

TITLE OF THE INVENTION

Process and apparatus for cracking ammonia

TECHNICAL FIELD OF THE INVENTION

5 The present invention is in the technical field of cracking ammonia to produce hydrogen gas and preferred embodiments relate specifically to a process and apparatus for the production of hydrogen gas from liquid ammonia.

BACKGROUND OF THE INVENTION

10 Global interest in renewable energy and using this renewable energy to generate "green" hydrogen has driven the interest in converting the "green" hydrogen to "green" ammonia, as ammonia is simpler to transport over distance of hundreds or thousands of miles. Particularly, shipping liquid hydrogen is not commercially possible currently but shipping ammonia, which is in a liquid state, is currently practiced.

For use in a commercial fuel cell, the ammonia must be converted back to hydrogen according to the reaction.



This is an endothermic process, *i.e.*, a process that requires heat, and hence higher temperatures will favor production of the products. The standard heat of reaction (per mole of ammonia) at 1 bar and 0°C is 45.47 kJ/mol. The endothermic nature of the process dictates the need for a furnace.

20 The process is known as cracking (or sometimes "dissociation") and is usually performed over a catalyst. The gas produced (or "cracked gas") is a mixture of hydrogen (H₂) and nitrogen (N₂) gases although, since the cracking reaction is an equilibrium reaction, there is also some residual ammonia. The amount of ammonia in the cracked gas, generally referred to as "ammonia slip", may be varied by changing the temperature at which the ammonia is heated with higher
25 temperatures favoring conversion thereby reducing the ammonia slip.

In most applications of crackers currently, the hydrogen and nitrogen mixture is utilised as is. However, as ammonia can be a poison to fuel cells, this stream, with ammonia suitably removed such as by scrubbing with water, can be used directly in a fuel cell. However, if the hydrogen is to be used in vehicle fueling, the nitrogen present provides a penalty to the process. The fuel to
30 a vehicle fueling system is compressed to significant pressure - up to 900 bar. This means that

the nitrogen, which is merely a diluent in the process, is also compressed, taking power, and taking storage volume and increasing anode gas purge requirement, decreasing efficiency. It is therefore beneficial where hydrogen is to be used in vehicle fueling, for the hydrogen and nitrogen to be separated.

- 5 There are many examples of ammonia cracking processes in the art, for example GB977830A, JP5330802A, CN111957270A, US2020/0398240A and KR2022/0085469A.

In addition, GB1142941A discloses a process for producing a fuel gas that is interchangeable with town gas. Ammonia is cracked to form a mixture of hydrogen and nitrogen and the mixture is then enriched by the addition of a gas having a calorific value higher than that of the mixture
10 such as methane, propane or butane or mixtures thereof. Liquid ammonia is pumped as a cold liquid and vaporised by a closed hot water circuit. The ammonia vapor is cracked over a suitable catalyst in the tubes of a direct fired tube furnace after having been superheated by heat exchange with the flue gas from the furnace. The cracked gas is scrubbed with water to recover residual ammonia which is eventually recycled to the ammonia feed to the catalyst-filled tubes of the
15 furnace. The purified cracked gas is enriched with propane and/or butane to produce the town gas product.

GB1142941A discloses that the cracking of the ammonia in a direct fired tube furnace in the presence of a suitable catalyst is preferred. However, the reference also discloses that other cracking processes could alternatively be used instead. In this context, GB1142941 mentions
20 heating the ammonia to a suitable temperature and then passing the ammonia through an unheated bed of ammonia cracking catalyst to crack a proportion of the ammonia into hydrogen and nitrogen, cooling the gas in the process. The unconverted ammonia could be recovered as described above. Alternatively, the gas mixture could be reheated and passed through a second bed of catalyst to reduce further the ammonia content and that this could be repeated as many
25 times as considered desirable.

GB1353751A discloses a process in which ammonia at a pressure in a range from 20 atm to 300 atm is cracked in two stages within heated reactor tubes. In the first stage, the gas at a temperature in a range from 450 to 800°C is passed through a layer of nickel-, iron- or cobalt-containing catalyst produced by co-precipitating with a carrier of aluminum oxide and magnesium
30 oxide or magnesium-aluminum spinel. The catalyst of the first stage may alternatively consist of either a ceramic material impregnated with iron or iron impregnated on a pre-formed carrier consisting of magnesium oxide and aluminum oxide and promoted with potassium oxide. After

the first stage, the gas then passes at a temperature in a range from 450 to 600°C through a layer of a doubly or triply promoted iron catalyst which forms the second stage.

WO2022/189560A discloses an ammonia cracking process involving a fired reactor having tubes filled with iron catalyst. Liquid ammonia is taken from storage, pumped and then pre-heated and evaporated to form ammonia gas which is heated by heat exchange with cracked gas. The heated ammonia gas is further heated by heat exchange against flue gas in the convection section of the fired reactor and is then fed to two adiabatic reactors in series (with interstage heating against flue gas) where it is partially cracked. The partially cracked gas is then heated by heat exchange with the flue gas in the convection section before being fed to the catalyst-filled tubes of the fired reactor to crack the remaining ammonia.

US11287089A discloses a hydrogen fueling system in which ammonia is cracked onsite into hydrogen and nitrogen in an ammonia cracker operating at a pressure in the range from 5 bar to 40 bar, and in which the hydrogen is compressed to a pressure of at least 30 MPa (300 bar) and stored ready for dispensing to vehicles. When being dispensed, the compressed gas from storage is cooled to a temperature in a range from -40°C to 5°C, by heat exchange with a heat exchange fluid, such as D-limonene, FP40 or a water/glycol mixture, circulating around a closed loop. The heat exchange fluid is cooled by heat exchange with the ammonia feed to the cracker and may be further cooled in a conventional refrigeration system. Where the ammonia feed is liquid, at least part of the duty required to vaporize the liquid ammonia is provided by the heat exchange fluid. The vaporization of liquid ammonia is usually effected at normal or sub-atmospheric pressure. US11287089A exemplifies a system producing 7.5 tons/day hydrogen gas.

There is, however, still a need generally for improved processes for the production of hydrogen from ammonia and specifically for processes that are more efficient in terms of energy consumption and/or that have higher levels of hydrogen recovery and/or that reduce or eliminate the need to combust fossil fuels.

BRIEF SUMMARY OF THE INVENTION

According to a first aspect of the present invention, there is provided a process for cracking ammonia comprising providing a heated ammonia gas at super-atmospheric pressure; combusting fuel with an oxidant gas in a furnace to heat catalyst-containing reactor tubes, each tube comprising an upstream layer of a first catalyst and a downstream layer of a second catalyst, and produce a flue gas; feeding the heated ammonia gas, or a partially cracked ammonia gas

derived therefrom, to the catalyst-containing reactor tubes to cause cracking of ammonia and produce a cracked gas comprising hydrogen gas, nitrogen gas and residual ammonia gas, wherein the first catalyst is more active for cracking ammonia than the second catalyst.

Using layers of different catalysts having different activities in the reactor tubes of an ammonia cracker is not new. In the known processes, the more active catalyst is used conventionally in a layer of the reactor tubes that is downstream of an upstream layer of less active catalyst, typically the final section of the catalyst bed at the outlet of the tubes, to ensure a close approach to equilibrium.

The Inventors have discovered, however, that using the more active catalyst in an upstream layer of the reactor tubes, typically the first section of the catalyst bed at the inlet of the tubes, is beneficial as it reduces nitriding of the metal of the reactor tubes.

Metal nitriding is a recognized problem in ammonia crackers. The amount of nitriding is dependent on the partial pressures of ammonia, nitrogen, and hydrogen and the temperature of the process. Nitriding is most severe at higher temperatures and higher partial pressure of ammonia and nitrogen, with ammonia being the greater risk at typical ammonia cracking temperatures (nitriding due to nitrogen is not likely until the temperature is excessively high for ammonia cracking, e.g., 1000°C or more). The exact extent to which the material of construction forms nitrides is currently unknown. However, there is data to indicate that below 650°C the extent of nitriding should be acceptable depending on the chosen material of construction.

