Ammonia and the Fertiliser Industry: The Development of Ammonia at Billingham

A history of technological innovation from the early 20th century to the present day

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It is over 100 years since the Haber-Bosch process began in 1913 with the world’s first ammonia synthesis plant. It led to the first synthetic fixed nitrogen, of which today over 85% is used to make fertiliser responsible for feeding around 50% of the world’s human population. With a growing population and rising living standards worldwide, the need to obtain reliable, economic supplies of this vital plant nutrient for crop growth is as important as ever. This article details the historic background to the discovery and development of a process “of greater fundamental importance to the modern world than the airplane, nuclear energy, spaceflight or television” (1, 2). It covers the role of the Billingham, UK, site in developing the process up to the present day. The technology was pioneered in Germany and developed commercially by BASF. In 1998 ICI’s catalyst business, now Johnson Matthey, acquired BASF’s catalytic expertise in this application and now Johnson Matthey is a world-leading supplier of catalyst and technology for ammonia production globally.

1. Introduction

Ammonia is the second most produced industrial chemical worldwide. Of the four chemicals, ammonia, methanol, hydrogen and carbon monoxide that rely on similar syngas processes for their production, ammonia is the most complex requiring the highest number of catalytic steps.

Ammonia is one of the most important chemicals produced globally with approximately 85% being used as fertiliser for food production (3). The other 15% of ammonia production is used in diverse industrial applications including explosives and polymers production, as a refrigeration fluid and a reducing agent in nitrogen oxides (NOx) emissions control. Ammonia synthesis from atmospheric nitrogen was made possible in the first part of the 20th century by the development of the Haber-Bosch process. It remains the only chemical breakthrough recognised by two Nobel prizes for chemistry, awarded to Fritz Haber in 1918 (4) and to Carl Bosch in 1931 (5). The development of ammonia synthesis directly addressed “The Wheat Problem” as foretold by Sir William Crookes in 1898 (6) whereby a shortage of available reserves (of wheat) would only allow the world’s population to continue to expand to about two billion which would be reached around 1930. Thus, in the early 20th century, the need to increase food production led to the development of the fertiliser industry.

Today, the global value of ammonia production is estimated to be over US$100 billion, with the largest individual plants being capable of producing 3300 metric tonnes per day (mtpd) or 3640 short tonnes per day (stpd) (7). To achieve this scale many improvements have been made over the last 100 years in both process and catalyst technology.

After describing historical aspects of the original ammonia technology development by Haber, Bosch et al. in Germany, and the background to
the requirement for efficiency improvements, this paper uses perspectives from Billingham, UK, to describe some of the technological contributions that came from there in the development of ammonia production.

2. The Growing Need for Nitrogen

In just over 100 years the ammonia production industry has grown massively and continues to do so to feed the ever expanding world population. The development of the remarkable iron catalyst by Alwin Mittasch (8) and the technology for the synthesis of ammonia from nitrogen and hydrogen by Fritz Haber and Carl Bosch led to BASF starting to operate the world’s first ammonia synthesis plant in 1913. Researchers estimate that about half of today’s food supply is dependent on the nitrogen originating from ammonia-based fertilisers (9). Between now and 2050, while the world population will grow by 30%, the demand for agricultural goods will rise by 70% and demand for meat by 200% (10). This is linked with fundamental shifts in the demand curve for food, especially caused by population growth, rising affluence leading to changes in diet in many countries and in some regions increasing use of food crops to produce fuel. The environmental, human health and climatic aspects of ammonia and fertilisers in the growth scenarios have been reviewed elsewhere (11, 12).

Ammonia production technology has and continues to advance under the competitive challenges in the industry that demands an ever more energy efficient process, with lower emissions that can operate with high reliability for extended periods between shutdowns. There have been dramatic increases in environmental performance and energy efficiency over the last 100 years, but with modern steam reforming processes energy utilisation is nearing the theoretical minimum (13) (Figure 1) and looking forward, specific energy consumption can only be reduced marginally, if at all, for the most efficient modern plants.

