



# The influence of $\text{NO}_x$ presence on the catalytic $\text{N}_2\text{O}$ decomposition over the supported double-promoted cobalt spinel catalyst

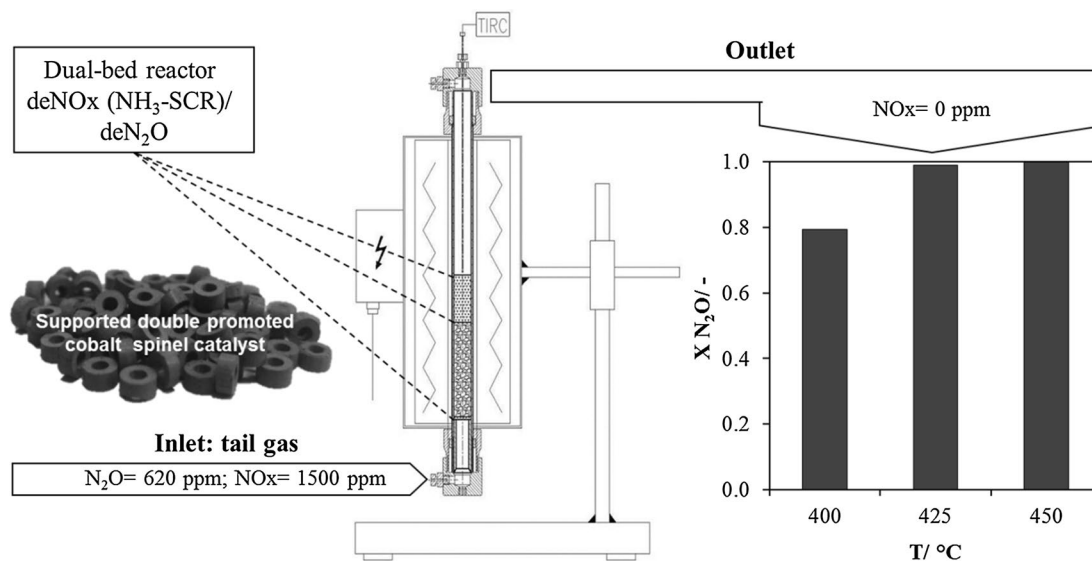
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## Abstract

The supported double-promoted cobalt spinel catalyst has been tested for  $\text{N}_2\text{O}$  decomposition in the model and real gas streams (tail gas) from nitric acid plant. Activity tests carried out in the model gas stream showed a high activity of the developed supported catalyst. However, during catalytic investigations performed under real industrial conditions it was found that this catalyst is very sensitive to  $\text{NO}_x$  poisoning. The negative influence of these oxides is visible even at their low concentration. However, the stability tests performed in real nitric acid plant gas streams showed that the developed catalyst preserves its initial activity for at least 40 days. Therefore, the authors propose to apply it in a dual-bed catalytic system consisting of  $\text{deNO}_x$  and  $\text{deN}_2\text{O}$  catalyst beds. At the first stage, tail gas is purified from  $\text{NO}_x$  in the  $\text{NH}_3$ -SCR process in the presence of commercial catalyst and at the second stage, catalytic decomposition of  $\text{N}_2\text{O}$  into  $\text{N}_2$  and  $\text{O}_2$  takes place over the developed supported cobalt spinel catalyst. The  $\text{deNO}_x$  catalyst bed applied in the proposed dual-bed catalytic system allowed for the complete conversion of  $\text{NO}_x$  from the tail gases. After installation of the supported cobalt spinel catalyst downstream of the  $\text{deNO}_x$  catalyst bed, a significant increase in  $\text{deN}_2\text{O}$  catalyst activity in the real tail gas stream was observed. Under optimal conditions, this system ( $F = 150 \text{ L/h}$ ;  $T = 425\text{--}450 \text{ }^\circ\text{C}$ ) allowed for the complete reduction of  $\text{N}_2\text{O}$  emission. It is shown that the proposed catalytic system is robust and allows to obtain high conversions of both  $\text{NO}_x$  and  $\text{N}_2\text{O}$ .