Operating ammonia crackers at elevated pressures, e.g., 10 to 50 bar, increases the partial pressure of the ammonia and nitrogen and may lead to more severe failure if nitriding results in brittle fracture of the reaction vessel or tubes.

This problem is of particular concern in fired reactors where the temperature surrounding the reaction tubes holding the process side catalyst is significantly above 650°C.

In a top fired furnace, the burners are located at the top and the burner flames extend down part of the length of the reactor tubes. The ammonia feed is fed to the top of the tubes and flow downwards, i.e., co-currently with the flue gas, to the bottom of the tubes. Since the tops of the tubes are nearest the burner flames, this region of the tubes has the highest heat flux. The tops of the tubes are also where the highest partial pressure of ammonia on the process side is found. Therefore, this region of the tubes has the highest potential for nitriding and hence is of highest concern.

Locating a more active catalyst in the first section of the catalyst bed of the reactor tubes leads to a stronger endotherm drawing more heat from the metal of the tubes and into the cracking process itself, thereby reducing the temperature of the inner wall of the tubes and enabling the inside wall temperature to be kept below a pre-defined limit of, e.g., 700°C (or 660°C to allow a margin).

- 5 Reducing the inside wall temperature in this way reduces nitriding of the metal walls of the tubes.

According to a second aspect of the present invention, there is provided a furnace, typically a top fired furnace, for cracking heated ammonia gas at super-atmospheric pressure comprising a radiant section comprising at least one inlet for fuel and oxidant gas in fluid flow communication with at least one burner, an ammonia feed inlet, catalyst-containing reactor tubes having upstream
10 ends in fluid flow communication with the ammonia feed inlet and downstream ends in fluid flow communication with an outlet for cracked gas, each tube comprising an upstream layer of a first catalyst and a downstream layer of a second catalyst; and a convection section in fluid flow communication with the radiant section and comprising an outlet for flue gas; wherein the first catalyst is more active for cracking ammonia than the second catalyst.

- 15 The catalyst bed within each reactor tube typically contains two or three layers (or sections) of catalysts. Beds having more than three layers of catalyst may also be suitable for specific applications.

The upstream layer is typically the first layer of the catalyst bed within each reactor tube and will comprise, e.g., contain or consist of, the first catalyst. The upstream layer has a length that may
20 be at least 20%, or at least 25%, or at least 30%, or at least 35% of the length of the entire catalyst bed. Typically, the length of the upstream layer is not more than 50% of the length of the catalyst bed. Usually, the upstream layer extends beyond the end of the flames of the burners.

The downstream layer is typically the second layer of the catalyst bed and will comprise, e.g., contain or consist of, the second catalyst. While there may be an intervening layer between the
25 upstream and downstream layers, this is not usually the case. Where there are only two layers in the catalyst bed, the length of the downstream layer will be the remainder of the bed. However, in some embodiments, there will be a layer of a third catalyst downstream of the layer of the second catalyst. The length of the layer of third catalyst would not typically be more than 10% of the entire bed length. Thus, where there are three layers in the catalyst bed, the downstream
30 layer will again make up the rest of the bed.

The third catalyst is ideally more active than the second catalyst so, if present, the layer of third catalyst enables a close approach to equilibrium.

The furnace of the second aspect of the invention is particularly suitable for carrying out the process of the first aspect of the invention.

According to a third aspect of the present invention, there is provided apparatus for cracking heated ammonia gas at super-atmospheric pressure comprising a source of liquid ammonia; a pump in fluid flow communication with the source of liquid ammonia for pumping liquid ammonia; and a furnace as defined in the second aspect wherein the ammonia feed inlet of the furnace is in fluid flow communication with the pump, wherein the apparatus further comprises at least one heat exchanger arranged for pre-heating liquid ammonia upstream of the pump; and at least one heat exchanger arranged for vaporizing pumped liquid ammonia and heating ammonia gas by heat exchange with flue gas and/or cracked gas located between the pump and the ammonia feed inlet of the furnace.

The apparatus of the third aspect of the invention is particularly suitable to carry out the process of the first aspect of the invention.

BRIEF DESCRIPTION OF THE DRAWINGS

Fig. 1 is a simplified flowsheet for an ammonia cracking process within which the present invention may be used;

Fig. 2 is a graph plotting how various temperatures and ammonia mole fraction vary as a function of cracker length for a comparative example using a nickel-based catalyst as the sole catalyst; and

Fig. 3 is a graph plotting how various temperatures and ammonia mole fraction vary as a function of cracker length for an example of the invention using an upstream layer of ruthenium-based catalyst and a downstream layer of nickel-based catalyst.

DESCRIPTION OF THE INVENTION

Amounts of components given in parts per million (or ppm) are calculated by weight unless otherwise stated. In addition, all percentages are calculated by mole unless otherwise stated. Moreover, any references to pressure are references to absolute pressure unless otherwise stated.

In the context of the present invention, the activity of a catalyst will be understood to refer to the rate of conversion of ammonia at a given partial pressure for a given amount of catalyst over a given period of time at a particular temperature and overall pressure. The units used to define

the activity of a heterogeneous catalyst are mole of ammonia converted per gram of catalyst (including substrate if present) per second (or mol. g⁻¹ s⁻¹).

5 The expression "in fluid flow communication" will be understood to mean that piping or other suitable conduits will be used to convey fluid from one specified location to another. During passage between the two locations, the fluid may flow through one or more other units which may be designed and/or arranged to alter the physical condition, *e.g.*, temperature (*e.g.*, a heat exchanger) and/or pressure (*e.g.*, a compressor or a pump) of the fluid, or the composition of the fluid through reaction of components within the fluid (*e.g.*, a catalytic reactor). The expression "in direct fluid flow communication" will be understood to mean that the fluid flows directly from the one location to the other, *i.e.*, does not flow through another such unit during its passage and hence there is essentially no change to the composition or physical condition of the fluid.

15 The term "super-atmospheric pressure" will be understood to mean a pressure that is significantly higher than atmospheric pressure, such as a pressure of at least 5 bar, *e.g.*, a pressure of at least 10 bar, or of at least 20 bar or of at least 30 bar. Typically, the pressure is no more than 60 bar.

15 The term "upstream" will be understood to mean in the opposite direction to the flow of fluid during normal operation. The term "downstream" will be interpreted accordingly.

A typical composition of the ammonia feed is shown in Table 1.

Component	Composition
Ammonia	99.5 – 99.8 wt%
Maximum water content	0.5 wt%
Minimum water content	0.2 wt%
Oil	5 ppm max
Inerts (<i>e.g.</i> , Ar)	<10 ppm
Iron	< 1 ppm

TABLE 1

Oil may be present in the ammonia due to a boil-off gas compressor used for the ammonia storage, either at the local storage tank, production location, or any other storage tank in between. The presence of oil is an issue because it presents a blockage and/or contamination risk. This may lead to malperformance in the heat exchangers or reduced catalyst activity in the reactors. Therefore, if present, the oil may need to be removed in some way. In this regard, the oil can be removed by passing the liquid ammonia through a bed of activated carbon. However, in preferred

embodiments, the catalyst used in the adiabatic reaction unit will crack the oil into shorter chain hydrocarbons which may react with any water present to form carbon monoxide, hydrogen, and methane.

Inert gases are not expected to be an issue, other than they could end up in the product hydrogen.

- 5 In this regard, helium can be present in ammonia derived from natural gas but ammonia from renewable hydrogen will not contain helium.

The liquid ammonia is typically pumped from the storage pressure (e.g., about 1 bar) to a pressure in a range from about 5 bar to about 60 bar, e.g., from about 10 bar to about 50 bar such as from about 10 bar to about 30 bar or from about 40 bar to about 50 bar. The temperature of the liquid ammonia increases slightly from the storage temperature (e.g., about -34°C) to about -32°C.

