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(54) Titre : PROCEDE POUR ELIMINER LE N_2O ET LES NO_x DU PROCEDE POUR LA PREPARATION D'ACIDE NITRIQUE ET INSTALLATION APPROPRIEE A CETTE FIN

(54) Title: METHOD FOR REMOVING N_2O AND NO_x FROM THE NITRIC ACID PRODUCTION PROCESS, AND AN INSTALLATION SUITABLE FOR SAME

(57) Abrégé/Abstract:
The invention relates to a method and installation for producing nitric acid by catalytically oxidising NH_3 with oxygen and then reacting the obtained NO_x with an absorption agent in an absorption tower, said tower comprising a catalyst bed for N_2O decomposition arranged in the process gas after the catalytic NH_3 oxidation and before the absorption tower in the direction of flow, and a catalyst bed for NO_x reduction and further N_2O reduction, arranged in the residual gas after the absorption tower in the direction of flow. The method and installation allow N_2O and NO_x emissions from nitric acid installations to be reduced in a particularly efficient manner.

Abstract

The invention relates to a method and installation for producing nitric acid by catalytically oxidising NH_3 with oxygen and then reacting the obtained NO_x with an absorption agent in an absorption tower, said tower comprising a catalyst bed for N_2O decomposition arranged in the process gas after the catalytic NH_3 oxidation and before the absorption tower in the direction of flow, and a catalyst bed for NO_x reduction and further N_2O reduction, arranged in the residual gas after the absorption tower in the direction of flow. The method and installation allow N_2O and NO_x emissions from nitric acid installations to be reduced in a particularly efficient manner.

Method for removing N_2O and NO_x from the nitric acid production process, and an installation suitable for same

Description

The invention relates to a method of removing N_2O and NO_x from the process for nitric acid production and also a suitable plant for carrying out this process.

The production of nitric acid is, on an industrial scale, generally carried out by the Ostwald process by catalytic oxidation of ammonia (NH_3) over Pt/Rh catalysts. Here, NH_3 is selectively oxidized to nitrogen monoxide (NO) which is then oxidized during the course of the further process to nitrogen dioxide (NO_2) and finally reacted with water in an absorption tower to form nitric acid. The Pt/Rh catalysts are configured as thin gauzes clamped on a wide area in a burner. A gas mixture composed of typically about 8-12% by volume of ammonia and air is passed through the gauzes, with a temperature of about 850-950°C being established at the gauzes due to the exothermic nature of the reaction.

An overview of the procedure for nitric acid production and its various process variants is given, for example, in Ullmans Encyclopedia of Industrial Chemistry, Vol. A 17, VCH Weinheim (1991) or in Winnacker-Küchler, Chemische Technik, Prozesse und Produkte, 5th edition, volume 3, Anorganische Grundstoffe, Zwischenprodukte, Chemische Technik, Dittmeyer, R. / Keim, W. / Kreysa, G. / Oberholz, A. (editors), Wiley-VCH, Weinheim, (2005).

Unfortunately, however, the oxidation of NH_3 to NO is not 100% selective but a certain proportion of nitrogen (N_2) and nitrous oxide (N_2O) is always also formed in addition to the desired NO.

Depending on the oxidation conditions, i.e. prevailing pressure, temperature and inflow velocity to the NH_3 combustion and also type and state of ageing of the Pt/Rh gauze catalysts, about 4-15 kg of N_2O are typically formed per metric ton of HNO_3. This results in typical N_2O concentrations of from about 500 to 2000 ppmv in the process gas.

The N_2O formed is not absorbed when the process gas is fed into the absorption tower and thus goes into the tailgas of HNO_3 production. Since the deNO_x stages installed here for reducing the residual content of NO and NO_2 (together referred to as NO_x) also generally do not bring about a reduction in the N_2O content, the N_2O finally goes more or less undiminished into the atmosphere. For example, the tailgas from a nitric acid plant in which the oxidation of NH_3 is carried out at intermediate pressure (about 4-5 bar abs) contains on average about 1000 ppmv of N_2O, which corresponds to an N_2O concentration in the process gas downstream of the NH_3 oxidation of about 830 ppmv.

While NO and NO_2 have long been known as compounds having ecotoxic relevance (acid rain, smog formation) and limit values for NO_x emissions and technical measures for reducing their amounts have become established worldwide, nitrous oxide has become a focus of environmental concern only in the last decade since it contributes to a not inconsiderable extent to the degradation of stratospheric ozone and to the greenhouse effect. A variety of solutions for removing N_2O, partly in combination with new processes for NO_x reduction have therefore been developed in recent years for the nitric acid process and employed in industrial plants for the production of nitric acid.

An overview of various measures for reducing the amounts of N_2O and NO_x in the HNO_3 process is given, for example, in: J. Perez-Ramirez et al., "Formation and control of N_2O in nitric acid production – Where do we stand today?" Appl. Catal. B Environmental 2003, 44 (2), 117-151, in M. Schwefer, R. Maurer, M. Groves, "Reduction of Nitrous Oxide Emissions from Nitric Acid Plants" Nitrogen 2000 International Conference, Vienna, Austria, March 2000, or in Integrated Pollution Prevention and Control Reference Document on Best Available Techniques for the Manufacture of Large Volume Inorganic Chemicals – Ammonia, Acids and Fertilisers, European Commission August 2007.

For the removal of N_2O alone, secondary measures which are directed at decomposition of N_2O in the process gas of HNO_3 production are frequently used. Here, specific catalysts are installed directly downstream of the NH_3 combustion underneath the Pt/Rh gauze catalysts. The process

gas here has temperatures of about 900°C, so that N_2O here requires only a little catalytic activation to decompose it. The aim of a secondary measure is to achieve very high degrees of removal of N_2O. An N_2O removal of > 80%, often even > 90%, is typically achieved. At an average amount of N_2O formed of 830 ppmv, which is typical, i.e. average, for an intermediate pressure plant; this corresponds to residual N_2O concentrations of < 165 ppmv, in particular < 80 ppmv, in the process gas or < 200 ppmv, in particular < 100 ppmv, in the tailgas of HNO_3 production. However, degrees of removal of > 95% cannot be achieved by means of this technology since the space available for accommodating the secondary catalyst underneath the Pt/Rh gauze catalysts is limited.