Worldwide ammonia production is largely based on modifications of the Haber-Bosch process in which $\text{NH}_3$ is synthesised from a 3:1 volume mixture of $\text{H}_2$:$\text{N}_2$ at elevated temperature and pressure in the presence of an iron catalyst. All the nitrogen used is obtained from the air and the hydrogen may be obtained by one of the following processes:
- Steam reforming of natural gas or other light hydrocarbons (natural gas liquids, liquefied petroleum gas or naphtha)
- Partial oxidation of heavy fuel oil or coal.

In ammonia production technology the type of feedstock plays a significant role in the amount of energy that is consumed and carbon dioxide ($\text{CO}_2$) produced. About 70% of global ammonia production is based on steam reforming concepts using natural gas, with the use of steam reforming of natural gas considered the best available technology from the point of view of energy use and $\text{CO}_2$ emissions, Table I (14). The use of coal and fuel oil are predominately restricted to China, which exhibits a strong divergence in the ammonia feedstock versus the rest of the world.

<table>
<thead>
<tr>
<th>Energy source</th>
<th>Process</th>
<th>Energy, GJ t$^{-1}$NH$_3$</th>
<th>CO$_2$ emissions, tonnes t$^{-1}$NH$_3$</th>
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<tr>
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<td>Steam reforming</td>
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<tr>
<td>Naphtha</td>
<td>Steam reforming</td>
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<td>Partial oxidation</td>
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![Figure 1. Historical efficiencies of ammonia process technologies](https://example.com)
accounts for 95% of global coal-based ammonia capacity with around 80% of the plants in China being coal-based.

The production of ammonia is a very energy demanding process, the energy use of the steam reforming process is about 28–35 GJ per tonne ammonia (GJ t\(^{-1}\)\(\text{NH}_3\)). **Figure 2** shows the theoretical, practical and operating level energy efficiencies for ammonia plants based on steam reforming. Energy efficiencies vary widely for ammonia plants currently in operation due to age, feedstock, energy costs and utility constraints. Most plants operate well above the practical minimum energy consumption with the best performers (top quartile) ranged between 28 and 33 GJ t\(^{-1}\)\(\text{NH}_3\) and an average efficiency of 37 GJ t\(^{-1}\)\(\text{NH}_3\). It has been estimated that if all plants worldwide were to achieve the efficiency of the best plants, energy consumption could fall by 20–25% (15). A feature of the industry is that most plants are being continually reviewed for improvements and revamp ideas can be subsequently implemented that improve efficiency.

### 3. Technology Development at Billingham, UK

Following pioneering work by Fritz Haber on the process (4), Alwin Mittasch on the catalyst (8), and Carl Bosch on the technology (5) the ammonia synthesis process came to Billingham, UK, in the early 20th century. An ammonia factory being located at Billingham, UK, grew out of the needs of World War I when the British government needed to develop technology to produce synthetic ammonia for producing explosives. Billingham was chosen partly for its proximity to a then-new North Tees electricity generating station nearby; although later developments to the process required less electric power than had been assumed. It is worth noting that even before the plant was begun the possibility for post-war use for fertiliser production was recognised. This was recorded in a report by the Chemical Society in 1916:

> “With some foresight a plant erected primarily for a military purpose might be easily adapted in peace time to agricultural objects” (16).

However by the time the plant (known as the Government Nitrate Factory) was completed, World War I was over. The site was put up for sale in 1919 (**Figure 3**), and was purchased by Brunner Mond & Co Ltd (16) who converted it to make ammonia-based fertilisers. The company was set up as a subsidiary called Synthetic Ammonia and Nitrates Ltd. This became part of ICI in December 1926,
when ICI was formed from the merger of Brunner Mond, Nobel Explosives, the United Alkali Company and the British Dyestuffs Corporation.

3.1 The Coke Oven Process of Syngas Production

The ammonia plants built at Billingham in the 1920s and 1930s employed the classic Haber-Bosch process based on coke, the same as the production technology used in Oppau, Germany. The first Billingham plant was a 24 mtpd (26 stpd) unit that made its first ammonia in December 1924. The original process is shown in Figure 4.