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The pumped liquid ammonia (at super-atmospheric pressure) is then pre-heated to its boiling point, ideally by appropriate heat integration within the process. Preferably, part of the pre-heating is achieved using a heat transfer circuit where the heat, e.g., from the intercooling and aftercooling of a PSA off gas compressor, is recovered using a heat transfer fluid such as an aqueous solution of a glycol, e.g., an aqueous solution comprising from about 50 wt. % to about 60 wt. % of a glycol such as ethylene glycol or propylene glycol, optionally together with heat from the cracked gas and/or flue gas, and used to heat the liquid ammonia. If no such integration is possible, such as if the compressor is not running, then an electric heater could be required to preheat the ammonia.

15

The pre-heated liquid ammonia is then evaporated, and the ammonia gas heated further prior to being fed to catalyst-containing reactor tubes or to an adiabatic reaction unit. In this regard, the ammonia gas is typically superheated, i.e., heated to a temperature above its boiling point, to a temperature of more than 350°C to ensure a useful rate of reaction in the catalyst-containing reactor tubes or the adiabatic reaction unit.

20

The duty for the heating and evaporation of the pre-heated liquid ammonia may be provided by the cracked gas, the flue gas or a combination of both the cracked gas and the flue gas. In preferred embodiments, the cracked gas is used to heat and evaporate the pre-heated liquid ammonia by heat exchange and then the ammonia gas is further heated by heat exchange with the flue gas.

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The heated ammonia gas may be fed directly to the catalyst-containing reactor tubes, i.e., without first partially cracking some of the ammonia.

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The reactor tubes of the furnace are filled with at least two different ammonia cracking catalysts having different activities, of which the first (more active) catalyst is located upstream of the second (less active) catalyst. In some embodiments, the bed in each reactor tube has only two layers, the upstream layer of the first catalyst and the downstream layer of the second catalyst.

5 However, in other embodiments, the bed may have a layer of a third catalyst downstream of the layer of the second catalyst in which the third catalyst is more active than the second catalyst which may enable shortening of the tubes. The catalytically active metal of the first and third catalysts may be different although in preferred embodiments, the metal is the same, *e.g.*, ruthenium.

10 The activity of the first (or third) catalyst is typically at least 50% greater than that of the second catalyst. However, the difference in relative activities is usually substantially more than 50%. In this regard, the activity of the first (or third) catalyst is usually greater than the activity of the second catalyst by a factor of at least two (*i.e.*, double), or at least three, or at least four, or at least five. In some embodiments, the activity of the first (or third) catalyst is greater than the activity of the
15 second catalyst by a factor of at least ten (*i.e.*, by one order of magnitude) or by a factor of at least fifteen.

A large number of metals are known in the art to catalyze the cracking of ammonia. These metals include transition metals such those in Group 6 of the Periodic Table, *e.g.*, chromium (Cr) and molybdenum (Mo); Group 8, *e.g.*, iron (Fe), ruthenium (Ru) and osmium (Os); Group 9, *e.g.*, cobalt
20 (Co), rhodium (Rh) and iridium (Ir); Group 10, *e.g.*, nickel (Ni), palladium (Pd) and platinum (Pt); and Group 11, *e.g.*, copper (Cu), silver (Ag) and gold (Au). Metalloids such as tellurium (Te) may also be used.

The activity of some of these metals as catalysts for ammonia cracking has been reported by Masel *et al* (Catalyst Letters, vol. 96, Nos 3-4, July 2004) to vary in the following order:

25
$$\text{Ru} > \text{Ni} > \text{Rh} > \text{Co} > \text{Ir} > \text{Fe} \gg \text{Pt} > \text{Cr} > \text{Pd} > \text{Cu} \gg \text{Te}$$

The metals may be unsupported but are usually supported on a suitable support, typically a metal oxide support such as silica (SiO₂), alumina (Al₂O₃), zirconia (ZrO₂) or a mixed metal oxide support such as spinel (MgAl₂O₄) or perovskite (CaTiO₃).

As would be understood by the skilled person, the activity of a supported metal catalyst will
30 typically depend in part on the loading of the catalytically active metal on the support. In this

regard, the loading of the metal will vary according to the specific requirements but will typically be in a range from about 0.1 wt % to about 70 wt %. The loading may be towards the lower end of the range, e.g., from about 0.1 wt % to about 10 wt % or from about 0.2 wt % to about 5 wt %, for the more active metals, e.g., ruthenium. For less active metals, e.g., nickel, the loading may be towards the upper end of the range, e.g., from about 20 wt % to about 65 wt %.

Supported metal catalysts may be unpromoted or may be promoted with at least one other metal, e.g., one or more Group 1 metals, e.g. lithium (Li), sodium (Na) and potassium (K); Group 2 metals, e.g., magnesium (Mg) and calcium (Ca), or Group 13 metals, e.g., aluminum (Al), to improve activity as is well known in the art.

Any conventional catalysts known for ammonia cracking may be used in the present invention. Suitable catalysts are disclosed in US2015/0217278A, Masel *et al* (above), Lamb *et al* (Int. J. Hydrogen Energy, 44 (2019) pp 3726-3736) and Boisen *et al* (J. Catalysis 230 (2005) pp 309-312).

Bimetallic catalysts, or catalysts containing two catalytically active metals, are also suitable for use with the present invention. Examples include the composite metal or metal alloys or metal nanoclusters supported on perovskites, composite oxides or nitrides, or mixed oxides or mixed nitrides disclosed in US2021/0001311A such as CoNi-MgSrCeO₄ and 1 wt % K-CoNi-MgSrCeO₄.

The catalysts of the present invention typically comprise, e.g., contain or consist of, metal-based catalysts. The catalytically active metals are typically selected from the transitional metals of the Periodic Table. Suitable transition metal-based catalysts typically have an activity at a temperature in the range of 475°C to 600°C suitable to achieve a reaction rate of at least 0.2 times, or at least 0.5 times, or at least equal to, or at least twice, or at least five times, the rate calculated according to equation 9 proposed by Lamb *et al*, i.e.,

$$r = 8.73 \exp [-76710/RT] \cdot (P_{NH_3})^{0.28} \cdot (P_{H_2})^{-0.42} \cdot (1-\beta^2)$$

where:

"r" is the reaction rate (or "activity") of the catalyst;

"RT" is the ideal gas constant "R" (8.314 J mol⁻¹ K⁻¹) multiplied by the temperature "T" in Kelvin;

P_{NH3} is the partial pressure of ammonia;

P_{H2} is the partial pressure of hydrogen;

β is define in the paper (see equation 5 proposed by Lamb *et al*) as

$$\beta = \frac{1}{K_e} \left(\frac{P_{N_2}^{0.5} P_{H_2}^{1.5}}{P_{NH_3}} \right)$$

P_{N2} is the partial pressure of nitrogen; and

- 5 K_e is the equilibrium constant for the reaction (see equations 6 & 7 proposed by Lamb *et al*).
a/).

The Inventors have realized that equation 9 of Lamb *et al* may be extrapolated to temperatures outside 475°C to 600°C, e.g., in the range from 450°C to 700°C.

- 10 Transition metals that may be particularly suitable for use as the predominant catalytically active metals of the catalysts in the catalyst bed of the reactor tubes are selected from chromium, manganese, iron, cobalt, nickel, ruthenium and copper, e.g., iron, cobalt, nickel and ruthenium.

The catalytically active metal of the first (more active) catalyst is preferably ruthenium and the catalytically active metal of the second (less active) catalyst is typically nickel.

- 15 In preferred embodiments, the first catalyst in the upstream layer is a ruthenium-based catalyst and the second catalyst in the downstream layer is a nickel-based catalyst. Where a third catalyst is present in these embodiments, that catalyst is preferable a ruthenium-based catalyst although the catalyst may be a different ruthenium-based catalyst to the first catalyst, e.g., the catalyst have a different support and/or catalyst loading and/or, if both are promoted, then promoted with a different metal.

- 20 Thus, the term "ruthenium-based catalyst" refers to a catalyst containing ruthenium as the sole (or at least predominant) catalytically active metal, i.e., the metal responsible for catalyzing the cracking reaction. Ruthenium may be the only metal in the catalyst or alternatively one or more other metals may be present, e.g., in a material supporting the ruthenium. The terms "nickel-based catalyst" and "iron-based catalyst" are intended to be interpreted accordingly.