However, the secondary measure offers the advantage of universal applicability, usually simple installation and a small catalyst requirement. In the ideal case, only replacement of packing elements which are often arranged underneath the gauze packings for flow equalization by the secondary catalyst is necessary, so that no additional apparatus costs are incurred. Particularly in the case of retrofitting, this is a clear advantage over N_2O removal from the tailgas of HNO_3 production (known as tertiary measure).

A disadvantage of secondary measures is, however, that owing to the limited space underneath the catalyst gauzes, a correspondingly finely divided catalyst having a high geometric surface area has to be used in order to achieve high degrees of removal of N_2O. This is associated with a correspondingly high pressure drop, which is ultimately reflected in a reduced production output of the HNO_3 plant. In addition, there is the risk that an only imprecisely definable loss of product can occur since the catalyst can, at 900°C, decompose not only N_2O but also NO to an unknown extent.

To remove NO_x from the tailgas of HNO_3 production, classical SCR catalysts based on TiO_2/V_2O_5 are usually employed in nitric acid plants (cf., for example, G. Ertl, H. Knözinger, J. Weitkamp: Handbook of Heterogeneous Catalysis, vol. 4, pages 1633-1668, VCH Weinheim (1997)). These operate in a temperature range from about 150 to 450°C and on an industrial scale are preferably operated in the range from 200 to

400°C, in particular from 250 to 350°C. With appropriate dimensioning of the catalyst beds, removal of NO_x down to residual concentrations of 40 ppm of NO_x, in special cases down to 20 ppm of NO_x, can be achieved in this way. In many nitric acid plants, such SCR catalysts are operated in the tailgas in combination with a secondary measure, i.e. together with N_2O removal in the process gas.

With regard to NO_x removal in the tailgas from HNO_3 production, iron-loaded zeolite catalysts also appear to be particularly advantageous since these also enable, unlike classical deNO_x catalysts based on TiO_2/V_2O_5, a certain proportion of N_2O to be removed at the same time, depending on the temperature. This is, for example, known from the disclosures in DE 101 12 444 A1 and in DE 102 15 605 A. In DE 101 12 444 A1, a gas containing N_2O and NO_x is firstly mixed with a gaseous reducing agent for NO_x, preferably with NH_3, and subsequently passed over the catalyst at a space velocity to be selected over the catalyst for the simultaneous removal of N_2O (by decomposition) and NO_x (by reduction) at a temperature of less than 450°C. In DE 102 15 605 A, the gas containing N_2O and NO_x is firstly mixed with ammonia as reducing agent for NO_x and additionally with hydrocarbons or carbon monoxide and/or hydrogen as reducing agent for N_2O and subsequently passed over iron-loaded zeolites for the removal of N_2O and NO_x, in each case by reduction, at a temperature of less than 450°C. A prerequisite for effective reduction of the N_2O in this process is complete reduction of NO_x. The removal of N_2O in the tailgas from HNO_3 production is referred to as tertiary measure.

Various possible ways of avoiding N_2O and NO_x emissions in nitric acid plants have thus been known to those skilled in the art from the prior art. Here, the abovementioned secondary and tertiary measures for removal of N_2O are competing technologies. A combination of these measures for removal of N_2O has hitherto not been realized on an industrial scale for cost reasons. In “Remarks and Comments on Nitric Acid Production Project Protocol — Public Draft Version 1.0 October 2009” by Groves and Rieck, it is merely mentioned that a secondary measure having poor removal performance could be supported by a tertiary measure in

order then to achieve an overall high degree of removal of N_2O. It is not stated how the coupling of these measures should be configured, for example whether the tertiary measure is a catalytic decomposition or reduction of N_2O or whether the removal of N_2O could be coupled with a $deNO_x$ stage or which devices or apparatuses could advantageously be used.

The process steps of, firstly, catalytic N_2O removal in the process gas and, secondly, N_2O and NO_x removal in the tailgas gives iron-loaded zeolite catalysts are to be combined with one another for the first time according to the present invention. Since the iron-loaded zeolite catalysts also have, as mentioned above, catalytic activity for N_2O decomposition or N_2O reduction, a further N_2O removal could be achieved in parallel to the NO_x reduction. Since less catalyst is generally required for the removal of N_2O in the process gas at high temperatures (850-950°C) than for removal of N_2O in the tailgas at low temperatures (T = < 500°C), it first appears to be advantageous from a technical and economic point of view to realize a very high degree of removal of N_2O by means of secondary measures and to use the iron-loaded zeolite catalysts in the tailgas more or less exclusively for the reduction of NO_x. It is known that the catalytic reduction of NO_x requires a significantly lower catalyst volume compared to the catalytic removal of N_2O. A person skilled in the art would therefore have, proceeding from the suggestions in the prior art, removed the N_2O largely in the process gas downstream of the ammonia oxidation and before introduction into the absorption tower and freed the resulting tailgas, which then would have had predominantly NO_x in terms of the nitrogen oxides, by selective catalytic reduction by means of ammonia in a $deNO_x$ stage based on Fe-zeolite catalysts downstream of the absorption tower.

In the implementation of this concept, it was, however, surprisingly found that the removal power of the $deNO_x$ stage was considerably reduced by the additional installation of a secondary measure. It was surprisingly found that at a very low content of N_2O, i.e. when a very high N_2O removal is achieved by the secondary measure, ammonia breakthrough occurs in the offgas stream from the NO_x reduction by means of NH_3 in the $deNO_x$ stage downstream of the absorption tower with increasing degrees of reduction of NO_x to values of from < 40 to < 3 ppmv, depending on the space velocity

selected over the catalyst. Since NH_3 is a very (eco)toxic compound, this is an extremely undesirable effect. Strict limits are imposed in many countries for NH_3 emissions or for NH_3 breakthrough from deNO_x plants. In addition, NH_3 breakthrough can in the presence of residual NO_x lead to formation of ammonium nitrate which can deposit in cooler parts of the plant. This must be avoided at all costs from a safety point of view since ammonium nitrate is an explosive substance.

It is therefore an object of the present invention to provide a process and a plant suitable therefor which ensures, by means of a combination of a secondary catalyst in the process gas stream with a deNO_x stage containing an iron-loaded zeolite catalyst in the tailgas stream, both a high degree of removal of NO_x and also of N_2O without ammonia breakthrough in the resulting offgas stream occurring.