The first stages of gas production were at atmospheric pressure. Alternate streams of steam and then air were fed into gas generators containing hot coke to make ‘water-gas’ (hydrogen-rich) and producer gas (nitrogen-rich). These streams were purified using iron oxides to remove hydrogen sulfide and a shift converter to convert most of the carbon monoxide to CO₂ and H₂. The ‘catalysed gas’ was compressed in reciprocating compressors. CO₂ was removed by counter-current scrubbing with circulating water and the scrubbed gas was further compressed, washed with copper liquor to remove residual CO and CO₂ and then fed as make-up gas to the synthesis loop which contained a large number of parallel converter vessels (Figure 5).

Using this technology the rise in output from the site is shown in Figure 6.

As well as scale improvement there were improvements in effectiveness. In 1929, A. H. Cowap, Chief Engineer, noted: “a striking feature is an ever increasing rapidity of work. The first large unit No. 3 Unit cost £5¼ million pounds and was completed in 27 months (of which 7 months was a labour stoppage for a coal strike). No. 4 and No. 5 units cost £11 million pounds and
have been completed in 2 months” (16). Despite improvements, by the late 1950s increasing costs of coal and the intrinsic inefficiency of syngas generation from coke had made this process uncompetitive.

### 3.2 Partial Oxidation and Plant Safety

The first step to improve process efficiency from coke-oven syngas production was utilisation of higher pressure oil gasification units, a Texaco gasification unit at Billingham for heavy fuel oil was later converted for naphtha feed. Syngas was produced at 30 bar (440 psi) pressure by reaction of the hydrocarbon with steam and a limited supply of oxygen at 1500°C (2732°F). The partial oxidation process reduced both the capital and operating costs of low pressure gas generation, eliminated the need for low pressure compression and offered greater feedstock flexibility. The principle disadvantage of the process was its requirement for an air separation plant to supply oxygen. In these early days the challenges for safe operations and engineering of these air separation units (ASU) were significant.

In 1959 at Billingham’s partial oxidation plant a serious explosion occurred during the commissioning of the ASU which led to three fatalities (17). This incident resulted in a long delay in the partial oxidation plant achieving beneficial operation by which time steam reforming technology development had advanced sufficiently to make it a more competitive route for syngas production.

Within the industry frequent explosions in oxygen plants encouraged engineers to meet and share information. The very first symposium to discuss safety in air and ammonia plants, called Safe Design and Operation of Low Temperature Air Separation Plants, was held in 1955. This meeting became an important event organised annually to improve the safety performance of the ammonia industry. It continues accomplishing these objectives by sharing information on incidents, safety practices, plant performance and technology improvements, with the 62nd meeting of AIChE Safety in Ammonia Plants and Related Facilities Symposium being held in 2017 (18).

### 3.3 Steam Reforming of Light Naphthas

Steam reforming of hydrocarbons provides the most economic source of hydrogen gas for ammonia synthesis. The general steam reforming reaction is shown in Equation (i):

\[
C_nH_{2n+2} + nH_2O \rightarrow nCO + (2n+1)H_2 \quad (i)
\]

The reaction was known to proceed at 700–800°C (1292–1470°F) over a promoted and supported nickel catalyst. ICI was amongst pioneers in methane steam reforming and commercial units had been built at Billingham in 1936 to reform propane/butane byproducts of hydrogenation of coal as part of synthetic hydrocarbons production (Oil Works). This reforming process was operating at atmospheric pressure.