## Graphical abstract



**Keywords** Low-temperature  $\text{N}_2\text{O}$  decomposition · Supported cobalt spinel catalyst · Dual-bed catalytic reactor

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## Introduction

Nitric acid plants constitute one of major industrial sources of nitrous oxide (I) where it is formed as a byproduct in the catalytic oxidation of ammonia on Pt–Rh gauzes. The formed nitrous oxide (I) does not take part in any technological reactions of nitric acid production process. Therefore, its total amount being formed in ammonia reactor is practically transferred to tail gases and is emitted to the atmosphere. Its final concentration depends on plant-operating parameters and it can reach the level of 2000 ppm (Pérez-Ramirez et al. 2003). Directive (2010) imposes on EU members, the obligation to reduce the emission of this gas below 150 ppm for existing plants and below 100 ppm for new nitric acid plants. Meeting the requirements related to N<sub>2</sub>O emissions, especially in existing nitric acid plants, needs catalytic processes to be implemented in the production technology. The analysis of operating parameters of nitric acid plant indicates that the process can be carried out in nitrous gas stream ( $T=750\text{--}940\text{ }^{\circ}\text{C}$ ) and/or tail gas stream ( $T=200\text{--}450\text{ }^{\circ}\text{C}$ ) (Pérez-Ramirez et al. 2003).

In studies of removing nitrous oxide (I) from tail gases, numerous catalytic systems were investigated (Kapteijn et al. 1996). Generally, they can be categorized into three groups: modified zeolites (Cheng et al. 2012; Hussain et al. 2012; Zhang et al. 2012; Melian-Cabrera et al. 2018), supported catalyst (Shen et al. 2009; Karaskova et al. 2012; Rico-Pérez and Bueno-López 2014) and modified oxide catalysts (Pasha et al. 2009; Karásková et al. 2010; Zabilskiy et al. 2015).

The analysis of literature reports implies that cobalt oxide is one of the most active substances in catalysts for N<sub>2</sub>O decomposition. Studies of decomposition process of N<sub>2</sub>O from the stream of tail gases with the use of doped spinel catalyst obtained using precipitation method and formed into tablets (Wilk et al. 2009; Inger et al. 2011a, b, 2013) showed that the catalyst is active and stable under industrial conditions (Inger et al. 2014). At the same time, it was found that the reaction rate is limited by internal diffusion resistance. As a result of this fact, a significant internal part of the shape does not take part in the contact process. Bearing in mind a high price of the main raw material, cobalt oxide, the production cost of catalyst may increase radically, making the proposed solution economically ineffective. Therefore, further research was focused on finding such a catalyst formula where its active phase would be the proposed double-promoted cobalt spinel placed on the support. During the course of studies, most promising supports in terms of their applicability were specified: cordierite monoliths (Grzybek et al. 2015) and aluminum oxide (Grzybek et al. 2017), and the method of placing the active phase on the support and the production technology of the final catalyst formula in pilot scale was developed (Wilk et al. 2016). As a result

of studies, it was found that the active phase placed onto support in the amount of 20% wt. sufficient to obtain high-activity catalyst. Therefore, the mass of costly component of the catalyst was significantly reduced.

A significant limitation for the application of the developed catalyst in the removal of N<sub>2</sub>O from tail gases is its resistance to inhibiting effect of other components, i.e. H<sub>2</sub>O, O<sub>2</sub>, and NO<sub>x</sub> in particular. Catalysts containing transition metals (e.g. cobalt, iron) are generally more sensitive to the presence of NO<sub>x</sub> in gas stream which can be easily absorbed on their active centers, especially at temperatures < 400 °C and NO<sub>x</sub> can block these active centers for N<sub>2</sub>O decomposition. Therefore, the use of developed catalyst in the industrial nitric acid plants requires a prior reduction of NO<sub>x</sub> to the lowest possible level.

The NO<sub>x</sub> removal process can be realized by a selective catalytic reduction of NO<sub>x</sub> with using NH<sub>3</sub> as a reducing agent. The low-temperature NH<sub>3</sub>-SCR method is successfully and widely used in existing nitric acid plants (Luck and Roiron 1989; Avila et al. 1993). As reported by (Pérez-Ramirez et al. 2000; Kim et al. 2011, 2012), a practical and simple implementation method may be the application of dual-bed catalyst systems to NO<sub>x</sub> reduction as well as the removal of harmful byproducts as N<sub>2</sub>O, especially in the low-temperature regime.