The pressure drop generated by the two catalyst stages should lead to no significant impairment of the possible throughput in HNO_3 production or to a deterioration in the economics of the process.

The object is achieved by a process for preparing nitric acid by catalytic oxidation of NH_3 by means of oxygen and subsequent reaction of the NO_x formed with an absorption medium, preferably with water, in an absorption tower, which comprises a catalyst bed for N_2O decomposition arranged in the process gas, i.e. in the flow direction downstream of the catalytic NH_3 oxidation and upstream of the absorption tower, and a catalyst bed for NO_x reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas, i.e. in the flow direction downstream of the absorption tower,

- wherein the amount of N_2O removed in the catalyst bed for N_2O removal arranged in the process gas is not more than that which results in an N_2O content of > 100 ppmv, preferably > 200 ppmv, particularly preferably > 300 ppmv and very particularly preferably from > 300 to 1000 ppmv, and a molar N_2O/NO_x ratio of > 0.25, preferably > 0.5, before entry of the tailgas into the catalyst bed for NO_x reduction and
- the catalyst bed for NO_x reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas contains at least one iron-

loaded zeolite catalyst and

- NH_3 is added to the tailgas before entry into the catalyst bed in such an amount that an NO_x concentration of < 40 ppmv, preferably < 20 ppmv, particularly preferably < 10 ppmv, very particularly preferably < 5 ppmv, in particular < 3 ppmv or extremely preferably from < 3 to 0 ppmv, results at the outlet from the catalyst bed and
- the operating parameters are selected in such a way that an N_2O concentration of < 200 ppmv, preferably < 100 ppmv, particularly preferably < 50 ppmv, very particularly preferably < 30 ppmv and extremely preferably from < 30 to 0 ppmv results.

The abovementioned removal of N_2O according to the invention in the catalyst bed in the process gas to resulting residual concentrations of > 100 ppmv, preferably > 200 ppmv, of N_2O, particularly preferably > 300 ppmv and very particularly preferably from > 300 to 1000 ppmv of N_2O, refers to the tailgas concentration directly before entry into the catalyst bed for NO_x reduction downstream of the absorption tower.

To achieve such tailgas concentrations before entry into the catalyst bed for NO_x reduction, a reduction of the N_2O content to values of > 83 ppmv, preferably > 165 ppmv, of N_2O, particularly preferably > 250 ppmv and very particularly preferably from > 250 to 1200 ppmv of N_2O, have to be achieved in the upstream catalyst bed arranged in the process gas, as long as no further decrease in the N_2O content is effected by any measures or reaction stages upstream of the catalyst bed for NO_x reduction in the tailgas.

Depending on the actual amount of N_2O formed in the NH_3 oxidation, typically amounts in the range from 500 to 2000 ppmv, the N_2O removal according to the invention in the catalyst bed in the process gas is 40-90%, preferably 45-80%, particularly preferably 50-70%, based on the amount of N_2O initially present.

In an advantageous embodiment of the process of the invention, the targeted setting of the N_2O removal in the catalyst bed in the process gas (secondary catalyst) is achieved by variation of the layer thickness or bed height of the catalyst bed and/or selection of the catalyst material and/or

selection of the geometry of the catalyst material.

As catalyst materials, it is possible to use, in particular, materials which are known per se for the high-temperature decomposition of N_2O.

To be able to work in the long term at the high temperatures required of typically from 800°C to 1000°C, the catalysts have to have a high thermal stability. For this reason, particularly suitable catalysts are, for example, high-temperature-resistant ceramic catalysts which contain a high-temperature-resistant ceramic material which can itself have catalytic properties and/or serves as support for one or more active components. The catalytically active component can be distributed homogeneously in the ceramic matrix or be present as a layer applied to the surface.

Suitable active components are noble metals, e.g. of the platinum group, and also, in particular, transition metal oxides and/or mixed oxides containing transition metal, preferably those having a perovskite structure, a perovskite-like structure or a spinel structure, as are described, for example, in (N. Gunasekaran et al., Catal. Lett. (1995) 34, (3, 4), pp. 373-382). The use of cobalt-containing oxides or mixed oxides, e.g. Co_2O_4 or $LaCoO_3$, is particularly advantageous.

Particular preference is given to using catalysts having a porous support composed of polycrystalline or vitreous inorganic material, a cerium oxide functional layer applied thereto and a layer of oxidic cobalt-containing material applied thereto. Variations thereof are disclosed in DE 10 2007 038 711 A1.

Further suitable catalyst materials are also described, for example, in EP 2 184 105, EP 1 301 275 or DE1984895.

The catalyst materials can be produced as shaped bodies of any size and geometry by shaping methods known in ceramic processing, e.g. dry pressing, granulation or extrusion.

The shape and the size or the equivalent diameter of the shaped bodies is

selected so that the desired N_2O removal is achieved at a very low pressure drop over the catalyst bed or packing when using the selected amount of catalyst.

Preferred geometries of the shaped bodies are cylinders, hollow cylinders, multi-hole cylinders, perforated and unperforated trilobes or polylobes or honeycomb structures.

The lower limits to the equivalent diameter of the shaped catalyst bodies is, according to the invention, typically > 1.5 mm, preferably > 3 mm and in particular > 5 mm, and the upper limit to the equivalent diameter is typically < 20 mm, preferably < 15 mm and in particular < 10 mm.

The equivalent diameter of a body or a particle is the diameter of a sphere having the same volume to surface area ratio as the particle. It can be calculated by the formula

$$d_e = 6V/A$$

where V = volume of the particle and A = surface area of the particle.

The pressure drop over the bed or packing of the shaped catalyst body is generally < 30 mbar, preferably < 25 mbar, particularly preferably < 20 mbar, very particularly preferably < 15 mbar, in particular < 10 mbar.

The bed or packing height of the catalyst bed in the process gas (secondary catalyst) is usually 3-30 cm, preferably 5-20 cm, particularly preferably 10-20 cm.