In the 1950s natural gas was not available in the UK (discovery and exploitation of North Sea gas was still some 15–20 years distant), however increasing quantities of light distillate hydrocarbons (naphtha) were available at falling prices. Sulfur free naphthas had been successfully reformed by the catalyst research group at Billingham in 1938 at atmospheric pressure. What was needed was the development of the process to operate at higher pressures to avoid compression costs. The world’s first pressurised steam naphtha reforming process
was designed at Billingham and was brought into commercial operation at Heysham, UK, in 1962 (19). The main problems were adequate desulfurisation of the feed, and suppression of carbon deposition on the reforming catalyst without the use of excessive steam ratios. Desulfurisation of the feed was addressed by development of feed purification technology involving hydrogenation catalysts (nickel-molybdenum, cobalt-molybdenum) along with zinc oxide absorbents capable of reducing sulfur to very low levels. The problem of carbon formation was solved by the development of new types of alkaliised catalysts (20).

Due to equilibrium considerations, to achieve a low methane slip a temperature of around 1000°C (1830°F) is required, however the metallurgical limit for a ten-year life of the available tube materials was a design exit temperature of 800°C (1470°F). To overcome this constraint, the new steam reforming process adopted two reforming stages as shown in Figure 7 (21). Now familiar to us as primary and secondary reformers, these unit operations are still present in nearly all ammonia plants.

### 3.4 Steam Reforming Modernisation

Having developed a viable steam reforming process, the syngas units at Billingham were modernised with four pressured naphtha steam reforming units built in 1962–1963 (Figure 8). Each unit included a primary (tubular) reformer with 4” (100 mm) internal diameter tubes and a reaction length of 20 ft (6 m). Operating at 14 bar (200 psi), the reformed gas, containing 10–12% CH₄, was collected in headers near the ground and passed to the air injection burner in the secondary reformer. After secondary reforming were waste heat recovery, two stage CO shift, further heat recovery, cooling and CO₂ removal. The process was rapidly adopted and by the mid-1960s over 100 steam reforming process licences had been sold from Billingham to the following reputed engineering contractor licensors: Power Gas Corporation (later Davy Power Gas, now Johnson Matthey), Foster Wheeler (now AMEC), Selas, M. W. Kellogg (now KBR), Friedrich Uhde GmbH (now thyssenkrupp Industrial Solutions GmbH) and Humphreys & Glasgow (now Jacobs).
The new steam reforming front end occupied an area of 14,160 m² (3.5 acres) – a little less than 10% of the area occupied by the coke based processes that it replaced. Using space freed up by the reformers, improvements to the gas purification and compression were introduced. The existing CO₂ removal process employed water washing at 55 bar (798 psi) and consumed significant energy leading to high capital and operating costs. Chemical absorbents with higher capacity for CO₂ removal had become available, such as the Benfield process (potassium carbonate) and Vetrocoke (arsenious oxide), and these were adopted on different Billingham plants in the early 1960s. These processes achieved CO₂ slips of <0.1% dry.

3.5 Development of Single Stream Plants

As can be seen with the four reformers above, in contrast to a modern single stream plant, in the early 1960s many units of process equipment (such as compressors and synthesis reactors) still had to be used in parallel. A ‘single stream’ concept emerged for new ammonia plants and Billingham engineers designed and engineered a 360 mtpd (397 stpd) ‘single stream’ plant. Commissioned at Severnside, UK, in 1963, it was, at that time, the largest single stream ammonia plant in the world. The plant used the steam naphtha reforming process, a hot potassium carbonate based CO₂ removal system, a copper liquor CO removal system and had two high temperature shift (HTS) beds, however parallel reciprocating compressors driven by electric motors were used for synthesis gas compression.

Although more active copper-based catalysts were known to be able to accomplish the shift reaction at ~250°C (482°F), they were very sensitive to poisoning by sulfur and could not be used with syngas made from coke. The virtually sulfur free syngas obtained from steam reforming allowed these copper-based shift catalysts to be used, achieving an equilibrium CO conversion of ~0.2% dry. ICI developed its own robust and reactive Cu catalyst for this application (22). The combined low level of residual carbon oxides from CO₂ removal and CO shift was now low enough that they could be made inert by methanation before the synthesis loop. As a result of all of the improvements considered so far – steam reforming, shift, CO₂ removal and methanation – by the mid-1960s it was possible to carry out all operations in single stream reactors.

