available.) The drying unit allows feeding the syngas to the synthesis loop directly upstream of the ammonia converter, leading to lower converter inlet ammonia concentration.

Other units

In most conventional plants, the hydrogen recovery is a membrane unit. While typically this is a cost-effective and reliable unit, it delivers both outlet streams at unfavourable pressures. The permeate stream, produced at low pressure, is the hydrogen-rich stream, to be returned to the synthesis loop. The retentate stream, produced at high pressure, is the methane-rich stream, to be sent to the low-pressure fuel system. In contrast, a cryogenic hydrogen recovery unit (as in service in the existing 3,300 t/d plants) returns both streams at suitable pressure levels. It avoids loading the first stages of the syngas compressor with the hydrogen-rich stream from it and therefore is selected for the new concept.

The plant front end remains mostly unchanged. That means there is no scale up versus the existing 3,300 t/d plants.

Overall economics

An economic comparison has been carried out with both plant versions designed to the same specific energy consumption of 6.74 Gcal/t (24.3 million Btu/st) ammonia. This accounts for combined consumption of natural gas (expressed by its LHV) and electricity, with credit for 52 bar (750 psi) steam export. The figure can be varied in both directions to reflect the economic conditions of a particular project.

The overall cost saving versus the dual pressure plant for engineering, supply and construction is estimated to be around 3.0% for a typical US Gulf Coast location.

Table 2: Syngas compressor details for built 3,300 t/d and newly proposed 3,300 t/d single pressure plant

Plant/capacity (t/d)	LP casing			HP casing			
	impeller 1-8 diameter (mm)	bearing span (mm)	shaft power (kW)	impeller 1-7 diameter (mm)	recycle impeller, diameter (mm)	bearing span (mm)	shaft power (kW)
3,300 dual pressure, as built	560	1,823	17,895	475 and 450	425	1,485	10,701
3,300 conventional	560 and 530	1,823	16,869	500	425	1,637	15,015

Source: tkfS

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Table 3: Primary reformer data

Plant capacity (t/d) and type	Reformer box volume (m³)	No. of reformer tubes
2,200 conventional, as built	3,025	288 5" tubes in 6 rows
3,300 dual pressure, as built	4,445	408 5" tubes in 8 rows
3,300, single pressure	4,445	408 5" tubes in 8 rows
4,700, dual pressure	7,088	570 5.12" tubes in 10 rows

Source: tkfS

Table 4: Secondary reformer data

Plant capacity (t/d) and type	Diameter (m)	Cylindrical height (m)
2,200 conventional, as built	4.5	17.4
3,300 dual pressure, as built	5.0	19.4
3,300, single pressure	5.0	19.4
4,700, dual pressure	5.5	20.5

Source: tkfS

Table 5: Dimensions of CO₂ removal equipment

Plant capacity (t/d) and type	Absorber internal Ø lower / upper part x cyl. height (m)	LP flash internal Ø (m), packing type
2,200 conventional, as built	5.0 / 3.3 x 42.6	6.1, PRM 50
3,300 dual pressure, as built	6.1 / 4.1 x 42.6	7.6, PRM 50
3,300, single pressure	6.1 / 4.1 x 42.6	7.6, PRM 50
4,700, dual pressure	7.1 / 4.6 x 42.6	7.9, IMTP 70 (8.6, PRM 50)

Source: tkfS

Application of dual pressure process

The next logical step towards higher ammonia capacities is to combine the 3,300 t/d synthesis loop with an upstream once-through synthesis. This can increase the capacity to beyond 4,000 t/d.

In this study, the target capacity is determined by the maximum amount of gas which can be handled by a synthesis gas compressor whose design data is close enough to referenced machines in successful operation. The capacity determined this way is 4,700 t/d.

Reforming

One way to unload the synthesis gas compressor is to operate the front end at higher pressure. In the concept presented here, the reformer outlet pressure is increased by 5 bar compared to the 3,300 t/d plant.