The aim of the study was to investigate the effect of NO<sub>x</sub> presence on the activity of the supported double-promoted cobalt spinel catalyst. Moreover, operating conditions for deN<sub>2</sub>O catalyst in a dual-bed catalytic system were determined, consisting of a deNO<sub>x</sub> and deN<sub>2</sub>O catalyst bed. In this case, in the first bed, the NO<sub>x</sub> was removed by NH<sub>3</sub>-SCR process over the commercial catalyst and in the second bed, N<sub>2</sub>O was catalytically decomposed into N<sub>2</sub> and O<sub>2</sub> over supported cobalt spinel catalyst (Wilk et al. 2016). It is shown that this dual-bed catalytic system is robust and allows to achieve high conversions of both NO<sub>x</sub> and N<sub>2</sub>O.

## Experimental

### Catalyst preparation

The raw material used to prepare support was the aluminum hydroxide oxide. It was mixed with graphite and plasticizer with the addition of water and then shaped into Raschig rings with the following dimensions: diameter × length × hole diameter = 4 × 2 × 1 mm. The shaped sample was then dried in the air and calcined at 1400 °C for 4 h.

Supported cobalt spinel catalyst was prepared by impregnation of α-Al<sub>2</sub>O<sub>3</sub> support with an aqueous solution of cobalt, zinc and potassium salts. The concentration of salts in the impregnating solution was adjusted to obtain 20 wt%

loading of the active phase in the whole mass of supported catalyst. After impregnation of the support, the samples were dried at 120 °C and then calcined in the air at 450 °C for 4 h (Wilk et al. 2016).

### The activity tests of the catalyst

All catalytic tests were performed in a tubular reactor with a diameter  $d=25$  mm, equipped with a heating mantle and a temperature control unit (Fig. 1). For the activity tests, a model gas mixture ( $N_2O/NO/N_2$ ) and the real tail gas were used. The model gas contained 1500 or 2000 ppm  $N_2O$ , 0–1000 ppm NO and  $N_2$  as a balance. Tail gases were taken directly from the pilot nitric acid plant and their composition was as follows: 420–650 ppm  $N_2O$ , 150–2450 ppm  $NO_x$ , 0.3–0.4 vol.%  $H_2O$ , 1.3–2.2 vol.%  $O_2$  and  $N_2$ . The initial concentration of ammonia after mixing with tail gases for SCR process was 0.1 or 0.2 vol.%. All the catalytic tests were performed at the temperature range of 300–450 °C, corresponding to the conditions of a tertiary  $N_2O$  abatement process in the nitric acid plants, at the gas flow within the range of 100–600  $Ndm^3/h$ .

In case of activity studies of  $deN_2O$  catalyst with no  $deNO_x$  catalyst applied, the catalytic bed layer with the volume of approx. 30  $cm^3$  was used, which was placed on the layer of inert bed in the form of aluminum oxide shapes with the volume of 50  $cm^3$ . In studies with the application of dual-bed catalytic system, the layer of  $deN_2O$  with the volume of 30  $cm^3$  and the layer of  $deNO_x$  with the volume of 50  $cm^3$  were applied.

Measurements of  $N_2O$  conversion at the assumed temperature of ( $T=const$ ) were carried out after 10 min of the moment it was stabilized. The amount of ammonia fed to the reactor during studies of  $deN_2O$  catalyst in dual-bed catalytic system was determined based on the intensity of tail gas flow and the initial concentration of  $NO_x$  in these gases. Tail

gases before mixing with gaseous  $NH_3$  were preheated to the temperature of 300 °C.

All studies were carried out with the use of formed catalysts in the shape that will be used in industrial practice. The composition of the gases (at the inlet and outlet of the reactor) was measured by a Gasetm™ CX 4000 FTIR Gas Analyzer (Gasetm Technologies).