- 25 Suitable ruthenium-based catalysts and nickel-based catalysts may be supported, for example, on alumina (as disclosed in Lamb *et al* or Masel *et al*) or spinel (as disclosed in Boisen *et al*), optionally promoted with a Group 1 or Group 2 metal.

As mentioned above, water is usually present in the ammonia as a contaminant. The water may be removed from the ammonia in which case catalysts which do not tolerate water, *e.g.*, iron-based catalysts, may be used in the reactor tubes. However, water is not removed in preferred embodiments to save capital and operational costs and to reduce energy consumption. In these
5 embodiments, catalysts that cannot tolerate water, *e.g.*, iron-based catalysts, are not used. Instead, the catalysts in the reactor tubes would be able to tolerate up to 1 mol. % water in the ammonia feed. Such catalysts include nickel-based and ruthenium-based catalysts.

In preferred embodiments, a ruthenium-based catalyst is used in the first layer within each tube so that the faster reaction rate allows the metals temperature to be kept to within the design limit
10 of 660°C. In these embodiments, the second layer of the tubes contains a lower cost but less active nickel-based catalyst. The catalysts are layered in this way in the preferred embodiments to use the endotherm of reaction inside the tubes to keep the tube metal cool in the region where the combustion is most intense on the outside of the tubes. Since ruthenium is more catalytically active it creates a stronger endotherm that keeps the inner tube wall cooler in the region of high
15 ammonia concentration on the process side which protects the tubes from excessive nitriding caused by high ammonia concentration and high temperatures.

The combustion process in the furnace is preferably at least partially internally fueled, *i.e.*, at least part of the fuel is either ammonia or an off gas generated during recovery of hydrogen from the cracked gas or a mixture of the two. However, a trim fuel such as C₁ to C₃ hydrocarbons or natural
20 gas may be used as required although using a hydrocarbon trim fuel will increase the carbon intensity of the process. It is generally desirable, however, to minimize, or even eliminate, the use of such a trim fuel to reduce the carbon intensity of the process.

The oxidant gas is typically air but may be an oxygen-enriched gas or pure oxygen as appropriate.

The feed to the catalyst-filled reactor tubes of the furnace may be at a temperature up to about
25 800°C provided the material of the reactor walls can withstand the higher temperatures. For lower temperature cycles, the feed is typically at a temperature in a range from about 400°C to about 600°C, or from about 450°C to about 550°C, *e.g.*, at about 500°. For higher temperature cycles, the feed may be at a temperature in the range of about 500°C to about 800°C, or from about 600°C to about 700°C, *e.g.*, at about 650°.

The cracking temperature and pressure typically dictate that the ammonia slip in the reactor tubes is no more than 3 mol. %, *e.g.*, from about 0.5 mol. % to about 1.5 mol. %.

In some embodiments, the heated ammonia gas is partially cracked in an adiabatic reaction unit comprising at least one catalyst bed to produce partially cracked ammonia gas for feeding to the catalyst-filled reactor tubes.

The mole fraction of ammonia in the gas passing through the adiabatic reaction unit is typically reduced by at least 20%, e.g., by at least 25%, or by at least 30%, or even by at least 35%, e.g., by about 40%, and potentially up to about 50%. Put another way, the mole fraction of ammonia may be reduced from 1 (or almost 1) in the heated ammonia gas to an amount in a range from about 0.5 to about 0.8, or to an amount in a range from about 0.5 to about 0.7, or to an amount in a range from about 0.55 to about 0.65, e.g., to about 0.6, in the partially cracked ammonia gas.

The adiabatic reaction unit may be incorporated into the design of the process to improve overall efficiency, specifically by using the heat available in the flue gas to provide heat to the adiabatic cracking process within this unit. In this regard, the temperatures around the adiabatic reaction unit are typically optimized to maximise the recovery of heat from the flue gas while avoiding temperatures greater than about 660°C because of concerns regarding the materials.

The key design parameter for the adiabatic reaction unit is the inlet temperature. A higher inlet temperature allows greater conversion in the unit since the ammonia cracking reaction is endothermic. However, higher temperatures put greater demands on the materials of construction and the catalyst. The inlet temperature is typically in a range from about 350°C to about 800°C and, for lower temperature cycles, may be in a range from about 400°C to about 600°C, or from about 400°C to about 450°C. For higher temperature cycles, the inlet temperature may be in a range from about 500°C to about 700°C, or from about 550°C to about 650°C.

Due to the high temperature and ammonia concentration, the cracking reactor vessels, e.g., the adiabatic reactors and the reactor tubes in furnace, must typically be constructed from materials resistant to ammonia and/or nitriding. Suitable materials include nickel-based alloys comprising at least 40 wt. % or at least 50 wt. % nickel. Such alloys typically have no more than 90 wt. %, or no more than 80 wt. % nickel. The alloys will typically comprise one or more other metals selected from chromium, cobalt, molybdenum and iron.

Specific examples of suitable nickel-based alloys include UNS N06600, N06625, N06601, N06617, N06025, N06230, N07214, N08811. In some embodiments, austenitic nickel-chromium based superalloys such as Inconel may be used.

The unified numbering system (UNS) is an alloy designation system that is widely accepted in North America. Each UNS number relates to a specific metal or alloy and defines its specific chemical composition, or in some cases a specific mechanical or physical property.

Other suitable materials include cobalt-based alloys such as UNS R30188. In addition, high temperature alloys that are less resistant to ammonia nitriding, such as UNS N08811 or cast alloys such as HPNb, HP Micro-Alloyed, MA-1 (MetalTek International, USA), may be suitable, particularly when surface modified or coated by a corrosion resistant layer such as aluminization; aluminization and then pre-oxidation; or a ceramics coating. Nitriding resistant alloys can also be used with surface modification or coating for improved performance.

The adiabatic reaction unit will typically comprise one or more adiabatic reactors, the or each reactor comprising a catalyst bed. The or each adiabatic reactor may be made from one or more of the ammonia-resistant and/or ammonia nitriding-resistant materials list above.

In preferred embodiments, the adiabatic reaction unit will comprise two or more adiabatic reactors, e.g., two, three, four, five or six reactors, with interstage heating as appropriate. The reactors may be arranged in series or in parallel or in a combination of series and parallel depending on the requirements of the process. However, in preferred embodiments, the adiabatic reaction unit will have two such reactors arranged in series with interstage heating of the intermediate partially cracked ammonia gas by heat exchange against the cracked gas or the flue gas.

Each adiabatic reactor has a bed comprising at least one catalyst suitable for cracking ammonia. Any conventional ammonia cracking catalyst may be used in the bed of the or each adiabatic reactor. Suitable catalysts are discussed above in the context of the catalysts for the reactor tubes.

In embodiments where water is removed from the ammonia prior to cracking, catalysts that are not tolerant to water, e.g., iron-based catalysts, may be used in the adiabatic reactor(s). However, water is not removed in preferred embodiments to save capital and operational costs and to reduce energy consumption. In these embodiments, the catalyst beds of the adiabatic reaction unit contain no iron-based catalyst and, instead, the use of water tolerant catalysts, i.e., metal-based catalysts that are tolerant to the presence of up to 1 mol. % water, are preferred. In this regard, either nickel-based catalysts or ruthenium-based catalysts, or a combination of nickel-based catalysts and ruthenium-based catalysts may be used in the bed(s) of the adiabatic reactor unit.

As mentioned above, ruthenium-based catalysts tend to be more active than nickel-based catalysts but are more expensive. Therefore, further optimization is possible by selection of the catalyst and, if more than one type of catalyst is used, by the ordering of layers of the catalysts within the bed(s) of the adiabatic reaction unit.

- 5 In some preferred embodiments having two adiabatic reactors in series, the catalyst bed of the first reactor comprises, *e.g.*, contains or consists of, a single layer of a first catalyst, *e.g.*, a ruthenium-based catalyst, and the catalyst bed of the second reactor comprises, *e.g.*, contains or consists of, an upstream layer of a second catalyst, *e.g.*, a nickel-based catalyst, that is typically less active than the first catalyst, and a downstream layer of a third catalyst, *e.g.*, a ruthenium-
10 based catalyst, that is typically more active than the second catalyst.