This has a couple of consequences, some favourable, some less favourable: The higher pressure favours the absorption in the CO₂ removal unit and helps to keep the absorber small. By the higher pressure, load is shifted from the syngas compressor to the process air and natural gas compressors, where the step to higher capacities is less risky due to existing references.

The primary reformer is still smaller than methanol plant reformers. Its dimensions are compared with the other presented plants in Table 3. The combustion air preheater at the end of the flue gas duct is split in two parts for reason of size.

The scale up of the secondary reformer is also not very critical. Table 4 shows characteristic data of several projects. Part of thyssenkrupp's standard design is the central riser pipe for the incoming reformed gas which acts as a support for the refractory arch in the bottom. Therefore, its span

is only half of the vessel diameter and half of that of competing designs.

The steam-to-carbon ratio has been increased compared to the reference plant to maintain the partial pressure of the reducing components (H₂ and CO) at the inlet to the HT shift reactor below a certain limit in order to avoid an adverse effect on the catalyst. The increase in steam flow has some impact on the process; however, its energy efficiency does not suffer from it. The higher throughput through the front end leads to higher HP steam production.

Process air compression

The proposed process air compressor is an integrally geared compressor, driven by a steam turbine (single-shaft machines are also available). References in this size exist from air compression units in other industries. Turbine-driven gear type compressors are also successfully employed in ammonia plants up to 2,200 t/d by thyssenkrupp and others.

Boiler feed water preheating

The boiler feed water heater upstream of the LT CO shift gets quite large. It can be built in one vessel but it seems to be cost-saving and beneficial for pressure-drop savings to install two parallel heat exchangers instead.

CO₂ removal

For this study BASF's OASE white® process (formerly known as aMDEA) has been selected for CO₂ removal because of thyssenkrupp's vast experience with this technology from its other large capacity plants.

Table 5 shows characteristic data of some of the CO₂ removal equipment. The LP flash vessel is the vessel with the largest diameter in the plant and is listed because it is most critical for transportation to site. For this reason, an effort was made to lower its diameter. Cost-effective packing material for the flash vessels in the reference plants as well as the new concepts is PRM 50. The only exception is the LP flash in the 4,700 t/d plant where IMTP 70 has been selected with licensors approval because it allows for a diameter reduction of 0.8 m.

Most of the recently built large ammonia plants are serving export-oriented fertilizer production and are located close to the sea, therefore, transportation of large vessels to site should not be critical.

Three motor-driven pumps with rated capacity of 60% of the total flow are foreseen for the semi-lean solution circulation,

Table 6: Syngas compressor details for the proposed 4,700 t/d plant

Plant/capacity (t/d)	LP casing		HP casing			Shaft power (kW)
	impeller 1-8 diameter (mm)	bearing span, (mm)	impeller 1-7 diameter (mm)	recycle impeller diameter (mm)	bearing span (mm)	
4,700 dual pressure	600	1,930	520	450	1,580	38,100

Source: tkfS

Table 7: Dimensions of ammonia synthesis equipment

Plant capacity (t/d) and type	Once-through converter: internal Ø x cyl. height (m)	Loop converters: internal Ø x cyl. height (m)
2,200 conventional, as built	–	3.0 x 11.3 2.9 x 10.2
3,300 dual pressure, as built	3.2 x 18.9	3.0 x 12.5 2.9 x 12.2
3,300, single pressure	–	3.2 x 11.6 3.2 x 19.4
4,700, dual pressure	3.2 x 18.9	–

Source: tkfS

two of them supported by expansion turbines installed in the rich solution flow from the absorber. This is the same concept as in the 3,300 t/d reference plant. All other pumps are installed as pairs with 2 x 100% capacity.