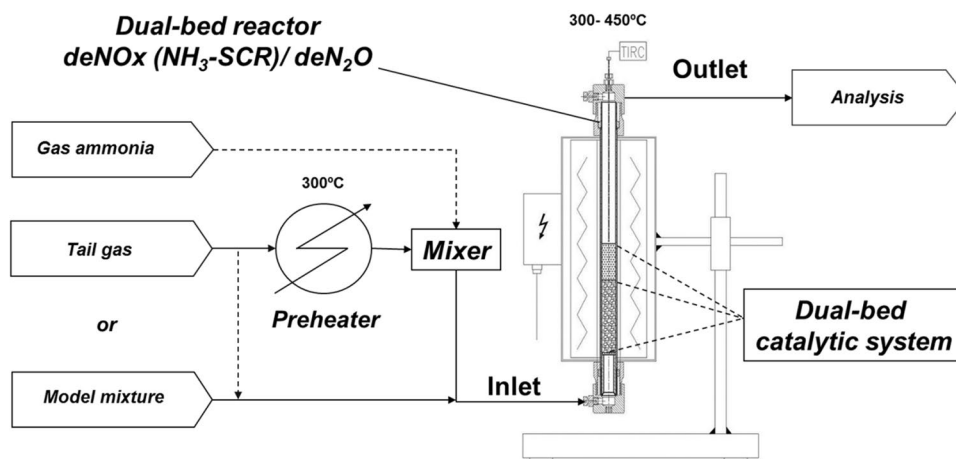
## Results and discussion

### The activity of $deN_2O$ catalyst

The results of the activity tests performed in the model gas mixture showed that the developed  $deN_2O$  catalyst has a very high activity in the absence of  $NO_x$ . At the temperature of 380 °C and with gas flow amounting to 300  $Ndm^3/h$ , the  $N_2O$  conversion above 85% was achieved. A double decrease of gas flow intensity allowed to achieve almost a 100%  $N_2O$  conversion at this temperature (Table 1).

The presence of NO in gas stream causes the shift of temperature window of  $deN_2O$  process towards higher temperatures (Fig. 2) and the temperature increase of decomposition process is directly proportional to NO concentration at the inlet to the reactor. It is particularly visible at temperatures of 300–370 °C. This interdependence is reflected in the temperature difference of a 50%  $N_2O$  conversion (Table 2). The increase of NO concentration in gas stream 100 ppm causes temperature increase  $T_{50\%}$  by 30 °C. However, at the temperature above 400 °C, the complete conversion of  $N_2O$  over supported cobalt spinel catalyst was achieved in all measurements, regardless of the initial content of NO in the inlet gas. The inhibiting effect of NO in gas stream may be linked with adsorption of NO particles on active centers of the catalyst, which contributes to blocking of these centers for  $N_2O$  decomposition. This can be prevented with the increase of  $deN_2O$  process temperature to the level above 400 °C. Such a

**Fig. 1** Scheme of the reactor used for the activity tests of supported cobalt spinel catalyst in a single-bed ( $deN_2O$ ) and dual-bed [ $deN_2O/deNO_x$  ( $NH_3$ -SCR)] catalytic system

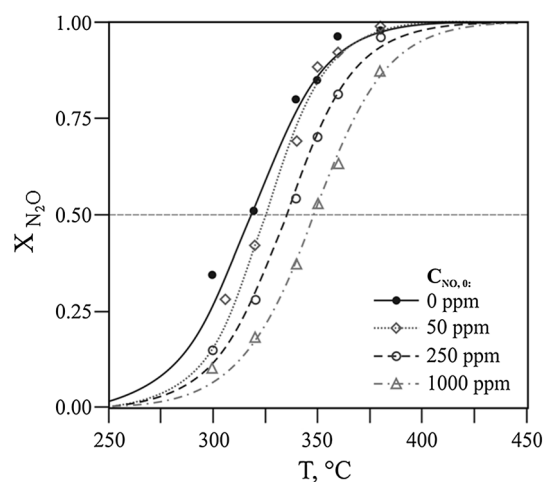


temperature level has a favorable effect on NO desorption from the catalyst surface, making the active surface available for N<sub>2</sub>O decomposition reaction. The reversible nature of poisoning the cobalt catalyst was described in paper Inger et al. (2011a, b).

During the experiments carried out under real industrial conditions (Table 3) despite the lower NO<sub>x</sub> content than in the model gas stream (Table 2), an additional temperature increase  $T_{50\%}$  was observed. For example, for model gas with NO<sub>x</sub> concentration = 250 ppm,  $T_{50\%}$  was 335 °C, while in the real tail gases with NO<sub>x</sub> concentration = 150 ppm,  $T_{50\%}$  was 380 °C. It can be concluded that other tail gas components (O<sub>2</sub> and H<sub>2</sub>O) can also have a negative effect on activity of the developed catalyst. Therefore, high final conversion of N<sub>2</sub>O can be attained by increasing the temperature reaction and/or extending the residence time of reaction gas in the catalyst bed (Table 3).