In these embodiments, the first and third catalysts may be identical. Alternatively, the first and third catalysts may be different, *e.g.*, contain different catalytically active metals or contain the same catalytically active metals but on different supports or contain the same catalytically active metals on the same supports but at different loadings.

- 15 In other preferred embodiments having two adiabatic reactors in series, the catalyst bed of the first reactor comprises, *e.g.*, contains or consists of, a single layer of a first catalyst, *e.g.*, a nickel-based catalyst, and the catalyst bed of the second reactor comprises, *e.g.*, contains or consists of, an upstream layer of a second catalyst, *e.g.*, a nickel-based catalyst, that typically has a similar activity to the first catalyst, and a downstream layer of a third catalyst, *e.g.*, a ruthenium-based
20 catalyst, that is typically more active than the first and second catalysts.

In these embodiments, the first and second catalysts may be identical. Alternatively, the first and second catalysts may be different, *e.g.*, contain different catalytically active metals or contain the same catalytically active metals but on different supports or contain the same catalytically active metals on the same supports but at different loadings.

- 25 In both sets of preferred embodiments, the volume of the upstream layer of the second catalyst in the bed of the second reactor may be from about 40% to about 90%, *e.g.*, from about 50% to about 70% or about 60%, of the total volume of the bed. In the absence of another layer of catalyst, the volume of the downstream layer of the third catalyst in the bed of the second reactor may be from about 10% to about 60%, *e.g.*, from about 30% to about 50%, or about 40%, of the
30 total volume of the bed.

The Inventors have realized that ruthenium-based catalysts are not only sufficiently tolerant to water but are also capable of cracking hydrocarbon oils into shorter hydrocarbons such as methane, together with carbon monoxide and hydrogen. Therefore, the use of these catalysts in the adiabatic reaction unit can eliminate the need for a dedicated unit upstream for removing the oil from the liquid ammonia.

It is also known that at higher temperatures, catalyst sintering reduces the activity and life of the catalyst. In this regard, the skilled person would be aware of the need to balance the improved conversion against the higher vessel costs and shorter catalyst life.

The heat from the cracked gas and the flue gas is typically used to heat the feed streams to the adiabatic reaction unit and furnace, thereby reducing the overall energy consumed by the process. In this regard, the temperature of the cracked gas will depend on the cycle being operated.

In lower temperature cycles, the temperature of the cracked gas may be up to about 700°C, *e.g.*, typically from about 550°C to about 700°C, or from about 600°C to about 650°C. The temperature of the flue gas may be up to about 750°C at its highest point. However, the temperature drops due to heat leak and is typically from about 600°C to about 700°C at the point where its heat can be utilized effectively.

In higher temperature cycles, the temperature of the cracked gas may be up to about 750°C, *e.g.*, typically from about 650°C to about 750°C, or from about 675°C to about 725°C. The temperature of the flue gas may be up to about 840°C at its highest point. However, the temperature drops due to heat leak and is typically from about 700°C to about 800°C at the point where its heat can be utilized effectively.

Where an adiabatic reaction unit is used to partially crack the heated ammonia gas and the partially cracked gas is heated to the feed temperature of the catalyst-filled reactor tubes of the furnace, at least some of the duty required to heat the partially cracked gas is provided by heat exchange with the cracked gas. In preferred embodiments, the cracked gas is not used to heat another process fluid before heating the partially cracked gas. Some of this heating duty may be provided in another way, *e.g.*, by heat exchange with the flue gas. However, all of this heating duty is preferably provided by the cracked gas.

In preferred embodiments, the composition of the ammonia is typically at least substantially unchanged from the liquid ammonia in storage to the heated ammonia gas being fed to the

adiabatic reaction unit. Oil present in the liquid ammonia may be removed at some point prior to partial cracking of the ammonia although, where ruthenium-based catalysts are used, the oil does not need to be removed. However, water is not typically removed so any water present in the liquid ammonia will also be present in heated ammonia gas.

- 5 Where the partially cracked gas is heated by heat exchange with the cracked gas, the temperature of the cracked gas is reduced. The cooled cracked gas is then typically further cooled by providing at least part of the heating duty required to generate the heated ammonia gas from liquid ammonia. After cooling, hydrogen may be recovered from the cracked gas as product. Recovery may be achieved in pressure swing adsorption (PSA) process or by using one or more selectively
10 permeable membranes, or by way of a combination of PSA and membrane separation. In preferred embodiments, hydrogen recovery is achieved in a PSA process alone, *i.e.*, without use of membrane separation.

In embodiments using a PSA process, an off gas comprising nitrogen gas, residual ammonia and residual hydrogen is generated. This off gas may be pre-heated and used as fuel for the
15 combustion in the furnace. Alternatively, while part of the offgas may be used as fuel, another part may be compressed and returned to the PSA process to improve hydrogen recovery.

Aspects of the invention include:

#1. A process for cracking ammonia comprising:

providing a heated ammonia gas at super-atmospheric pressure;

- 20 combusting fuel with an oxidant gas in a furnace to heat catalyst-containing reactor tubes, each tube comprising an upstream layer of a first catalyst and a downstream layer of a second catalyst, and produce a flue gas;

feeding the heated ammonia gas, or a partially cracked ammonia gas derived therefrom, to the catalyst-containing reactor tubes to cause cracking of ammonia and produce a cracked gas
25 comprising hydrogen gas, nitrogen gas and residual ammonia gas,

wherein the first catalyst is more active for cracking ammonia than the second catalyst.

#2. A process according to #1 wherein the first catalyst is a ruthenium-based catalyst.

#3. A process according to #1 or #2 wherein the second catalyst is a nickel-based catalyst.

- #4. A process according to any of #1 to 3 wherein the catalyst-containing reactor tubes comprise a layer of a third catalyst downstream of the layer of the second catalyst, the third catalyst being more active for cracking ammonia than the second catalyst.
- #5. A process according to #4 wherein the third catalyst contains the same catalytically active metal as the first catalyst.
- #6. A process according to #4 or #5 wherein the third catalyst is a ruthenium-based catalyst.
- #7. A process according to any of #1 to #6 comprising:
- pumping liquid ammonia containing at least 0.1 mol. % water to produce pumped liquid ammonia;
- pre-heating the pumped liquid ammonia to produce a pre-heated liquid ammonia;
- vaporizing the pre-heated liquid ammonia to produce ammonia gas; and
- heating the ammonia gas to produce the heated ammonia gas at super-atmospheric pressure,
- wherein the water from the liquid ammonia is present in the heated ammonia gas.
- #8. A process according to #7 wherein at least some, preferably a majority, e.g., over 50% and optionally up to 80%, of the heating duty required to provide the heated ammonia gas is provided by heat exchange with the cracked gas.
- #9. A process according to #7 or #8 wherein the water is present in the heated ammonia gas in an amount of no more than 1 mol. %.
- #10. A process according to any of #1 to #9 wherein the catalyst-containing reactor tubes contain no iron-based catalyst.
- #11. A process according to any of #1 to #10 comprising partially cracking the heated ammonia gas in an adiabatic reaction unit comprising at least one catalyst bed to produce the partially cracked ammonia gas for feeding to the catalyst-filled reactor tubes.
- #12. A process according to #11 wherein the catalyst bed(s) of the adiabatic reaction unit comprise(s) at least one catalyst selected from a nickel-based catalyst and a ruthenium-based catalyst.

#13. A process according to #11 or #12 wherein the catalyst bed(s) of the adiabatic reaction unit contain(s) no iron-based catalyst.

#14. A furnace for cracking heated ammonia gas comprising:

a radiant section comprising at least one inlet for fuel and oxidant gas in fluid flow communication with at least one burner, an ammonia feed inlet, catalyst-containing reactor tubes having upstream ends in fluid flow communication with the ammonia feed inlet and downstream ends in fluid flow communication with an outlet for cracked gas, each tube comprising an upstream layer of a first catalyst and a downstream layer of a second catalyst; and

a convection section in fluid flow communication with the radiant section and comprising an outlet for flue gas;

wherein the first catalyst is more active for cracking ammonia than the second catalyst.