Syngas compression

For the equipment selection of the synthesis gas compression, the same considerations as for the 3,300 t/d plant are applied. Table 6 shows the main data. Despite the high capacity, the main dimensions are not much larger than those of the 3,300 t/d single-pressure plant: The bearing span of the LP casing is of course larger due to the higher volumetric flow rate processed by the impellers, but only by 5.9%. In contrast, the HP casing bearing span is even 3.5% smaller than that of the 3,300 t/d single-pressure plant (see Tables 2 and 6).

The reason for this is that the mass flow to the HP loop is approximately the same for both plants, but the dual pressure plant has a lower gas inlet temperature since the gas is coming from the first ammonia separator (see Fig. 2) and hence a lower volume flow. The fact that the dimensions are close to each other suggest that the rotor dynamics are not much different for both machines and can help to build trust that selecting the 4,700 t/d process does not involve high risk for the machine.

Ammonia synthesis

The ammonia synthesis loop is identical to the 3,300 t/d single pressure plant. The upstream once-through synthesis is equipped with a converter having the same

dimensions as the ones in service in the 3,300 t/d dual pressure plants. Due to its higher inlet flow rate it produces approximately 1,400 t/d ammonia. This concept of combining equipment already being used elsewhere ensures no new risk is created by the increase in total plant capacity. The converter dimensions are listed in Table 7.

Overall economics

There is a high MP steam export from the process plant of 136 t/h. Most of the large ammonia plants built in recent years are located within fertilizer complexes, where urea plants or other consumers for the export steam exist, so the steam export should be welcome and should lead to lower auxiliary boiler firing.

The amount of CO₂ produced is sufficient for production of approx. 7,500 t/d urea, about twice the size of many urea plants being built today.

For the purpose of the study, the specific energy consumption is kept at the same level as for the 3,300 t/d plant. Compared to a two-train plant, the concept presented has a clear cost advantage. Of course there are some elements in the cost calculation which do not have a cost regression with capacity, for example the number of heavy lifts during construction. But despite that, the savings by having less but larger equipment prevails as expected.

Study conclusions

The results of the detailed studies for improvement can be summarised as follows:

- By deleting the once-through synthesis and correspondingly increasing the size of the synthesis loop, the 3,300 t/d plant can now be offered with less equipment than the operating reference plants. This saves investment cost without adding risk.
- For capacities of >3,300 t/d, the once-through synthesis can be added to this type of plant, to increase the capacity up to 4,700 t/d.

The discussions with equipment vendors and own engineering resources confirmed there are no obstacles to both versions of the plant, but clear cost advantages against the benchmarks.

Casale A6000™ ammonia process

The philosophy followed by Casale when designing a new plant is to combine all of the latest and proven technological achievements to obtain the best compromise of investment costs, plant productivity and utilities consumption.

Casale has extensive experience of revamping projects, where the actual plant configurations and the desired targets differ case by case and require huge flexibility in design. The long activity in the revamping business has allowed Casale companies to develop a unique know-how and experience with practically all the most widespread technologies for syngas applications. This familiarity with very different designs permits an open-minded selection of the most suitable solutions to be adopted in each

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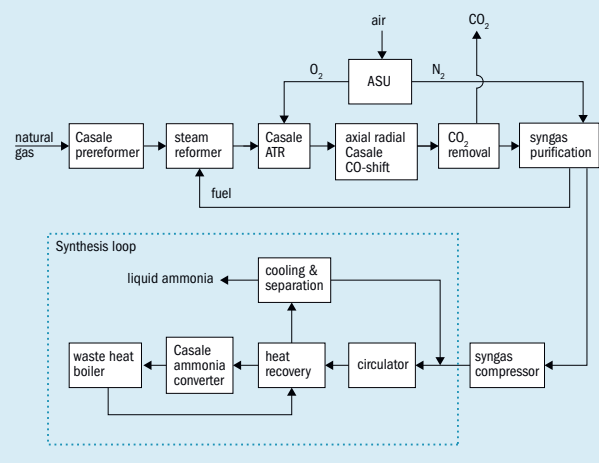
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Fig. 4: Casale A6000™ ammonia process



Source: Casale

specific case, without proposing a fixed standard configuration regardless of the different bases of design. In addition, Casale has developed a solid experience in the design and supply of new plants and has completed six grass-root ammonia plants, all 2,050 t/d capacity, with two on-stream and four in the engineering phase, as well as 53 ammonia synthesis loops, with a capacity of up to 2,000 t/d.