The increase of NO<sub>x</sub> concentration at the inlet of the reactor resulted in a decrease of N<sub>2</sub>O conversion, regardless of the gas flow rate. The exception was the results obtained at the lowest measured gas flow rate for which N<sub>2</sub>O conversions observed with the NO<sub>x</sub> content of 150 and 650 ppm were at the similar level.

The stability of developed catalyst was also investigated. In Fig. 3, the results of N<sub>2</sub>O decomposition for the fresh and used catalysts (after 500 h and 1000 h of operation) are shown. The experimental data are expressed as a function of N<sub>2</sub>O conversion ( $X_{N_2O}$ ) versus temperature. The stability tests ( $T=350\text{--}450$  °C,  $F=150$  Ndm<sup>3</sup>/h, 5000 GHSV, tail gas composition: 600–650 ppm N<sub>2</sub>O, 44–520 ppm NO<sub>x</sub>, 2.2% H<sub>2</sub>O and 2.2% O<sub>2</sub>) showed that the developed catalyst preserves its initial activity for at least 40 days, wherein the N<sub>2</sub>O conversion during this period remains at the level of > 0.95 for 420 °C, 0.62–0.80 for 400 °C and 0.34–0.55 for 380 °C. Differences in conversion curves as presented in Fig. 3 are the result of a large slope of the curves, which is caused by a large increase in the catalyst activity in the temperature range 350–420 °C. Moreover, differences in conversion curves may also result from a small difference in the



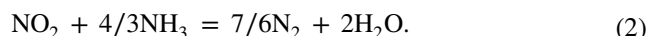
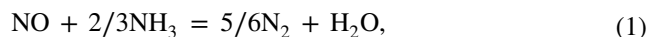
**Fig. 2** The effect of NO<sub>x</sub> content in a model gas mixture on the N<sub>2</sub>O conversion over supported cobalt spinel catalyst (NO=0–1000 ppm, N<sub>2</sub>O=2000 ppm,  $F=150$  Ndm<sup>3</sup>/h; 5000 GHSV)

composition of tail gases or from accuracy of the measurement method ( $\Delta T_{50\%} \leq 10$  °C).

The results presented above show that the supported double-promoted cobalt spinel catalyst can be applied in the industrial nitric acid installations if prior reduction of NO<sub>x</sub> in the tail gas stream to the lowest possible level was carried out. Therefore, further investigations were focused on determining the activity of the developed deN<sub>2</sub>O catalyst in the presence of the second catalyst bed for NO<sub>x</sub> reduction using NH<sub>3</sub>-SCR process.

### Dual-bed catalytic system

The temperature window for selective catalytic reduction falls in the range of 200–450 °C depending on the type of the catalyst applied. The process efficiency is determined with the amount of NH<sub>3</sub> fed into the inlet mixture according to the following reactions:



In the case of initial concentration of the reducing agent about 0.1 vol.%, low NO<sub>x</sub> conversion was observed (Fig. 4a). No ammonia was found at the outlet of the reactor, that the ammonia has completely reacted. It was

**Table 1** Comparison of the activity of developed deN<sub>2</sub>O catalyst in a model gas mixture containing 1500 and 2000 ppm N<sub>2</sub>O/N<sub>2</sub>

Model gas mixture			
$F$ , Ndm <sup>3</sup> /h; GHSV	$C_{\text{NO}_x}$ ppm	$C_{\text{N}_2\text{O}}$ ppm	N <sub>2</sub> O conversion, %
			$T$ , °C
			350    380
300/10,000	0	1500	87.3
150/5000	0	1500	91.5    100.0
150/5000	0	2000	87.8    99.2

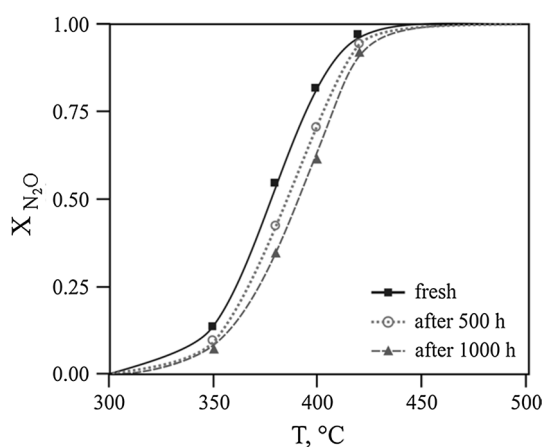
**Table 2** The temperature of 50% N<sub>2</sub>O conversion for different NO<sub>x</sub> concentration in the model gas mixture at the inlet of the reactor