#15. A furnace according to #14 wherein the first catalyst is a ruthenium-based catalyst.

#16. A furnace according to #14 or #15 wherein the second catalyst is a nickel-based catalyst.

#17. A furnace according to any of #14 to #16 wherein each catalyst-containing reactor tube comprises a layer of a third catalyst downstream of the layer of the second catalyst, the third catalyst being more active for cracking ammonia than the second catalyst.

#18. A furnace according to #17 wherein the third catalyst contains the same catalytically active metal as the first catalyst.

#19. A furnace according to #17 or #18 wherein the third catalyst is a ruthenium-based catalyst.

#20. A furnace according to any of #14 to #19 wherein the catalyst-containing reactor tubes contain no iron-based catalyst.

#21. Apparatus for cracking heated ammonia gas comprising:

a source of liquid ammonia;

a pump in fluid flow communication with the source of liquid ammonia for pumping liquid ammonia; and

a furnace as defined in any of #14 to #20 wherein the ammonia feed inlet is in fluid flow communication with the pump,

wherein the apparatus further comprises:

at least one heat exchanger arranged for pre-heating liquid ammonia upstream of the pump; and

at least one heat exchanger arranged for vaporizing pumped liquid ammonia and heating ammonia gas by heat exchange with flue gas and/or cracked gas located between the pump and the ammonia feed inlet of the furnace.

#22. An apparatus according to #21 comprising an adiabatic reaction unit for partially cracking heated ammonia gas at super-atmospheric pressure, the unit comprising an inlet for heated ammonia gas at super-atmospheric pressure in fluid communication with the pump, at least one catalyst bed having an upstream end in fluid communication with the inlet and a downstream end in fluid communication with an outlet for partially cracked ammonia gas,

wherein the ammonia feed inlet of the furnace is in fluid flow communication with the outlet for partially cracked ammonia gas of the adiabatic reaction unit.

The invention will now be described by way of example only with reference to the figures.

In Fig. 1, a stream 2 of liquid ammonia at about -32°C is removed from storage (not shown) and fed to a pump P101 where it is pumped to produce a stream 4 of pressurised liquid ammonia at a pressure of about 46 bar which is pre-heated by heat exchange with a heat transfer fluid which in this case is an aqueous solution of a glycol, typically about 55 wt. % ethylene glycol or propylene glycol, in heat exchanger E271 to produce stream 6 of pre-heated liquid ammonia at about 55°C. An electric heater may be used to ensure that the temperature of the glycol solution being fed to the heat exchanger E271 is sufficient to pre-heat the liquid ammonia to the required temperature.

The pre-heated liquid ammonia in stream 6 is further heated by heat exchange in heat exchanger E312 to produce a stream 8 of further heated liquid ammonia. The further heated liquid ammonia in stream 8 is then evaporated by heat exchange in heat exchanger E311 to produce a stream 10 of ammonia vapor. The ammonia vapor in stream 8 is then superheated by heat exchange in heat exchanger E310 to produce a stream 12 of heated ammonia gas at about 260°C.

The heated ammonia gas in stream 12 is further heated by heat exchange in heat exchanger E2102 to produce a stream 14 of superheated ammonia gas at about 420°C. For convenience, heat exchanger E2102 is indicated as a single unit. However, there may be in reality two separate exchangers with a selective catalytic reactor (SCR) located in between.

The superheated ammonia gas in stream 14 is fed at about 420°C and about 43 bar to a first adiabatic reactor vessel C141 and passed through a bed of ruthenium-based catalyst. Some of

the ammonia gas is cracked over the catalyst to form a stream 16 of intermediate gas that contains some cracked ammonia. The mole fraction of ammonia in the gas passing through the first adiabatic reactor vessel C141 drops from almost 1 to about 0.9.

The intermediate gas is at about 360°C before being heated by heat exchange in heat exchanger E2103 to produce a stream 18 of superheated intermediate gas which is then fed at about 590°C to a second adiabatic reactor vessel C142 and passed through a bed containing an upstream layer of nickel-based catalyst and a downstream layer of a ruthenium-based catalyst to produce a stream 20 of partially cracked ammonia gas. The mole fraction of ammonia in the gas passing through the second adiabatic reactor vessel C142 drops from about 0.9 to about 0.6.

The catalyst bed of the second adiabatic reactor vessel C142 has two layers with a layer of nickel-based catalyst over a layer of ruthenium-based catalyst to use the heat more efficiently and hence maximize ammonia conversion. The catalyst volumes are also optimized, *i.e.*, minimizing the volume of the ruthenium-based catalyst by limiting the exit temperature of the second adiabatic reactor vessel to about 390°C. The Inventors found that reducing this temperature any further increases the volume required of the ruthenium-based catalyst.

The ruthenium-based catalyst is the same in both the first and second adiabatic reactor vessels. However, different ruthenium-based catalysts may be used.

The partially cracked ammonia in stream 20 is heated by heat exchange in a heat exchanger (or "economizer") E305 prior to being fed as stream 22 at a pressure of about 38 bar to catalyst-filled tubes in the radiant section F201 of a furnace (or reactor). Heating the feed to the tubes increases the amount of cracking that can be done with the heat from the burners by reducing the duty required to heat the partially cracked stream to the reaction temperature. Utilizing the cracked stream from the tubes allows efficient use of this high temperature stream. The inlet temperature of the direct fired tube furnace is limited to about 500°C to limit the inner wall temperature of the cracker tubes.

A stream 62 of air passed through forced draft fan K212 before being pre-heated by heat exchange in heat exchanger E2141 to produce a stream 64 of pre-heated air. The pre-heated air of stream 64 may be mixed with a stream 70 of natural gas as trim fuel and fed to the burners (not shown) of the furnace F201. Pre-heating the air in this way helps reduce the fuel requirement.

The tubes in the radiant section F201 of the furnace are filled with two types of ammonia cracking catalysts in two different layers. A ruthenium-based catalyst is used in the first layer within each

tube so that the faster reaction rate allows the metals temperature to be kept to within the design limit of about 660°C. The second layer in the tubes, downstream of the first layer, contains a lower cost but less active nickel-based catalyst.

5 A stream 24 of cracked gas exits the radiant section F201 of the direct fired tube furnace at about 640°C and is then fed to economizer E305 to provide the duty required to heat the partially cracked ammonia, thereby reducing the temperature of the cracked gas to about 530°C.

The economizer E305 is depicted as a shell-and-tube style heat exchanger with the partially cracked ammonia gas passing through the tubes and the cracked gas passing through the shell side. However, this arrangement could be reversed or indeed a different style of heat exchanger
10 could be used.

A stream 26 of cracked gas is then fed from the economizer E305 to heat exchanger E310 to provide the duty required to superheat the ammonia gas, thereby further reducing the temperature of the cracked gas to about 389°C.

15 A stream 28 of cracked gas is then fed from heat exchanger E310 to heat exchanger E311 to provide the duty required to evaporate the further heated liquid ammonia, thereby reducing the temperature of the cracked gas even further to about 109°C.

A stream 30 of cracked gas is then fed from heat exchanger E311 to heat exchanger E312 to provide the duty required to further heat the heated pressurized liquid ammonia, thereby reducing the temperature of the cracked gas further again to about 70°C.

20 Each of the heat exchangers E310, E311 and E312 is depicted as an individual shell-and-tube style heat exchanger with the ammonia passing through the tubes and the cracked gas passing through the shell side. However, this arrangement could be reversed. Alternatively, the heat exchangers could be combined into a single shell-and-tube style heat exchanger or indeed a different style of heat exchanger could be used.

25 A stream 32 of cracked gas from heat exchanger E312 is then cooled further by heat exchange with the heat transfer fluid in cooler E323 and then fed as stream 34 to PSA system U501 where it is separated into a stream 40 of hydrogen gas which is removed as product, and a stream 42 of PSA off gas comprising nitrogen gas, residual hydrogen gas and residual ammonia gas. The hydrogen gas in stream 40 may be fed to a hydrogen liquefaction unit (not shown) to produce
30 liquid hydrogen.

