FIELD STUDY, DESIGN, AND CATALYST COST OF SELECTIVE CATALYTIC REDUCTION PROCESS

By Hsunling Bai,[1] Shang-Hsiu Lee,[2] Chia-Hsin Lin,[3] Jyh Feng Hwang,[4] and Chang-Liang Hsieh[5]

ABSTRACT: A pilot TiO_2/V_2O_5 based selective catalytic reduction (SCR) system was installed in a glass manufacturing factory. The effects of important operation parameters such as the reaction temperature, space velocity, and inlet NH_3/NO_x molar ratio on the NO removal were evaluated. It was observed that >1 mol of NO_x is reacted for every 1 mol of NH_3 injected. This observation is different from that reported in the literature for laboratory experiments. The key design parameter values of the reaction rate and the NH_3 adsorption rate were determined based on results of the pilot tests using a theoretical SCR model developed by the same research group. The SCR model is a simple analytical solution that accounts for the simultaneous mass transfer and chemical reaction of an SCR reactor. The model results were compared to field data of NO_x conversion efficiency and NH_3 slip, and they were in reasonable agreement. The required catalyst volume and the cost of catalysts were then determined with the assistance of the design model.

INTRODUCTION

The selective catalytic reduction (SCR) process is a commercialized technology for an efficient NO_x control. The NO conversion efficiency of an SCR device can be well above 85% with proper design and operation. The NO_x in flue gas is removed by the injection of NH_3. The NH_3 and NO_x gaseous molecules react in a heterogeneous catalytic reactor, and nontoxic N_2 and H_2O gases are formed during the catalytic reaction. For commercial applications, catalysts are used as monolithic materials with honeycomb structures or in parallel-plate geometry (Boer et al. 1990; Casagrande et al. 1999).

Cho (1994) discussed operating parameters that affect the performance of an SCR system. The conversion efficiency of NO_x depends on the amount of injected NH_3, type of catalysts, residence time of the gas in the reactor, and specific surface area of the catalysts, etc. Cho (1994) indicated that increasing the ammonia injection quantity leads to better SCR performance. This phenomenon has been seen in the literature for an NH_3/NO inlet molar ratio of <1. And experimental studies also revealed that, for $NH_3/NO > 1$, the NH_3 concentration has almost no effect on the NO removal (Beeckman and Hegedus 1991; Lefers et al. 1991; Svachula et al. 1993). In addition, it was observed that the NH_3/NO reacts at a stoichiometric ratio of less than or equal to 1.0 (Beeckman and Hegedus 1991; Lefers et al. 1991).

Although the SCR process is mature, worldwide application of SCR systems is growing with a speed slower than expected. This may be partly due to limited information available in combining experimental data with the proper design of an SCR system to reduce its cost. The SCR cost is so high that many of the industrial plants withdraw their willingness for installation. In addition, the reported data as well as theoretical design models in the literature were usually based on laboratory test results in ''clean air'' conditions without the presence of other air pollutants.

The same research group has developed theoretical models (Bai and Chwu 1997, 1998) for the design of a full-scale SCR reactor. Their models can be applied using either a numerical scheme that solves the 2D mass transfer equation (Bai and Chwu 1997) or a simple analytical solution of the 2D equation (Bai and Chwu 1998). Both models yield almost identical prediction results. However, because of the lack of field SCR performance data, both models were validated using laboratory experimental data only.

This paper intends to provide test data of a pilot study conducted in a glass manufacturing factory. The effects of important operation parameters such as the reaction temperature, space velocity and inlet NH_3/NO_x molar ratio on the NO conversion efficiency were evaluated. Results from the analytical design model of Bai and Chwu (1998) were compared to the experimental data, and the model was used to design the catalyst volume required for a full-scale reactor. The catalyst cost was then estimated based on the new design.