$C_{\text{NO}_x}$ , ppm	0	50	250	1000
$T_{50\%}$ , °C	318	325	335	348

**Table 3** The activity of supported catalyst in the form of Raschig rings

$F$ , Ndm <sup>3</sup> /h; GHSV	$C_{\text{NO}_x}$ , ppm	$C_{\text{N}_2\text{O}}$ , ppm	$\text{N}_2\text{O}$ conversion, %				
			$T$ , °C				
			350	380	400	420	450
150/5000	150	550	23.3	49.1	80.0	n.m	n.m.
150/5000	650	440	11.4	56.4	81.8	93.1	n.m.
150/5000	1300	510	10.8	35.3	65.7	84.8	n.m.
300/10,000	150	550	n.m.	16.4	38.2	67.3	n.m.
300/10,000	650	440	n.m.	9.1	27.3	52.3	n.m.
300/10,000	1300	518	n.m	5.4	17.0	40.2	n.m.
600/20,000	150	550	0.0	0.0	3.6	11.8	46.4
600/20,000	1300	550	0.0	0.0	2.7	3.6	20.0

Measurements of real tail gases. Gase composition:  $\text{NO}_x$  and  $\text{N}_2\text{O}$  concentration in table,  $\text{H}_2\text{O}$ : 0.3–0.4 vol.%,  $\text{O}_2$ : 1.3–2.2 vol.% and  $\text{N}_2$ : rest



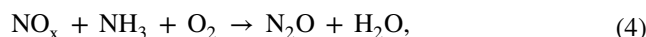
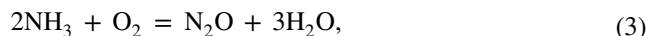
**Fig. 3** Conversion curves for the fresh and used supported cobalt spinel catalyst ( $F=150$  Ndm<sup>3</sup>/h, 5000 GHSV), tail gas composition: 600–650 ppm  $\text{NO}_x$ , 0.3–0.4 vol.%  $\text{H}_2\text{O}$ , 2.2 vol.%  $\text{O}_2$ , 440–520 ppm  $\text{N}_2\text{O}$ )

concluded that ammonia was deficient. The use of 0.2 vol.%  $\text{NH}_3$  excess allowed for the complete  $\text{NO}_x$  reduction in the temperature range from 250 to 400 °C. However, the presence of ammonia at the outlet of the reactor was noted (Fig. 4b), which means that an excess of ammonia was used in relation to the initial concentration of  $\text{NO}_x$  (1480 ppm) in tail gases. Limitations of the laboratory system (control of the tail gas and ammonia flow rate) made it impossible to determine the optimal amount of ammonia, which on the one hand would allow to completely reduce  $\text{NO}_x$ , on the other hand, it would not cause ammonia enter into the outlet of the reactor.

Increasing the temperature above 400 °C caused a rapid drop in the catalyst activity. Furthermore, increasing the temperature above 300 °C caused the formation of additional  $\text{N}_2\text{O}$ .

As reported by Wang et al. (2015),  $\text{N}_2\text{O}$  is the major byproduct of the de $\text{NO}_x$   $\text{NH}_3$ -SCR process, regardless of

the type of the catalyst used. In general, there are three main pathways for  $\text{N}_2\text{O}$  formation during  $\text{NH}_3$ -SCR process (Liatti et al. 1999; Yates et al. 2005; Martín et al. 2007):