All of the PSA off gas in stream 42 may be sent directly as fuel (stream 60) for combustion in the furnace F201. Alternatively, stream 42 may be divided into two portions.

5 A first portion of the PSA off gas in stream 44 is heated by heat exchange in a heat exchanger E2112 to produce a stream 60 of warmed PSA off gas which is then fed to the burners in the furnace F201, together with air feed 64 and, optionally the natural gas feed 70 as required. A minimum amount of natural gas is used as trim fuel to provide the balance of fuel required in the fired section.

10 A second portion may be sent as stream 46 to a multistage compression unit K681 for compression. The compression unit K681 has five stages with an intercooler between each stage, together with an aftercooler following the last stage. Heat is recovered from the compressed gas in the intercoolers and the aftercooler by heat exchange with the heat transfer fluid. Heat may also be recovered from the lube oil and, where a positive displacement compression unit is used, from the cylinder(s) of the compression unit using the heat transfer fluid.

15 For convenience, the intercoolers and aftercooler are indicated by a single heat exchanger (labelled as E6816A-E) that recovers heat from a stream 48 of compressed PSA off gas by heat exchange with a stream 52 of the heat transfer fluid to produce a stream 50 of cooled compressed PSA off gas and a stream 54 of warmed heat transfer fluid.

20 The heat transfer fluid heated in the cooler E323 and in the intercoolers and aftercooler E6816A-E, *etc.*, is then used to provide the duty required to preheat the liquid ammonia by heat exchange in heat exchanger E271.

The cooled compressed PSA off gas in stream 50 is fed to a phase separator C6816 in which any condensate is removed as stream 56. The compressed PSA off gas is then recycled to the PSA system U501 as stream 58 to recover further hydrogen. In this way, the recovery of hydrogen may be increased from 85% (without the recycle) to 95% (with the recycle).

25 As indicated above, the process can be operated without compression unit K681, resulting in lower hydrogen recovery in the PSA unit 501. Reducing hydrogen recovery will obviously result in less hydrogen gas product. However, reduced hydrogen recovery may still be desirable as the carbon intensity (CI) for the process is lowered as more hydrogen would be in the off gas thereby reducing the requirement for natural gas as trim fuel and reducing carbon dioxide emissions.

30 A stream 72 of flue gas at about 686°C passes from the radiant section F201 to the convection section 90 of the furnace F201 where it first provides the duty required to heat the intermediate

gas from stream 16 in heat exchanger E2103 thereby reducing the temperature of the flue gas which is then used (as stream 74) to provide the duty required to further heat the heated ammonia gas from stream 12 in heat exchanger E2102 thereby reducing the temperature of the flue gas further. Therefore, the flue gas provides heating duty in a direction countercurrent to the flow of feed gas to the radiant section F201 of the direct fired tube furnace.

The cooled flue gas is then used (as stream 76) to provide the duty required to heat the air from stream 62 in heat exchanger E2142 thereby reducing the temperature of the flue gas further. The further cooled flue gas is then used (as stream 78) to provide the duty required to heat the PSA off gas from stream 44 in heat exchanger E2112 thereby cooling the flue gas further.

The cooled flue gas leaves the convection section 90 of the direct fired tube furnace F201 as stream 80 at a temperature of about 121°C, *i.e.*, above the condensation point of water, passes through an induced draft fan K211 and then leaves the process as stream 82. All of the practical energy has been extracted from the flue gas at this point which may now be vented to atmosphere, optionally after further processing if required depending on its composition.

Oil may be present in the liquid ammonia in an amount up to about 5 ppm from boil-off gas compressors (not shown) used with ammonia storage tanks (not shown), either at the location where the ammonia is produced, or onsite where the ammonia is cracked, or indeed anywhere in transit between the two sites. The presence of oil in the ammonia may cause difficulties as the ammonia cracking catalysts may not tolerate the oil. Thus, it may be desirable to remove the oil before the ammonia is exposed to the catalysts. Oil may be removed by passing the ammonia through a bed of activated carbon.

If oil is to be removed from the ammonia, then an oil removal unit (not shown) may be located in stream 2 (*i.e.*, in the feed line to the pump P101), in stream 4 (*i.e.*, between pump P101 and glycol heater E271), in stream 6 (*i.e.*, between glycol heater E271 and heat exchanger E312), in stream 8 (*i.e.*, between heat exchangers E312 and E311) or in stream 10 (*i.e.*, between heat exchangers E311 and E310).

EXAMPLES

Comparative Example

The process depicted in Fig. 1 has been simulated by computer (Aspen Plus, ver. 10, Aspen Technology, Inc., Massachusetts, USA) for a plant designed to produce 30 tonnes/day hydrogen (stream 40).

The reactor was simulated as having 60 tubes, each tube having an internal diameter of 4.313 in (0.11 m) and containing a 40 ft (12.2 m) catalytic bed containing a nickel-based catalyst as the sole catalyst.

The activities of the ruthenium-based catalyst in the adiabatic reactors and the nickel-based catalyst in both the adiabatic reactors and the reactor tubes were modelled using rate equation No. 9 given in Lamb *et al* (Int. J. Hydrogen Energy, 44 (2019) pp 3726-3736) as a basis. For the purpose of the simulation, it was assumed that the activity of the ruthenium-based catalyst conformed to the rate equation but that the activity of the nickel-based catalyst was 20% of that predicted by the rate equation.

Various temperatures and the mole fraction of ammonia are plotted as a function of cracker length in Fig. 2. The results indicate that the tubes have a maximum outer wall temperature of about 750°C and a maximum inner wall temperature of about 724°C.

Example 1

The simulation according to the Comparative Example was repeated using an arrangement of catalysts in the reactor tubes according to the present invention. In this regard, each tube was simulated as having a 15 ft (4.6 m) upstream layer of a ruthenium-based catalyst and a 25 ft (7.6 m) downstream layer of the same nickel-based catalyst as in the Comparative Example.

All other features of the simulation were unchanged, including the arrangement of catalysts in the beds of the adiabatic reactors, and the assumed activities of both types of catalyst.

The various temperatures and ammonia mole fraction are plotted as a function of cracker length in Fig. 3. The results indicate that the tubes have a maximum outer wall temperature of 668°C and a maximum inner wall temperature of 648°C, both of which are significantly less than the equivalent figures of 750°C and 724°C respectively from the Comparative Example.

At the lower temperatures, the extent of nitriding of the metal from which the tubes are formed will be significantly less.

The heat and mass balance data for the simulation of Example 1 is provided in Table 2.