Early design memos for the Billingham plants in the 1920s had discussed the relative economics of reciprocating and centrifugal compression. They showed that the relative efficiency of rotary compressors for the later compression stages would be low except at high throughputs (Figure 9). This low efficiency was one of the reasons for excluding centrifugal compressors for all but the low pressure stages and this reasoning still prevailed at the beginning of the 1960s; however it was now challenged. A second plant for ICI Severnside, with a capacity of 545 mtpd (600 stpd), led to what was described as “possibly the most important event in the history of the development” of the single stream ammonia plant (23). Figure 10 (24) showing the increase in output of ammonia converters from 1930 to the mid-1960s illustrates the revolution in scale taking place. By then the normal ammonia unit size was already 600 mtpd (662 stpd) capacity and plants with lower capacity than this were being regarded as small.

In late 1962, a meeting was held in which Ron Smith, Vice President of Operations at M. W. Kellogg, opined that the capacity of ammonia plants was bound to increase and queried why the synthesis loop pressure could not be reduced from 325 bar (4700 psi) to 150 bar (2200 psi), thus removing the need for reciprocating compressors. Although not successful for that plant bid soon afterwards discussions began on how a large plant could be built – and the plant design was established which quickly became a new technology era that came...
In January 1964, M. W. Kellogg was awarded a contract for two ‘1000 stpd’ (900 mtpd) plants to be built at Billingham. The design incorporated a number of important features:

- The steam naphtha reforming process at 31 bar (450 psi) pressure
- A loop pressure of 131 bar (1900 psi) allowing the use of centrifugal compressors
- Improved plant efficiency by recovering heat to generate 103.5 bar (1500 psi) high pressure steam superheated to 450°C (850°F) for use on steam turbine drives.

The steam was generated at a higher pressure than that required by the process, so energy was recovered by expanding the steam through turbines to the pressure level required by the process. This greatly enhanced process efficiency. Within a few weeks a third plant was announced, they were the largest plants built at that time.

### 3.6 M. W. Kellogg Ammonia Units

As the M. W. Kellogg plants incorporated the steam naphtha reforming process, Billingham engineers worked closely with their counterparts from M. W. Kellogg in the design of the reformers, shown in Figure 11. As by the mid-1960s exploration was ongoing for North Sea gas this was considered and a feature of the reforming process was that the plants could be readily converted to lighter hydrocarbons.

In keeping with its status as an operator, designer, technology licensor and catalyst manufacturer, ICI continued to develop its own technology. Two Billingham designed ammonia plants, constructed in Kanpur, India, in 1969, featured the first application of ICI’s single nozzle secondary burner (Figure 12); the forerunner of a design used in many ICI (and subsequently Johnson Matthey) designed plants using autothermal reforming technology.

### 3.7 Use of Natural Gas Feedstock

In the 1970s Billingham’s ammonia plants changed from naphtha feeds to run on the newly commercialised natural gas from the North Sea, however the favourable gas contract was on an interruptible supply basis, meaning that with short notice the feedstock could be cut when demand
for natural gas was high. If natural gas supply was interrupted the plants were configured to switch feedstock on-line to liquefied petroleum gas (LPG) propane feedstock (which was stored locally in underground salt caverns), bringing a demand for catalysts that could cope with feedstock flexibility. This brought new requirements for a catalyst with lower potash and higher activity in order to optimise the reformer for this feedstock. By the end of the decade there were two light potash catalysts: 25-3 (1.6% K₂O) for natural gas feeds and 46-9 (2.2% K₂O) for LPG feeds. By the end of the 1970s, ICI Katalco had a product range very similar to the present: 57-series non potash, 25-series light potash, and 46-series naphtha catalyst. By this point the catalyst beds were operating at temperatures up to 1000°C (1832°F) and 35.6 bar (516 psi), primarily due to improvements in metallurgy.