EXPERIMENTAL METHODS

A schematic of the experimental setup used in this study is shown in Fig. 1. Waste gas from a glass manufacturing plant was extracted and passed into a small-scale (0.18-m-wide, 0.18-m-high, and 0.41-m-long) SCR reactor. The SCR inlet gas flow rates were varied to obtain the desired space velocity in the reactor, and they were between 1.8 and 4.2 m^3/min. There were two honeycomb catalysts (0.15-m-wide, 0.15-m-high, and 0.20-m-long) installed in the SCR reactor. The V_2O_5/TiO_2 based catalysts were obtained from a local supplier (KJ

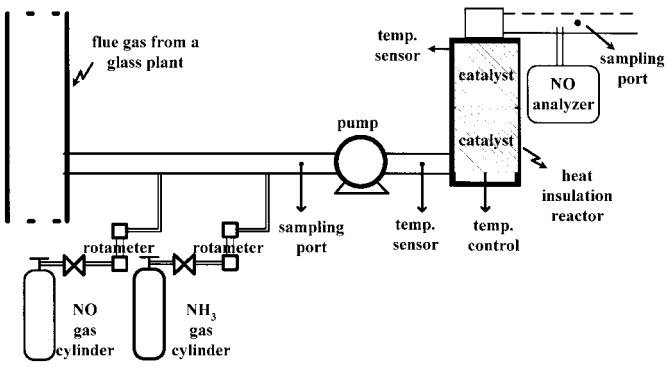

FIG. 1. Schematic of Experimental Setup Used in This Study

[1]Prof., Inst. of Envir. Engrg., Nat. Chiao-Tung Univ., No. 75 Po-Ai St., Hsinchu 300, Taiwan. E-mail: hlbai@green.ev.nctu.edu.tw

[2]Grad. Student, Inst. of Envir. Engrg., Nat. Chiao-Tung Univ., No. 75 Po-Ai St., Hsinchu 300, Taiwan.

[3]Grad. Student, Inst. of Envir. Engrg., Nat. Chiao-Tung Univ., No. 75 Po-Ai St., Hsinchu 300, Taiwan.

[4]Mgr., Air Pollution Control Lab., Waste Treatment Technol. Div., Energy and Resour. Lab., Industrial Technol. Res. Inst., Hsinchu 310, Taiwan.

[5]Engr., Air Pollution Control Lab., Waste Treatment Technol. Div., Energy and Resour. Lab., Industrial Technol. Res. Inst., Hsinchu 310, Taiwan.

Note. Associate Editor: Mark Rood. Discussion open until January 1, 2002. To extend the closing date one month, a written request must be filed with the ASCE Manager of Journals. The manuscript for this paper was submitted for review and possible publication on August 30, 2000; revised April 18, 2001. This paper is part of the ***Journal of Environmental Engineering***, Vol. 127, No. 8, August, 2001. ©ASCE, ISSN 0733-9372/01/0008-0735–0740/$8.00 + $.50 per page. Paper No. 22428.

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Environmental Technology Co., Ltd.) in Taipei, Taiwan. The cell size of the catalyst was 50 cell/in.^{2}, which corresponds to 0.3 cm of channel pitch.

Because the SCR reactor was located downstream of a spray dryer absorber and a bag house, the contents of SO_2 as well as particles were below the detection limits. The total fluoride (measured as F^-) and NO_x concentrations, as provided by the glass manufacturing plant, were 2.7 mg/m^3 and 105 ppm, respectively, at 128°C and 14.3% oxygen content. Because the NO_x content in the duct was low, an artificial NO source was introduced into the SCR reactor by means of an NO gas cylinder. The justified NO_x inlet concentrations were around 1,030 ± 80 ppm. An SO_x/NO_x analyzer (Testo Model 350, Germany) was used for continuous monitoring of inlet and outlet concentrations. The analyzer was precalibrated in the laboratory using standard gases. An ammonia gas cylinder coupled with a needle valve was used to provide the amount of ammonia required for reaction. The diameter of the duct for NH_3 and NO_x gases mixing was 7.5 cm. For the flow rates tested in this study, the flow Reynolds numbers in the mixing zone were >10^4. That is, the gas flow in the NH_3/NO mixing zone was turbulent and ensured good mixing for all tests. The ammonia inlet and outlet concentrations were verified by installing two midget impingers in series that contain 0.1 N H_2SO_4 solutions. The collected samples were later measured for the NH_4^+ concentration by an ion chromatography method (Dionex Model 2003I).