After passage through the secondary catalyst and subsequent cooling, the process gas is fed into the absorption tower of the HNO_3 plant. Here, the NO_x formed is reacted with H_2O to form nitric acid and leave a tailgas which, depending on the dimensions of the absorption tower and on the prevailing pressure and temperature at the outlet of the absorption tower, has a residual content of about 200-2000 ppmv of NO_x and an N_2O content of > 100 ppmv, preferably > 200 ppmv, particularly preferably > 300 ppmv and very particularly preferably from > 300 to 1000 ppmv.

After stepwise heating of the tailgas, this is then passed through a catalyst bed containing at least one iron-loaded zeolite catalyst to effect NO_x reduction and a further decrease in the content of N_2O. Here, NH_3 is added to the tailgas for NO_x reduction before entry into the catalyst bed in such an amount that an NO_x concentration of < 40 ppmv, preferably < 20 ppmv, particularly preferably < 10 ppmv, very particularly preferably < 5 ppmv, in particular < 3 ppmv or extremely particularly preferably from < 3 to 0 ppmv, results at the outlet of the catalyst bed.

The other operating parameters such as temperature, pressure and space velocity and/or any addition of specific reducing agents for N_2O are selected so that an N_2O concentration of < 200 ppmv, preferably < 100 ppmv, very particularly preferably < 50 ppmv, in particular < 30 ppmv and extremely preferably from < 30 to 0 ppmv, results.

The decrease in the content of N_2O in the catalyst bed arranged in the tailgas is typically at least 50%, preferably at least 70%, particularly preferably at least 80% and very particularly preferably from 90 to 100%, based on the content of N_2O at the inlet into this catalyst bed. This degree of removal can be achieved by appropriate setting of the abovementioned operating parameters and/or by addition of specific reducing agents for N_2O, preferably hydrocarbons. The measures and also the dimensioning of the catalyst bed in order to achieve this degree of removal are known to those skilled in the art.

For the purposes of the present description, the term space velocity refers to the volume of gas mixture (measured at 0°C and 1.014 bara) per hour divided by the volume of catalyst. The space velocity can thus be adjusted via the volume flow of the gas and/or via the amount of catalyst.

The tailgas is usually passed through the catalyst bed at a space velocity of from 200 to 200 000 h^{-1}, preferably from 5 000 to 100 000 h^{-1}, in particular from 5 000 to 50 000 h^{-1}. The pressure in the tailgas before entry into the catalyst bed is generally from 1 to 50 bar, preferably at least 2 bar, in particular at least 3 bar, very particularly preferably from 4 to 25 bar. The temperature of the tailgas before entry into the catalyst bed is generally

300-600°C, preferably 330-520°C.

The setting of the abovementioned parameters is known to those skilled in the art, for example from DE 101 12 444 A1 or from DE 102 15 605 A.

In DE 101 12 444 A1, the removal of N_2O is effected over Fe-zeolite catalysts of the Fe-ZSM-5 type by pure decomposition which is catalyzed by residual NO_x. For reduction of the NO_x, an amount of from 0.9 to 1.3 mol of NH_3 for a mol of NO_x to be reduced, in particular from 1.0 to 1.2 mol of NH_3 per mol of NO_x to be reduced, is added to the tailgas stream. This addition of NH_3 can be applied directly to the process of the invention.

In an embodiment as per DE 102 15 605 A, the N_2O removal is achieved over Fe-zeolite catalysts of the Fe-BEA type by addition of appropriate reducing agents for N_2O, preferably hydrocarbons such as methane or propane. The amount of hydrocarbon (HC) required is about 0.2-1 mol of HC/1 mol of N_2O at the inlet into the catalyst bed. Preference is given to amounts of 0.2-0.7 mol of HC/1 mol of N_2O, in particular 0.2-0.7 mol of HC/1 mol of N_2O. The NO_x content is in this case to be reduced completely, i.e. to values of < 10 ppmv, preferably < 5 ppmv, in particular < 1 ppmv. The addition of appropriate amounts of nitrogen-containing reducing agents is necessary here. In the case of NH_3, these are, based on the NO_x entry concentration, about 1-2 mol of NH_3/mol of NO_x, preferably 1.2-1.8 mol of NH_3/mol of NO_x, in particular 1.3-1.7 mol of NH_3/mol of NO_x. In this case, too, the addition of hydrocarbons and NH_3 can be applied directly to the process of the invention.

In a particular embodiment of the invention, the catalyst bed for NO_x reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas is divided into a plurality of reaction zones or physically separate reaction stages. A gradated introduction of NH_3 into the individual reaction zones or into the physically separate reaction zones of the catalyst bed arranged in the tailgas is preferably carried out.

The way in which the reducing agents are introduced into the gas stream to be treated can be chosen freely for the purposes of the invention, as long as the reducing agent is fed in upstream of the catalyst bed. The

introduction can, for example, be effected into the entry line upstream of the vessel or directly upstream of the catalyst bed. The reducing agent can be introduced in the form of a gas or else a liquid or aqueous solution which vaporizes in the gas stream to be treated. The introduction is carried out by means of a suitable device such as an appropriate pressure valve or appropriately configured nozzles which open into a mixer for the gas stream to be purified and the reducing agent introduced. When different reducing agents are used for NO_x and N_2O, they can be fed and introduced into the gas to be purified either separately or together.

As catalysts, use is made of iron-loaded zeolite catalysts which, based on the mass of zeolite, contain up to 25% of iron, but preferably from 0.1 to 10%.

Iron-loaded zeolite catalysts which are particularly preferably used according to the invention essentially contain > 50% by weight, in particular > 70% by weight, of one or more iron-loaded zeolites. Thus, for example, it is possible for not only an Fe-ZSM-5 zeolite but also a further iron-containing zeolite, e.g. an iron-containing zeolite of the FER type, to be present in the catalyst used according to the invention.

In addition, the catalyst used according to the invention can contain further additives known to those skilled in the art, e.g. binders.