In the 1970s, it was recognised that appropriately formulated low-temperature shift (LTS) catalysts could be self-guarding not only in regard to sulfur, but also towards chloride. It was also recognised that the benefits in terms of shift activity and bed life accruing from the use of fresh LTS catalyst outweighed the cost savings realised by reusing discharged LTS catalyst. All LTS catalysts subsequently developed by ICI and Johnson Matthey were therefore optimised to maximise their self-guarding capability.

These catalyst systems were utilised in the three M. W. Kellogg ammonia plants and also in the ICI designed Ammonia IV plant (Figure 13). Designed for 1000 mtpd (1100 stpd), this was commissioned in 1978 and was able to achieve a throughput of about 1125 mtpd, (1240 stpd) without significant modification.

### 3.8 The Ammonia V Process

Ammonia technology continued to develop and Billingham-based engineers were tasked with the design of a fifth ammonia plant for Billingham (Ammonia V or 'AMV'). Economic considerations meant that capital cost had to be reduced whilst improving plant efficiency. Although market conditions in the early 1980s meant that the plant was never built at Billingham, the designs for Ammonia V evolved into the AMV process. The first AMV design was commissioned at Courtright, Canada, in August 1985 (Figure 14), producing 1120 mptd (1234 stpd) at a total energy requirement of 29 GJ per metric tonne (lower heating value, LHV). Ammonia production was achieved 43 hours after feed gas introduction, believed to be a record at that time (25). The AMV process also featured a low pressure synthesis loop operating at about 85 bar (1230 psi) featuring a new cobalt-promoted high-activity ammonia synthesis catalyst (KATALCO™ 74-1) which had been developed specifically for the project.

As a highly efficient process operating with a low steam ratio, the plant was one of the first to suffer from byproduct formation and pressure drop increase due to HTS over reduction. Copper was added to the HTS catalyst formulation to create an
over reduction resistant formulation which was first installed in 1987.

3.9 The Leading Concept Ammonia Process

By the mid-1980s, the two ammonia plants at Severnside were becoming uncompetitive and a decision had to be made: improve their efficiency, replace them or close the site. Improving the efficiency was thought unfeasible and it was decided to develop a new process to replace them. This led to the leading concept ammonia (LCA) process technology being developed at Billingham (Figure 15).

The LCA process used a combination of new equipment, new catalysts and improved construction and procurement techniques. The range of developments included:

- KATALCO 61-2 (the first low-temperature hydro-desulfurisation (HDS) catalyst)
- PURASPEC™ 2020 (the first low-temperature sulfur removal absorbent)
- KATALCO 83-1 (the first application of a process gas heated reformer (GHR), isothermal shift catalyst specifically developed to resist the high operating temperature)
- KATALCO 11-4 (a low-temperature methanation catalyst)
- KATALCO 74-1 (a catalyst which could be used in an ammonia synthesis loop at 80 bar (1160 psi) pressure, even lower than in the AMV process).

The unique process together with extensive automation start-up sequences meant the plants were amongst the most automated ever. The second plant at Severnside made ammonia only 19 hours after natural gas was first introduced.

3.10 Catalyst Developments

Catalyst developments continued into the 1990s. Figure 16 illustrates the dramatic improvement in the activity of one particular catalyst which resulted from a combination of on-going development and the incorporation of learning from the development of the technology for LTS catalysts.

A step change occurred in 1997 due to the acquisition of the BASF syngas catalyst business by ICI’s catalyst business (since acquired by Johnson Matthey). The acquisition of the BASF activities allowed the knowledge of two historic companies to be combined and in this case the best of both companies created a new improved LTS catalyst. Methanol is an unwanted byproduct that may be formed in LTS reactors and is the main volatile organic compound (VOC) emitted from ammonia production plants. It is formed as a byproduct in both high-temperature and low-temperature shift. Through the 1990s byproduct methanol was an increasing concern for plants as environmental emissions came under closer attention. More selective catalysts became available that made less methanol. BASF previously had low methanol LTS catalysts, K3-110 and K3-111, which suffered from issues relating to physical strength and poisons resistance. ICI had LTS products with good strength characteristics, but could not mimic the BASF low methanol recipe due to patent protection. The combination of the two businesses meant that a low methanol, high-strength product could be developed. The results were KATALCO 83-3K, launched in 1997, and KATALCO 83-3X,
which was launched in 1998 with 90% reduction in methanol byproduct formation compared to previous generations of catalyst (Figure 17).