THEORETICAL MODEL

The results of the SCR performance tests (Svachula et al. 1993; Cho 1994) indicated that the main reactions in the SCR reactor are

$$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O \quad (1)$$

$$6NO_2 + 8NH_3 \rightarrow 7N_2 + 12H_2O \quad (2)$$

Reaction (1) is assumed to be dominant in the model of Bai and Chwu (1998). This is because NO contributes to >90% of NO_x in typical flue gas emissions from combustion processes. Therefore the NH_3/NO molar reaction ratio is 1.0. The theoretical model accounts for simultaneous diffusion, and chemical reaction in the SCR reactor is a 2D mass transfer equation

$$U\frac{\partial C_i}{\partial Z} = D_i\left(\frac{\partial^2 C_i}{\partial X^2} + \frac{\partial^2 C_i}{\partial Y^2}\right) \quad (3)$$

where U = flow velocity in the ducts; D_i = diffusivity of gas species i; i = NO or NH_3; C_i = concentration of gas species i; Z = main flow direction in the SCR reactor; and X and Y = directions perpendicular to the main gas stream. Eq. (3) is subject to the following boundary conditions:

$$Z = 0; \quad C_i = C_i^0 \quad (4)$$

$$X = 0; \quad \frac{\partial C_i}{\partial X} = 0 \quad (5)$$

$$Y = 0; \quad \frac{\partial C_i}{\partial Y} = 0 \quad (6)$$

$$X = \frac{d_h}{2}; \quad -D_i\frac{\partial C_i}{\partial X} = R_{NO} \quad (7)$$

$$Y = \frac{d_h}{2}; \quad -D_i\frac{\partial C_i}{\partial Y} = R_{NO} \quad (8)$$

where C_i^0 = inlet concentration of species i; i = NO or NH_3; d_h = channel pitch; and R_{NO} = surface reaction rate of NO. It may be difficult to find the complete analytic solution of (3); however, with the assumptions of plug flow and equal diffusivity of NH_3 and NO gas molecules, an analytic solution of (3) that describes the NO conversion efficiency η_{NO} is derived

$$\eta_{NO} = 1 - \frac{\beta^2}{\sin^2\beta}\exp(-2\beta^2 Z^*)\sum_{i=0}^{N}\sum_{j=0}^{N}\cos\left(\beta\frac{j}{N}\right)\cos\left(\beta\frac{i}{N}\right) \quad (9)$$

where Z^* = dimensionless main flow direction ($Z^* = 4LD_{NO}/U_m d_h^2$); L = reactor length; and U_m = average flow velocity. The definition of β is

$$\beta\tan(\beta) = \frac{Da}{2}\phi^*_{NH_3} \quad (10)$$

where Da = Damköhler number ($Da = K_S d_h/D_{NO}$) and

$$\theta^*_{NH_3} = \frac{K^*_{NH_3}C^*_{NH_3}}{1 + K^*_{NH_3}C^*_{NH_3}} \quad (11)$$

where $\theta^*_{NH_3}$ = percentage coverage of ammonia; $K^*_{NH_3}$ = dimensionless adsorption constant of ammonia ($=K_{NH_3}C^0_{NO}$); and $C^*_{NH_3}$ = dimensionless NH_3 concentration ($=C_{NH_3}/C^0_{NO}$). The assumption of equal diffusivity of NO and NH_3 leads to negligible errors in terms of NO conversion efficiency. Detailed results of the assumptions are referred to Bai and Chwu (1998). The plug flow assumption yields maximum deviations of around 10% in the NO conversion efficiency as compared to the laminar flow model. And the deviation can be corrected by a correction factor (Bai and Chwu 1998).

There are two unknown parameters that must be determined from experimental data, the surface reaction rate constant K_S and the ammonia adsorption constant K_{NH_3}. These two constants are functions of temperature and can be expressed by the Arrhenius equation and the adsorption equation

$$K_S = k_0\exp\left(-\frac{E_a}{R_g T}\right) \quad (12)$$

$$K_{NH_3} = K_0\exp\left(-\frac{\Delta H}{R_g T}\right) \quad (13)$$

where E_a = activation energy; ΔH = heat of adsorption; k_0 and K_0 = preexponential factors in the Arrhenius equation and the adsorption equation, respectively; T = reaction temperature; and R_g = gas constant (=8.314 J/mol °K).

RESULTS AND DISCUSSIONS

Determination of Reaction Rate and Adsorption Constant

Using a series of field test data at various operating temperatures, the surface reaction rate constant K_S and the NH_3 adsorption constant K_{NH_3} were determined by a least-squares linear regression method using (12) and (13). For detailed procedures in using the theoretical model to obtain K_S and K_{NH_3} refer to Bai and Chwu (1997). And the two constant values obtained in this study are described as follows.