For larger scale single train ammonia plants Casale has developed Casale A6000™, a patented process suitable for capacities of >6,000 t/d of ammonia. It has been specifically developed to reach the highest single train ammonia capacity, thus taking maximum advantage of economies of scale, while still ensuring that all critical items are referenced and available from multiple vendors.

Based on several independent reviews, conventional ammonia process technology is expected to reach its maximum single train capacity at about 3,500-4,000 t/d of ammonia. Any significant further increase beyond that capacity would be restricted by limitations on the size of critical items such as the steam reformer, RG boiler, syngas turbo-compressor train, CO2 removal columns, synthesis reactor, synloop equipment, high pressure synloop and steam piping.

The A6000™ process either avoids or drastically debottlenecks all those critical items,

thereby permitting a much higher capacity. This is achieved by three major concepts:

- a nitrogen-free front-end based on a pre-reformer, steam reformer and oxygen-blown autothermal reformer (ATR);
- enhanced carbon dioxide removal, and purification based on Liquid nitrogen wash;
- inert-free ammonia synthesis loop with two Casale synthesis converters.

The syngas generation of the A6000™ process is based on the combination of a steam reformer with an oxygen-fired Casale autothermal reformer (ATR).

This line-up enables operating the reformer at a mild tube outlet temperature, and entails conditions beneficial for the downstream CO2 removal, syngas purification and compression.

Installation of a pre-reformer permits operation of the reforming at a steam to carbon ratio of 2.0, which has significant debottlenecking effects on the whole section. The relatively small duty of the reformer and overall plant efficiency are key factors to achieve a substantial reduction of the CO2 stack emissions, compared to a conventional process.

This front end configuration is typical of the Casale world scale methanol process. The front end of a A6000™ ammonia process is equivalent to a 5,000 t/d Casale methanol process.

The main units of the new Casale ammonia plant are shown in Fig. 4. The main process sections are:

- feed gas desulphurisation;
- air separation, O2 and N2 compression;
- prereforming (Casale proprietary);
- primary reforming;
- autothermal reforming (Casale proprietary);
- medium temperature shift conversion (Casale proprietary);
- syngas drying;
- syngas purification (nitrogen wash);
- syngas compression;
- ammonia synthesis (Casale proprietary converter);
- refrigeration;
- ammonia recovery.

The size of all critical items (steam reformer, ASU, machinery, reactors and columns) in Casale's design is such that they are commercially available and can be supplied for a capacity of 7,000 t/d.

Advantages of Casale design

The A6000™ provides major benefits for an ammonia producer; for a capacity of 7,000 t/d, compared to two parallel 3,500 t/d ammonia plants using conventional technology, it requires:

- 20% lower total investment cost;
- 30% less plot space;
- 45% lower total catalyst volume;
- about half the number of operators.

Additional benefits of A6000™ include:

- equivalent gas consumption lower than conventional ammonia plants (<6.7 Gcal/t);
 - 25% lower carbon dioxide stack emissions;
 - optional balanced NH3-CO2 for total conversion to urea.
- Other features are:
- minimisation of the plant capital cost;
 - enhancement of the plant reliability thanks to the reduced number of unit operations;
 - decoupling of frontend-backend operation thanks to the introduction of the liquid nitrogen wash;
 - minimisation of operating cost;
 - minimisation of land cost.

Reference

1. Noelker K and Meissner C (tKIS): "4,700 mtpd single-train ammonia plant based on proven technology", 2016 Ammonia Safety Symposium, Denver, USA.

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