Catalysts used according to the invention are very particularly preferably based on zeolites into which iron has been introduced by means of solid-state iron exchange. For this purpose, the commercially available ammonium zeolites (e.g. NH_4-ZSM-5) and the appropriate iron salts (e.g. $FeSO_4 \times 7\ H_2O$) are usually used as starting materials and these are mixed intensively with one another by mechanical means in a bore mill at room temperature (Turek et al.; Appl. Catal. 184, (1999) 249-256; EP-A-0 955 080). The catalyst powders obtained in this way are subsequently calcined in air at temperatures in the range from 400 to 600°C in a box furnace. After calcination, the iron-containing zeolites are intensively washed in distilled water and, after filtering off the zeolite, dried. The iron-containing zeolites obtained in this way are subsequently admixed with suitable binders and mixed and, for example, extruded to form cylindrical

catalyst bodies. Suitable binders are all binders customarily used; the most widely used binders here are aluminum silicates such as kaolin. It is naturally also possible to use iron-loaded zeolites produced by ion exchange in the liquid phase, for example those produced from the H form and/or the NH_4 form of the zeolites by exchange with an aqueous solution of iron salts.

Preference is given to using iron-loaded zeolite catalysts in which the zeolite is selected from the group consisting of the types MFI, BEA, FER, MOR, FAU and/or MEL and very particularly preferably from the group consisting of the types MFI and BEA and FER.

If the further N_2O removal in the catalyst bed in the tailgas is effected by decomposition into N_2 and O_2, very particular preference is given to using iron-loaded zeolites of the MFI and/or BEA and/or FER type, in particular an iron-loaded ZSM-5 zeolite.

If the further N_2O removal in the catalyst bed in the tailgas is effected by reduction of the N_2O by means of hydrocarbons, very particular preference is given to using iron-loaded zeolites of the MFI, BEA, FER, MOR, FAU and/or MEL type, in particular iron-loaded zeolites of the MFI and/or BEA type.

In the process of the invention or in the apparatus of the invention, the use of zeolites in which the lattice aluminum has been partly isomorphously replaced by one or more elements, for example replaced by one or more elements selected from among B, Be, Ga, Fe, Cr, V, As, Sb and Bi, is also included in the catalyst bed in the tailgas. The use of zeolites in which the lattice silicon has been isomorphously replaced by one or more elements, for example replaced by one or more elements selected from among Ge, Ti, Zr and Hf, is likewise included.

Precise information on the make-up or structure of the zeolites which are preferably used according to the invention is given in the Atlas of Zeolite Structure Types, Elsevier, 4th revised Edition 1996.

Very particular preference is given to using the above-defined zeolite catalysts which have been treated with steam ("steamed" catalysts) in the process of the invention or in the apparatus of the invention. The lattice of the zeolites is dealuminated by such a treatment; this treatment is known per se to those skilled in the art. These hydrothermally treated zeolite catalysts surprisingly display a particularly high activity in the process of the invention.

Preference is given to hydrothermally treated zeolite catalysts which have been loaded with iron and in which the ratio of extra-lattice aluminum to lattice aluminum is at least 1:2, preferably from 1:2 to 20:1.

The catalyst bed in the tailgas can be configured freely for the purposes of the invention. Thus, for example, the catalyst or catalysts can be arranged in a catalyst bed through which the gas flows axially or laterally, preferably radially, and which is arranged in one or more vessels.

In a further embodiment of the invention, one or more further stages for removal of N_2O and/or NO_x is/are arranged between the catalyst bed for N_2O decomposition arranged in the process gas and the catalyst bed for NO_x reduction and effecting a further decrease in N_2O arranged in the tailgas. In these stages, processes known per se for decreasing the amount of N_2O and NO_x are used. This can preferably be effected catalytically.

The invention also provides a nitric acid plant in which a catalytic removal of the N_2O formed in the catalytic NH_3 oxidation is carried out in the process gas and a further reduction of the N_2O content and a reduction of the NO_x content is carried out in the tailgas downstream of the absorption tower.

The plant comprises at least the following elements:

A) reactor for the catalytic oxidation of NH_3 by means of oxygen to produce an NO_x-containing process gas,
B) absorption tower for reacting the NO_x formed from the process gas with an absorption medium, preferably water, leaving a tail gas

containing NO_x and N_2O,

C) at least one first catalyst bed for N_2O decomposition through which the process gas flows and which is arranged downstream of the catalytic NH_3 oxidation and upstream of the absorption tower,

D) at least one second catalyst bed for NO_x reduction and effecting a further decrease in the N_2O content, through which the tailgas flows and which is arranged downstream of the absorption tower, and

E) at least one device for feeding gaseous reducing agent into the tailgas, which is arranged downstream of the absorption tower and upstream of the second catalyst bed, where

F) the first catalyst bed contains a catalyst suitable for the decomposition of N_2O, preferably a catalyst which contains transition metal oxides and/or transition metal-containing mixed oxides, preferably mixed oxides having a perovskite structure, a perovskite-like structure or a spinel structure, and/or noble metals as active component and

G) the second catalyst bed contains a catalyst containing iron-loaded zeolites.

Further preferred embodiments of the apparatus of the invention are described in the dependent claims.

Examples 1, 3 and 5 and also comparative examples 2, 4 and 6 below illustrate the invention without restricting it.

Examples 1, 3 and 5 and also comparative examples 2, 4 and 6 demonstrate the effect of the N_2O entry concentration on the achievable NO_x removal for the example of a deNO_x stage which contains an iron-loaded zeolite catalyst. The N_2O entry concentration selected in examples 1, 3 and 5 corresponds to an N_2O content which results from operation according to the invention of the catalyst stage in the process gas stream. The N_2O inlet concentration in the comparative examples 2, 4 and 6 gives a comparison with operation which is not according to the invention of the catalyst stage in the process gas stream.

The catalysts used in experiments 1 to 6 were iron-loaded zeolites of the ZSM-5 type (examples 1 to 4) or iron-loaded zeolites of the BEA type

(examples 5 and 6) which had been produced by solid-state ion exchange starting out from ZSM-5 or BEA zeolite powder in the ammonium form.

Detailed information on the preparation may be found in M. Rauscher, K. Kesore, R. Mönning, W. Schwieger, A. Tißler, T. Turek: "Preparation of highly active Fe-ZSM-5 catalyst through solid state ion exchange for the catalytic decomposition of N_2O" in Appl. Catal. 184 (1999) 249-256. The catalyst powder obtained was calcined in air at 823 k for six hours, washed and dried overnight at 383 K. After addition of an appropriate binder, the powder was extruded to give cylindrical catalyst bodies.