![Relative LTS selectivity of the KATALCO catalysts, measured by methanol production rate. The numbers above the bars refer to the catalyst series](image)

### 3.11 Developments in Catalyst Shape

The effect of shape on reforming catalysts has been recognised for a long time (21). For steam reforming catalysts, the reaction occurs in a very thin layer at the surface of the pellet. Therefore, developments focused on techniques to develop the shape, maximising the external surface area of the catalyst pellets whilst at the same time considering the resistance to flow caused by the way the catalyst packs in the tube. The shape of the steam reforming catalysts evolved from the original cubes (circa 1930s) to Raschig rings (circa 1940s) to ICI Katalco 4-hole (circa 1980s) and finally the current KATALCO QUADRALOBE™ shape. At each iteration, for similar sized pellets the activity increased and the pressure drop decreased (20). Increasing the catalyst activity also allowed the reforming reaction to progress at a lower temperature, which meant the tubes were also at a lower temperature as shown by the measured tube wall temperature (TWT). The lower the peak maximum TWT, the longer the tube metallurgy lasts before failure, with a difference of as little as 20°C (68°F) doubling the tube life.

Since the 1990s design tools such as finite element analysis have been used to assist with the design and optimisation of catalyst shape. The latest development for the steam reforming process is the CATALCEL™ technology, which Johnson Matthey purchased in 2014. CATALCEL SSR™ is a stackable structural catalytic reactor for the production of hydrogen from natural gas. It is made from a high-temperature stainless steel foil coated with a reforming catalyst. This structure allows higher heat transfer and can provide significant capacity increase to reformers or lower pressure drop compared to standard pelleted catalysts.

Further developments have been made by shaping the pellets in some of the other reactors which follow the steam reformer in the production of syngas at the front-end of the plant, notably the HTS and methanator. For example, KATALCO 71-5F (Figure 18) is a shaped 5-lobe pellet HTS catalyst which exhibits lower pressure drop, increased strength and increased voidage. Similarly, for the methanation reactor, KATALCO 11-6MC (Figure 19) uses a 4-hole clover leaf shape to provide lower pressure drop with increased bed voidage. The benefit of pressure drop reduction in the front end varies from plant to plant depending on the individual process constraints. Generally...
pressure drop is welcome and a small increase in energy efficiency can be gained if it is reduced.

3.12 The Dual-Pressure Process

In 1998, ICI and Uhde (now thyssenkrupp Industrial Solutions (tkIS)) formed an alliance in the field of ammonia technology resulting in a variety of new developments, the most public of which was the dual-pressure ammonia process (28). The resulting 3300 mtpd (3640 stpd) plant was a step change in the scale of plant design available and offered a reduction of specific production costs through economies of scale. These are still being built today as the world’s largest ammonia plants.

The key innovation in the Uhde dual-pressure ammonia process was an additional medium-pressure once-through ammonia synthesis step operating at around 110 bar (1595 psi), connected in series with the conventional high-pressure ammonia synthesis loop at around 200 bar (2900 psi), Figure 20. The first plant based on this process was the SAFCO IV ammonia plant in Al Jubail, Saudi Arabia, started up in 2006. With a capacity of 3300 mtpd (3640 stpd) it was by far the largest ammonia plant worldwide, Figure 21.

Since then tkIS’s Uhde dual-pressure process has been implemented in other similar scale plants recently commissioned in regions of the world with an abundance of low cost natural gas feedstock (Table II).