	2	4	6	12	14	16	18	20	22	24
Composition, mol%										
Hydrogen	0.00	0.00	0.00	0.00	0.00	7.75	7.75	29.83	29.83	73.93
Nitrogen	0.00	0.00	0.00	0.00	0.00	2.58	2.58	9.94	9.94	24.64
Ammonia	99.81	99.81	99.81	99.81	99.81	89.49	89.49	60.07	60.07	1.33
Water	0.19	0.19	0.19	0.19	0.19	0.18	0.18	0.15	0.15	0.10
Oxygen	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ethane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Pentane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Butane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Pentane	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ethylene Glycol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Flowrate, kmol/hr	447.7	447.7	447.7	447.7	447.7	472.1	472.1	558.9	558.9	882.9
Flowrate, kg/hr	7626.4	7626.4	7626.4	7626.4	7626.4	7626.4	7626.4	7626.4	7626.4	7626.4
Pressure, bar (a)	1.0	46.4	46.0	44.8	42.8	41.3	40.3	38.8	38.4	35.4
Temperature, °C	-33.7	-32.2	53.0	260.0	420.0	357.0	590.0	387.0	500.0	639.9
	26	34	44	40	60	70	64	72	76	82
Composition, mol%										
Hydrogen	73.93	73.93	12.42	100.00	12.42	0.00	0.00	0.00	0.00	0.00
Nitrogen	24.64	24.64	82.78	0.00	82.78	1.00	77.27	78.53	78.53	78.53
Ammonia	1.33	1.33	4.48	0.00	4.48	0.00	0.00	0.00	0.00	0.00
Water	0.10	0.10	0.32	0.00	0.32	0.00	1.05	15.69	15.69	15.69
Oxygen	0.00	0.00	0.00	0.00	0.00	0.00	20.73	1.15	1.15	1.15
Argon	0.00	0.00	0.00	0.00	0.00	0.00	0.92	0.56	0.56	0.56
Carbon Dioxide	0.00	0.00	0.00	0.00	0.00	0.50	0.03	4.07	4.07	4.07
Methane	0.00	0.00	0.00	0.00	0.00	93.96	0.00	0.00	0.00	0.00
Ethane	0.00	0.00	0.00	0.00	0.00	4.20	0.00	0.00	0.00	0.00
Pentane	0.00	0.00	0.00	0.00	0.00	0.30	0.00	0.00	0.00	0.00
Butane	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00
Pentane	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00
Ethylene Glycol	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Flowrate, kmol/hr	882.9	882.9	262.8	620.0	262.8	27.8	435.1	713.1	713.1	713.1
Flowrate, kg/hr	7626.4	7626.4	6376.4	1250.0	6376.4	472.7	12553.2	19402.2	19402.2	19402.2
Pressure, bar (a)	35.0	33.4	1.4	32.7	1.2	5.0	1.1	1.0	1.0	1.0
Temperature, °C	530.1	50.0	35.0	50.0	138.0	30.0	236.7	685.9	304.4	131.8

TABLE 2

The data indicate that, with a 1.33 mol. % ammonia slip (stream 24) from the cracker (tube furnace 201) and 95% recovery of hydrogen in the PSA unit, 7626 kg/h of ammonia is required (stream 2) as feed, together with 472.7 kg/h of natural gas (stream 70) as fuel in addition to the PSA off gas (stream 60) in order to fire the cracker.

- 5 For a given hydrogen recovery and ammonia slip, the effect of using the cracked gas (instead of the flue gas) to provide the duty required to heat the partially cracked gas (stream 20) to the feed temperature of the catalyst-filled tubes of the cracker (F201) is to reduce the overall carbon intensity (CI) process.

10 While the invention has been described with reference to the preferred embodiments depicted in the figure, it will be appreciated that various modifications are possible within the spirit or scope of the invention as defined in the following claims.

In this specification, unless expressly otherwise indicated, the word "or" is used in the sense of an operator that returns a true value when either or both of the stated conditions are met, as opposed to the operator "exclusive or" which requires only that one of the conditions is met. The word "comprising" is used in the sense of "including" and incorporates "consisting of" rather than meaning "consisting of" exclusively.

20 All prior teachings above are hereby incorporated herein by reference. No acknowledgement of any prior published document herein should be taken to be an admission or representation that the teaching thereof was common general knowledge in Australia or elsewhere at the date thereof.

CLAIMS

1. A process for cracking ammonia comprising:
 - providing a heated ammonia gas at super-atmospheric pressure;
 - combusting fuel with an oxidant gas in a furnace to heat catalyst-containing reactor tubes,
 - 5 each tube comprising an upstream layer of a first catalyst and a downstream layer of a second catalyst, and produce a flue gas;
 - feeding the heated ammonia gas, or a partially cracked ammonia gas derived therefrom, to the catalyst-containing reactor tubes to cause cracking of ammonia and produce a cracked gas comprising hydrogen gas, nitrogen gas and residual ammonia gas,
 - 10 wherein the first catalyst is more active for cracking ammonia than the second catalyst.
2. The process of Claim 1 wherein the first catalyst is a ruthenium-based catalyst.
3. The process of Claim 1 wherein the second catalyst is a nickel-based catalyst.
4. The process of Claim 1 wherein the catalyst-containing reactor tubes comprise a layer of a third catalyst downstream of the layer of the second catalyst, the third catalyst being more active
- 15 for cracking ammonia than the second catalyst.
5. The process of Claim 4 wherein the third catalyst contains the same catalytically active metal as the first catalyst.
6. The process of Claim 4 wherein the third catalyst is a ruthenium-based catalyst.
7. The process of Claim 1 comprising:
 - 20 pumping liquid ammonia containing at least 0.1 mol. % water to produce pumped liquid ammonia;
 - pre-heating the pumped liquid ammonia to produce a pre-heated liquid ammonia;
 - vaporizing the pre-heated liquid ammonia to produce ammonia gas; and
 - heating the ammonia gas to produce the heated ammonia gas at super-atmospheric
 - 25 pressure,
 - wherein the water from the liquid ammonia is present in the heated ammonia gas.

8. The process of Claim 7 wherein at least some of the heating duty required to provide the heated ammonia gas is provided by heat exchange with the cracked gas.
9. The process of Claim 7 wherein the water is present in the heated ammonia gas in an amount of no more than 1 mol. %.
- 5 10. The process of Claim 1 wherein the catalyst-containing reactor tubes contain no iron-based catalyst.
11. The process of Claim 1 comprising partially cracking the heated ammonia gas in an adiabatic reaction unit comprising at least one catalyst bed to produce the partially cracked ammonia gas for feeding to the catalyst-filled reactor tubes.
- 10 12. The process of Claim 11 wherein the catalyst bed(s) of the adiabatic reaction unit comprise(s) at least one catalyst selected from a nickel-based catalyst and a ruthenium-based catalyst.
13. The process of Claim 11 wherein the catalyst bed(s) of the adiabatic reaction unit contain(s) no iron-based catalyst.
- 15 14. A furnace for cracking heated ammonia gas comprising:
- a radiant section comprising at least one inlet for fuel and oxidant gas in fluid flow communication with at least one burner, an ammonia feed inlet, catalyst-containing reactor tubes having upstream ends in fluid flow communication with the ammonia feed inlet and downstream ends in fluid flow communication with an outlet for cracked gas, each tube comprising an upstream
- 20 layer of a first catalyst and a downstream layer of a second catalyst; and
- a convection section in fluid flow communication with the radiant section and comprising an outlet for flue gas;
- wherein the first catalyst is more active for cracking ammonia than the second catalyst.
15. The furnace of Claim 14 wherein the first catalyst is a ruthenium-based catalyst.
- 25 16. The furnace of Claim 14 wherein the second catalyst is a nickel-based catalyst.
17. The furnace of Claim 14 wherein each catalyst-containing reactor tube comprises a layer of a third catalyst downstream of the layer of the second catalyst, the third catalyst being more active for cracking ammonia than the second catalyst.

18. The furnace of Claim 17 wherein the third catalyst has the same catalytically active metal as the first catalyst.

19. The furnace of Claim 17 wherein the third catalyst is a ruthenium-based catalyst.

20. The furnace of Claim 14 wherein the catalyst-containing reactor tubes contain no iron-based catalyst.

21. Apparatus for cracking heated ammonia gas comprising:

a source of liquid ammonia;

a pump in fluid flow communication with the source of liquid ammonia for pumping liquid ammonia; and

a furnace as defined in Claim 14 wherein the ammonia feed inlet is in fluid flow communication with the pump,

wherein the apparatus further comprises:

at least one heat exchanger arranged for pre-heating liquid ammonia upstream of the pump; and

at least one heat exchanger arranged for vaporizing pumped liquid ammonia and heating ammonia gas by heat exchange with flue gas and/or cracked gas located between the pump and the ammonia feed inlet of the furnace.

22. The apparatus of Claim 21 comprising an adiabatic reaction unit for partially cracking heated ammonia gas at super-atmospheric pressure, the unit comprising an inlet for heated ammonia gas at super-atmospheric pressure in fluid communication with the pump, at least one catalyst bed having an upstream end in fluid communication with the inlet and a downstream end in fluid communication with an outlet for partially cracked ammonia gas,

wherein the ammonia feed inlet of the furnace is in fluid flow communication with the outlet for partially cracked ammonia gas of the adiabatic reaction unit.