At a nominal degree of exchange of 100% and a modulus (SiO_2/Al_2O_3 ratio) of in each case about 25, the iron content of the catalyst samples before shaping was in each case about 5%.

To carry out the experiments for examples 1 to 6, the extrudates obtained were crushed and a particle size fraction of 0.5-1.25 mm was sieved out. Of this, 1.75 g (of the catalyst Fe-ZSM-5) or 1.50 g (of the catalyst Fe-BEA) were then in each case diluted to a bed volume of 12 ml with glass beads and introduced into a suitable flow tube reactor.

The operating temperature in the reactor tube was set by means of electric heating. The analysis of the gas streams entering and leaving the reactor was carried out by means of an FTIR spectrometer (model 5700, from Thermo) which was equipped with a heated 2 m long-path gas cell.

The precise experimental and operating conditions of the individual experiments are shown in Table 1 below.

Table 1: Operating conditions for experiments 1 to 6

Experiment			**1**	**2**	**3**	**4**	**5**	**6**
Process parameters	T	°C	440	440	420	420	400	400
	SV*)	h^{-1}	60 000	60 000	40 000	40 000	50 000	50 000
	P	bara	4.0	4.0	4.0	4.0	4.0	4.0
Gas composition on entry into the experiment reactor	NO_x	ppmv	1001	1012	1007	1000	510	502
	N_2O	ppmv	549	83	571	92	501	98
	H_2O	% by vol.	~0.31	~0.31	~0.31	~0.31	~0.32	~0.32
	O_2	% by vol.	1.0	1.0	1.0	1.0	1.0	1.0
	N_2O/NO_x	mol/mol	0.55	0.08	0.57	0.09	0.98	0.20
	NH_3	ppmv	1138	1088	1306	1175	664	548

*)SV = space velocity

The results of experiments 1 to 6 are shown in Table 2.

Table 2: Experimental results

Experiment			**1**	**2**	**3**	**4**	**5**	**6**
Gas composition at the outlet of the experimental reactor	NO_x	ppmv	32	81	8	27	1	16
	N_2O	ppmv	107	11	111	11	210	27
Degree of removal at the outlet of the experimental reactor	NO_x	%	96.8	80.0	99.2	97.3	99.8	96.8
	N_2O	%	80.5	86.7	80.6	88.0	58.1	72.4

*)SV = space velocity

NH_3 as reducing agent for NO_x is added in an amount corresponding to the maximum amount which can be added until analytically significant NH_3 breakthrough (about 1-2 ppmv) occurs; i.e. the indicated residual concentration of NO_x corresponds to the minimum possible residual concentration of NO_x, and the degree of removal of NO_x corresponds to the maximum degree of removal which can still just be achieved by addition of NH_3 without NH_3 breakthrough occurring.

As can be seen from Tables 1 and 2, a very much higher NO_x removal can in each case be achieved in the case of a high N_2O entry concentration (experiments 1, 3 and 5) than under otherwise identical conditions in the case of a reduced N_2O entry concentration, as in examples 2, 4 and 6. Thus, the residual NO_x concentration can be decreased from 81 ppmv to 32 ppmv in experiment 1 according to the invention compared to comparative experiment 2. In experiment 3 according to the invention, the residual NO_x concentration can be reduced from 27 to 8 compared to comparative experiment 4.

Finally, more or less complete NO_x reduction can be achieved in experiment 5 compared to experiment 6. This is of particular significance, since, as mentioned at the outset, in the case of complete NO_x reduction, a further N_2O reduction can be achieved by addition of hydrocarbons, preferably by means of methane, according to the process described in DE 102 15 605 A.

Overall, it can be seen from the above examples that in the case of an excessive reduction of the N_2O content in the upstream catalyst bed in the process gas, as is usually sought and realized, the deNO_x performance of the downstream catalyst bed through which the tailgas flows is significantly reduced and the desired degree of removal of NO_x can sometimes no longer be achieved.

Without knowledge of this wholly unexpected relationship, a person skilled in the art would have only the possibility of appropriately adapting the design of the downstream catalyst bed through which the tailgas flows, i.e. at given process parameters (pressure, temperature, volume flow),

appropriately increasing the size of the reactor or the catalyst volume of this catalyst bed. However, it is much more convenient to adapt the removal performance of the catalyst stage for N_2O decomposition in the process gas according to the invention so that a residual concentration of N_2O of > 100 ppmv, preferably > 200 ppmv and particularly preferably > 300 ppmv, and a molar N_2O/NO_x ratio of > 0.25, preferably > 0.5, result before entry into the catalyst bed.

The decreased degree of removal of N_2O is compensated for in the downstream catalyst bed through which the tailgas flows, i.e. through the catalyst bed charged with Fe-zeolite catalyst in the tailgas, which, according to the invention also brings about N_2O removal in parallel to the NO_x reduction. In experiments 1 to 4, in which the N_2O removal is effected by decomposition into N_2 and O_2, this is only slightly dependent on the N_2O entry concentration and under the process conditions (pressure, temperature, space velocity) selected is in the range from 80 to 90%.

Claims:

1. A process for preparing nitric acid by catalytic oxidation of NH_3 by means of oxygen and subsequent reaction of the NOx formed with an absorption medium in an absorption tower, which comprises a catalyst bed for N_2O decomposition arranged in the process gas downstream of the catalytic NH_3 oxidation and upstream of the absorption tower and a catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas downstream of the absorption tower,

- wherein the amount of N_2O removed in the catalyst bed for N_2O removal arranged in the process gas downstream of the catalytic NH_3 oxidation and upstream of the absorption tower is not more than that which results in an N_2O content of > 100 ppmv and a molar N_2O/NO_X ratio of > 0.25 before entry of the tailgas into the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas and
- the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas contains at least one iron-loaded zeolite catalyst and
- NH_3 is added to the tailgas before entry into the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas in such an amount that an NOx concentration of < 40 ppmv results at the outlet from the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas and
- the operating parameters of pressure, temperature and space velocity are selected in/over the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas in such a way that an N_2O concentration of < 200 ppmv results, with the proviso that the N_2O concentration at the outlet from the catalyst bed for NOx reduction is less than the N_2O content entering the catalyst bed for NOx reduction.