4. Ammonia Production Today

Figure 22 shows the current plant capacity and year of construction for all operating ammonia plants. There is a clear progression of increasing plant scale with time. Market needs for individual plants will differ, leading to a range in plant capacities. There are however preferred plant sizes which have become ‘standard’ in the industry for which references and documented plant designs exist. These can be clearly seen in Figure 22 at capacities of 600 mtpd, 1000 mtpd, 1360 mtpd, 1500 mtpd, 2000–2200 mtpd and most recently 3300 mtpd (3640 stpd). It is notable that, as well as being the largest production units in the world, the emission limits for the new US fertiliser projects (ammonia and downstream plants) are amongst the lowest in the world, with the design levels for emissions of NOx, N₂O, CO and volatile organic compounds (VOC) being significantly lower.
below current recognised best available techniques (BAT) values (26).
In just over 100 years, the nitrogen fertiliser industry based on ammonia production has grown massively (**Figure 23**). Drivers behind this growth have been, and remain, increasing global population (**Figure 24**) (9) coupled with increased plant size to achieve better economies of scale. Although the picture is more complex than this (for example, one could ask which came first: fertiliser or population growth?), together this has created demand for increased capacity and increased reliability from that capacity.

### 5. Conclusion

Over the last century, scientists and engineers have made a significant contribution to the nitrogen industry. Some of these have been based at Billingham, UK, whose heritage now resides with Johnson Matthey and the challenge is to continue

<table>
<thead>
<tr>
<th>Plant</th>
<th>Location</th>
<th>Capacity, mtpd</th>
<th>Start-up year</th>
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<td>Al Jubail, Saudi Arabia</td>
<td>3300</td>
<td>2006</td>
</tr>
<tr>
<td>Saudi Arabian Mining Company, Ma’aden</td>
<td>Raz Az Zwor, Saudi Arabia</td>
<td>3300</td>
<td>2011</td>
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<tr>
<td>CF Industries, Donaldsonville, Ammonia 6</td>
<td>Donaldsonville, LA, USA</td>
<td>3300</td>
<td>2016</td>
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<tr>
<td>Saudi Arabian Mining Company, Ma’aden 2</td>
<td>Raz Az Zwor, Saudi Arabia</td>
<td>3300</td>
<td>2016</td>
</tr>
</tbody>
</table>

**Table II 3300 mtpd tkIS Uhde Ammonia Plants with Johnson Matthey Catalyst**

![Graph showing trends of plant capacity vs. year of construction](image)
Fig. 23. Global Haber-Bosch ammonia production from mid-20th century to the present. Over 99% of fixed nitrogen production today is by the Haber-Bosch process (2) (Copyright The Fertilizer Institute, used with permission)

Fig. 24. Demographic drivers for Haber-Bosch nitrogen and its use in fertiliser: “...the lives of around half of humanity are made possible by Haber-Bosch nitrogen” (2, 9) (Copyright The Fertilizer Institute, used with permission)
this legacy and make an equally significant contribution to the future of this vital industry. The fundamental ammonia synthesis process and catalysts developed by Haber-Bosch and Mittasch can still be clearly recognised in even the most modern ammonia plants. However, the process efficiencies and environmental performances have been dramatically improved over the last 100 years, most particularly in the preparation of the synthesis gas, benefiting both ammonia production and other syngas-based processes. Because energy utilisation within modern processes is near the theoretical minimum, specific energy consumption can be reduced only marginally, if at all. There are many future challenges for ammonia and the fertiliser industry, which fall outside the scope of this historical overview.

For now, the ammonia industry will be with us more or less in its present form for decades to come (27). The present production capacity for synthetic ammonia of over 175 million metric tonnes per year will continue to grow at 1–2% every year to satisfy the increasing demands for food and ammonia-based intermediates from an increasing number of people enjoying increasing welfare.

Trademarks
KATALCO, PURASPEC, QUADRALOBE, CATACEL and SSR are trademarks of Johnson Matthey.

References
18. 62nd Annual Safety in Ammonia Plants and Related Facilities Symposium, New York, USA, 10th–14th September, 2017

**Further Reading**

J. Larsen, D. Lippmann and C. W. Hooper, 'A New Process for Large-Capacity Ammonia Plants', *Nitrogen & Methanol*, 2001, 253, (September-October), 41

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