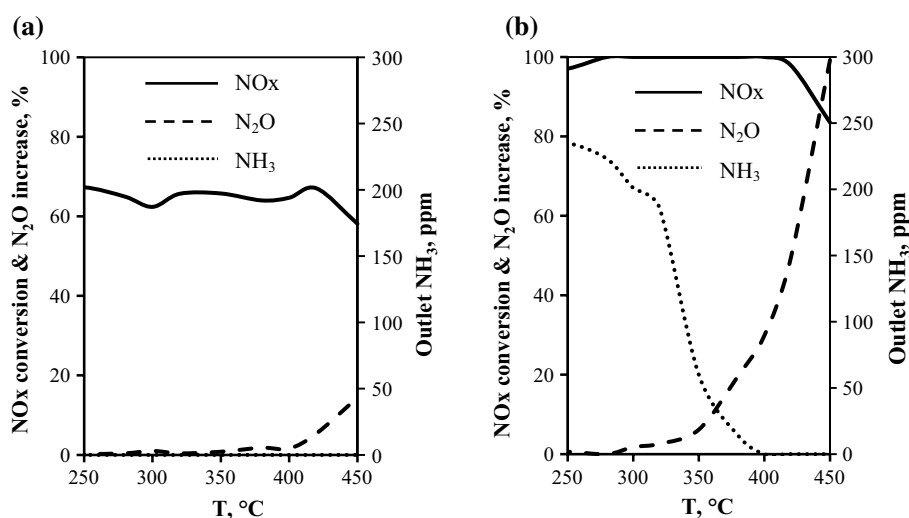


The results shown in the Fig. 4b show that the excess of  $\text{NH}_3$  used in the de $\text{NO}_x$   $\text{NH}_3$ -SCR process reacts to  $\text{N}_2\text{O}$  (according to the reactions 3 and 4). The formation of additional amount of  $\text{N}_2\text{O}$  in the stream of gas leaving the de $\text{NO}_x$  catalyst bed is not a problem as nitrous oxide will be decomposed in the next catalytic bed over the developed de $\text{N}_2\text{O}$  catalyst. However, in case of the application of excessive amount of ammonia the possibility of ammonia slip at the outlet of reactor is a problem. In the current case, ammonia slip occurred at temperatures lower than 400 °C whilst with the temperature exceeding this level, ammonia slip was converted into  $\text{N}_2\text{O}$ . Thus, in the proposed dual-bed catalytic system consisting of de $\text{NO}_x$  commercial catalyst and the developed supported de $\text{N}_2\text{O}$  catalyst, it is necessary to use the stoichiometric or slightly higher amount of  $\text{NH}_3$  reducing agent (for a complete  $\text{NO}_x$  reduction). On the one hand, it will ensure complete  $\text{NO}_x$  conversion; on the other hand, it will not cause the formation of a large amount of additional  $\text{N}_2\text{O}$ .

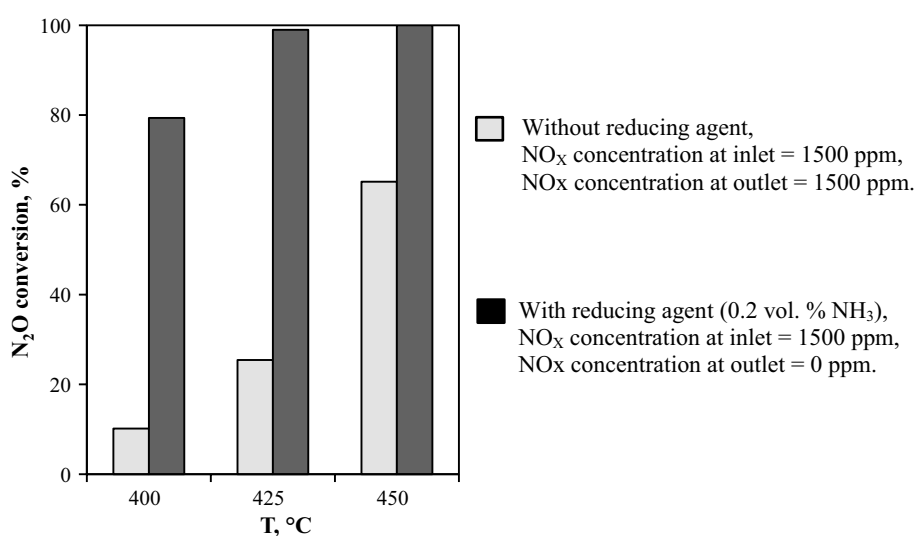
During the next stage of the study,  $\text{N}_2\text{O}$  conversion in the dual-bed catalytic system was investigated under real industrial conditions (tail gas composition:  $\text{N}_2\text{O}=620$  ppm,  $\text{NO}_x=1500$  ppm,  $\text{O}_2=1.35$  vol.%,  $\text{H}_2\text{O}=0.30$  vol.%). The results of the developed catalyst activity are shown in Fig. 5. The results show that even for the best process conditions ( $F_{\text{GR}}=300$  L/h,  $T=450$  °C) without reducing agent, tested catalytic system allows for a maximum of