2. The process as claimed in claim 1, wherein the targeted setting of the N_2O removal in the catalyst bed arranged in the process gas is achieved by variation of the layer thickness or the layer height of the catalyst bed and/or selection of the catalyst material and/or selection of the geometry of the catalyst material.

3. The process as claimed in claim 1 or 2, wherein the catalyst bed arranged in the process gas contains a catalyst suitable for the decomposition of N_2O.

4. The process as claimed in claim 3, wherein the catalyst suitable for the decomposition of N_2O is a catalyst containing, as active component, transition metal oxides and/or transition metal-containing mixed oxides and/or noble metals.

5. The process as claimed in claim 4, wherein the mixed oxides have a perovskite structure, a perovskite-like structure or a spinel structure.

6. The process as claimed in any one of claims 1 to 5, wherein the catalyst material of the catalyst bed arranged in the process gas contains cobalt-containing oxides or mixed oxides as active components.

7. The process as claimed in claim 6, wherein the catalyst material is a catalyst material which comprises a porous support composed of polycrystalline or vitreous inorganic material, a cerium oxide functional layer applied thereto and a layer of oxidic cobalt-containing material applied thereto.

8. The process as claimed in any one of claims 1 to 6, wherein the catalyst material of the catalyst bed arranged in the process gas is configured as a shaped body which has the geometry of a cylinder, hollow cylinder, multi-hole cylinder, perforated and unperforated trilobes or polylobes or honeycomb structures.

9. The process as claimed in claim 8, wherein the catalyst material of the catalyst bed arranged in the process gas is configured as a shaped body whose equivalent diameter has a lower limit of > 1.5 mm and has an upper limit of < 20 mm.

10. The process as claimed in claim 9, wherein the catalyst material of the catalyst bed arranged in the process gas is configured as a shaped body whose equivalent diameter has a lower limit of > 3 mm.

11. The process as claimed in claim 10, wherein the catalyst material of the catalyst bed arranged in the process gas is configured as a shaped body whose equivalent diameter has a lower limit of > 5 mm.

12. The process as claimed in any one of claims 9 to 11, wherein the catalyst material of the catalyst bed arranged in the process gas is configured as a shaped body whose equivalent diameter has an upper limit of < 15 mm.

13. The process as claimed in claim 12, wherein the catalyst material of the catalyst bed arranged in the process gas is configured as a shaped body whose equivalent diameter has an upper limit of < 10 mm.

14. The process as claimed in any one of claims 1 to 13, wherein the pressure drop over the catalyst material of the catalyst bed arranged in the process gas is < 30 mbar and, where, the bed height of the catalyst material of the catalyst bed arranged in the process gas is 3-30 cm.

15. The process as claimed in claim 14, wherein the pressure drop over the catalyst material of the catalyst bed arranged in the process gas is < 25 mbar.

16. The process as claimed in claim 15, wherein the pressure drop over the catalyst material of the catalyst bed arranged in the process gas is < 20 mbar.

17. The process as claimed in claim 16, wherein the pressure drop over the catalyst material of the catalyst bed arranged in the process gas is < 15 mbar.

18. The process as claimed in claim 17, wherein the pressure drop over the catalyst material of the catalyst bed arranged in the process gas is < 10 mbar.

19. The process as claimed in any one of claims 14 to 18, wherein the bed height of the catalyst material of the catalyst bed arranged in the process gas is 5-20 cm.

20. The process as claimed in claim 19, wherein the bed height of the catalyst material of the catalyst bed arranged in the process gas is 10-20 cm.

21. The process as claimed in any one of claims 1 to 20, wherein the N_2O removal in the catalyst bed in the process gas is 40-90%, based on the amount of N_2O initially present.

22. The process as claimed in claim 21, wherein the N_2O removal in the catalyst bed in the process gas is 45-80%, based on the amount of N_2O initially present.

23. The process as claimed in claim 22, wherein the N_2O removal in the catalyst bed in the process gas is 50-70%, based on the amount of N_2O initially present.

24. The process as claimed in any one of claims 1 to 23, wherein the space velocity at which the tailgas is passed over the catalyst material of the catalyst bed arranged in the tailgas is from 200 to 200 000 h^{-1}, the pressure in the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas is from 1 to 50 bar, and the temperature in the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas is from 300°C to 600°C.

25. The process as claimed in claim 24, wherein the space velocity at which the tailgas is passed over the catalyst material of the catalyst bed arranged in the tailgas is from 5 000 to 10 000 h^{-1}.

26. The process as claimed in claim 24, wherein the space velocity at which the tailgas is passed over the catalyst material of the catalyst bed arranged in the tailgas is from 5 000 to 50 000 h^{-1}.

27. The process as claimed in any one of claims 24 to 26, wherein the pressure in the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas is at least 2 bar.

28. The process as claimed in claim 27, wherein the pressure in the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas is at least 3 bar.

29. The process as claimed in claim 28, wherein the pressure in the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas is from 4 to 25 bar.

30. The process as claimed in any one of claims 24 to 29, wherein the temperature in the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas is from 330°C to 520°C.

31. The process as claimed in any one of claims 1 to 30, wherein the operating parameters pressure, temperature and space velocity in/over the catalyst bed arranged in the tailgas are set and/or hydrocarbons are added as reducing agent for N_2O in this catalyst bed in such a way that the decrease in the content of N_2O in this catalyst bed is at least 50%, based on the content of N_2O at the entry into this catalyst bed.

32. The process as claimed in claim 31, wherein the decrease in the content of N_2O in this catalyst bed is at least 70% based on the content of N_2O at the entry into this catalyst bed.

33. The process as claimed in claim 32, wherein the decrease in the content of N_2O in this catalyst bed is at least 80% based on the content of N_2O at the entry into this catalyst bed.

34. The process as claimed in claim 33, wherein the decrease in the content of N_2O in this catalyst bed is from 90 to 100% based on the content of N_2O at the entry into this catalyst bed.

35. The process as claimed in any one of claims 1 to 34, wherein an amount of from 0.9 to 1.3 mol of NH_3 per mol of NOx to be reduced is added to the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas.

36. The process as claimed in claim 35, wherein the amount of from 1.0 to 1.2 mol of NH_3 per mol of NOx to be reduced is added to the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas.

37. The process as claimed in any one of claims 1 to 34, wherein hydrocarbons are mixed into the tailgas before entry into the catalyst material of the catalyst bed arranged in the tailgas, where an amount of 0.2-1 mol of hydrocarbon/1 mol of N_2O to be reduced are added and, based on the NO_x entry concentration, 1-2 mol of NH_3/mol of NOx, are added.

38. The process as claimed in claim 37, wherein the amount of 0.2-0.7 mol of hydrocarbon/1 mol of N_2O to be reduced are added.

39. The process as claimed in claim 37 or 38, wherein based on the NOx entry concentration 1.2-1.8 mol of NH_3/mol of NOx are added.

40. The process as claimed in claim 39, wherein based on the NOx entry concentration 1.2-1.7 mol of NH_3/mol of NOx are added.

41. The process as claimed in any one of claims 1 to 40, wherein the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas is divided into a plurality of reaction zones or physically separate reaction stages and gradated introduction of NH_3 into the individual reaction zones or into the physically separate reaction stages of the catalyst bed arranged in the tailgas is carried out.

42. The process as claimed in any one of claims 1 to 41, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas contains, based on the mass of zeolite, up to 25% of iron.

43. The process as claimed in claim 42, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas contains, based on the mass of zeolite from 0.1 to 10% of iron.

44. The process as claimed in any one of claims 1 to 43, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas contains > 50% by weight of an iron-loaded zeolite or a plurality of iron-loaded zeolites.

45. The process as claimed in claim 44, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas contains > 70% by weight of an iron-loaded zeolite or a plurality of iron-loaded zeolites.

46. The process as claimed in any one of claims 1 to 45, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas is a zeolite of the MFI, BEA, FER, MOR, FAU and/or MEL type.

47. The process as claimed in claim 46, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas is an iron-loaded zeolite of the MFI and/or BEA and/or FER type.

48. The process as claimed in any one of claims 1 to 47, wherein the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas comprises a zeolite whose lattice aluminum has been completely or partly isomorphously replaced by one or more elements, where the elements are selected from the group consisting of B, Be, Ga, Fe, Cr, V, As, Sb and Bi, or comprises a zeolite whose lattice silicon has been completely or partly replaced by one or more elements selected from the group consisting of Ge, Ti, Zr and Hf and/or the iron-loaded zeolite catalyst of the catalyst bed arranged in the tailgas comprises a zeolite which has been hydrothermally pretreated with steam.

49. The process as claimed in claim 48, wherein the zeolite which has been thermally pretreated with steam has a ratio of extra-lattice aluminum to lattice aluminum of at least 1:2.

50. The process as claimed in claim 49, wherein the zeolite which has been thermally pretreated with steam has a ratio of extra-lattice aluminum to lattice aluminum of from 1:2 to 20:1.

51. The process as claimed in any one of claims 1 to 50, wherein the gas flows axially, laterally or radially through the catalyst bed arranged in the tailgas.

52. The process as claimed in any one of claims 1 to 51, wherein one or more further stages for N_2O and/or NOx removal are arranged between the catalyst bed for N_2O decomposition arranged in the process gas and the catalyst bed for NOx reduction and effecting a further decrease in the amount of N_2O arranged in the tailgas.

53. A nitric acid plant being configured for performing the process according to any one of claims 1 to 52 in which a catalytic removal of the N_2O formed in the catalytic NH_3 oxidation is carried out in the process gas and a further reduction of the N_2O content and a reduction of the NOx content is carried out in the tailgas downstream of the absorption tower, which comprises at least the following elements:

A) reactor for the catalytic oxidation of NH_3 by means of oxygen to produce an NOx-containing process gas,

B) absorption tower for reacting the NOx formed from the process gas with an absorption medium, leaving a tailgas containing NOx and N_2O,

C) at least one first catalyst bed for N_2O decomposition through which the process gas flows and which is arranged downstream of the catalytic NH_3 oxidation and upstream of the absorption tower in the flow direction,

D) at least one second catalyst bed for NOx reduction and effecting a further decrease in the N_2O content, through which the tailgas flows and which is arranged downstream of the absorption tower in the flow direction, and

E) at least one device for feeding gaseous reducing agent into the tailgas, which is arranged downstream of the absorption tower and upstream of the second catalyst bed in the flow direction, where

F) the first catalyst bed contains a catalyst suitable for the decomposition of N_2O and

G) the second catalyst bed contains a catalyst containing at least one iron-loaded zeolite.

54. The nitric acid plant as claimed in claim 53, wherein the absorption tower of (B) is configured for reacting the NOx formed from the process gas with an absorption medium of (B) that is water.

55. The nitric acid plant as claimed in claim 53 or 54, wherein the catalyst suitable for the decomposition of N_2O of (F) is a catalyst which contains, as active component, transition metal oxides and/or transition metal-containing mixed oxides and/or noble metals.

56. The nitric acid plant as claimed in claim 55, wherein the mixed oxides have a perovskite structure, a perovskite-like structure or a spinel structure.

57. The nitric acid plant as claimed in any one of claims 53 to 56, wherein the catalyst material of the first catalyst bed contains cobalt-containing oxides or mixed oxides as active components.

58. The nitric acid plant as claimed in claim 57, wherein the catalyst material comprises a porous support composed of polycrystalline or vitreous inorganic material, a cerium oxide functional layer applied thereto and a layer of oxidic cobalt-containing material applied thereto.

59. The nitric acid plant as claimed in any one of claims 53 to 58, wherein the bed height of the catalyst material of the first catalyst bed is 3-30 cm.

60. The nitric acid plant as claimed in claim 59, wherein the bed height of the catalyst material of the first catalyst bed is 5-20 cm.

61. The nitric acid plant as claimed in claim 60, wherein the bed height of the catalyst material of the first catalyst bed is 10-20 cm.

62. The nitric acid plant as claimed in any one of claims 53 to 61, wherein the iron-loaded zeolite catalyst in the second catalyst bed is a zeolite of the MFI, BEA, FER, MOR, FAU and/or MEL type.

63. The nitric acid plant as claimed in claim 62, wherein the iron-loaded zeolite catalyst in the second catalyst bed is an iron-loaded zeolite of the MFI and/or BEA and/or FER type.