**Fig. 4** The process of selective catalytic reduction of  $\text{NO}_x$  with using  $\text{NH}_3$  in the presence of the commercial catalyst samples in the case of **a**  $\text{NH}_3$  deficiency (0.1 vol.%) and **b**  $\text{NH}_3$  excess (0.2 vol.%). Tail gas composition:  $\text{N}_2\text{O} = 500$  ppm,  $\text{NO}_x = 1480$  ppm,  $\text{O}_2 = 1.13$  vol.%,  $\text{H}_2\text{O} = 0.30$  vol.%,  $F = 600 \text{ Ndm}^3/\text{h}$ , 20,000 GHSV



**Fig. 5** Comparison of supported CoZnK catalyst activity in a dual-bed catalytic system in the absence of the reducing agent ( $\text{NH}_3$ ) in the tail gas stream and after its addition ( $F_{\text{GR}} = 300 \text{ Ndm}^3/\text{h}$ , 10,000 GHSV, tail gas composition:  $\text{N}_2\text{O} = 620$  ppm,  $\text{NO}_x = 1500$  ppm,  $\text{O}_2 = 1.35$  vol.%,  $\text{H}_2\text{O} = 0.30$  vol.%; reducing agent 0.2 vol.%  $\text{NH}_3$ )



65%  $\text{N}_2\text{O}$  conversion. However, under optimal conditions ( $F_{\text{GR}} = 300 \text{ L/h}$ , 5000 GHSV,  $T = 425\text{--}450$  °C), the dual-bed catalytic system with reducing agent allows for the complete reduction of  $\text{NO}_x$  and complete conversion of  $\text{N}_2\text{O}$  from the tail gases. Despite the possibility of formation of the additional amount of nitrous oxide in de $\text{NO}_x$  catalytic bed, at the temperature over 400 °C, no presence of  $\text{NO}_x$ ,  $\text{NH}_3$  and  $\text{N}_2\text{O}$  was found in tail gases stream.

The proposed dual-bed catalytic system can be used in the process of  $\text{N}_2\text{O}$  decomposition from the tail gases, especially in the high- and dual-pressure nitric acid plants. In addition, the  $\text{N}_2\text{O}$  abatement in the temperature used is a major challenge for low-temperature de $\text{NO}_x$  systems (Yates et al. 2005; Martín et al. 2007). Usually, significantly higher operating parameters are required for an effective course of  $\text{N}_2\text{O}$  catalytic decomposition compared to the typical temperature of de $\text{NO}_x$  process. The presented method of the two-stage tail gas purification can be then an attractive solution to this problem. In this case, the system of two catalysts was

applied, the effectiveness of which (> 400 °C) is sufficient for achieving the residual emissions of  $\text{N}_2\text{O}$ ,  $\text{NO}_x$  and  $\text{NH}_3$  close to 0.

## Conclusions

Activity tests of the supported double-promoted cobalt spinel catalyst carried out in the model gas stream showed a high activity of the developed catalyst. However, during catalytic investigations performed under real industrial conditions it was found that this catalyst is very sensitive to  $\text{NO}_x$  poisoning.

As a solution, a dual-bed catalytic system was proposed, consisting of the two consecutive catalysts layers, de $\text{NO}_x$  commercial catalyst ( $\text{NH}_3\text{-SCR}$ ) and developed de $\text{N}_2\text{O}$  catalyst for a successive removal of these nitrogen oxides from the tail gas stream.

The use of stoichiometric amount or slightly higher amount of ammonia in deNO<sub>x</sub> SCR process with respect to the initial concentration of NO<sub>x</sub> in the tail gas ensures the complete NO<sub>x</sub> reduction and does not cause ammonia slip. After applying deNO<sub>x</sub> catalyst bed before the supported double-promoted cobalt spinel catalyst bed (dual-bed configuration), a high activity of the developed deN<sub>2</sub>O catalyst was found at temperature range of 425–450 °C.

The proposed dual-bed catalytic system can be used in the process of N<sub>2</sub>O decomposition from the tail gases, especially in the high- and dual-pressure nitric acid plants if sufficiently high temperature of tail gases is ensured.

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## Compliance with ethical standards

**Conflict of interest** On behalf of all authors, the corresponding author states that there is no conflict of interest.

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