Surface Reaction Rate Constant

To reduce the uncertainty in determining two unknown parameters at the same time, the value of K_S was obtained under the NH_3/NO inlet molar ratio of >1.0. This is because that, for the NH_3/NO inlet molar ratio of >1.0, the effect of K_{NH_3} becomes negligible (Bai and Chwu 1997). Therefore only K_S is the dominent factor influencing the NO conversion efficiency of the SCR process. The experimental data were at the NH_3/NO inlet molar ratio of 1.0–1.4, operating temperature range of 250°–400°C, and gas hourly space velocity of 13,550 h^{-1}. The results are shown in Fig. 2(a). As a result of linear regression based on (12), the following is obtained:

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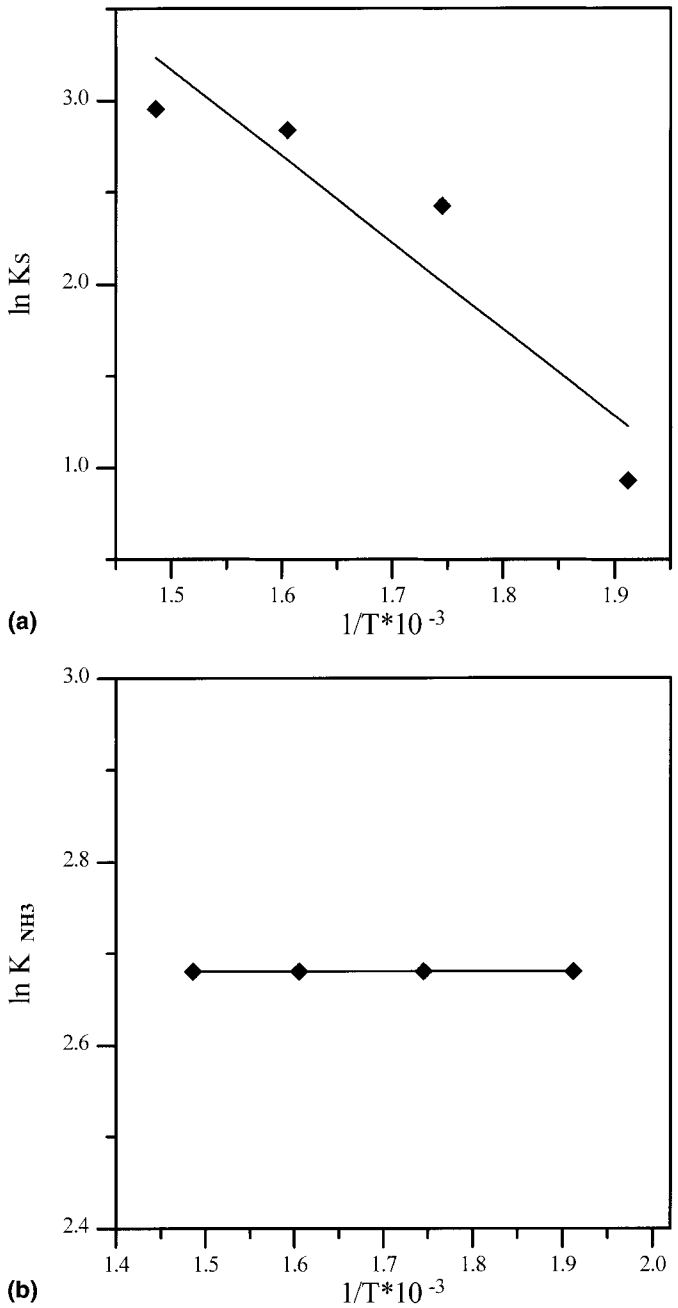

FIG. 2. Variation of: (a) Surface Rate Constant K_S (cm/s) with Respect to Temperature (°K) (Symbols Are Average Value of K_S Obtained from Experimental Test and Straight Line Is Linear Regression Result; R^2 Value for Linear Regression Was 0.86 for Total of Eight Experimental Data); (b) Dimensionless NH_3 Adsorption Constant $K^*_{NH_3}$ with Respect to Temperature (°K) (Symbols Are Average Value of K_{NH_3} Obtained from Experimental Test and Straight Line Is Linear Regression Result; R^2 Value for Linear Regression Was 1.0 for Total of Eight Experimental Data)

$$K_S = 2.76 \times 10^4 \exp\left(-\frac{4.71 \times 10^3}{T}\right) \quad \text{(cm/s)} \qquad (14)$$

Ammonia Adsorption Constant

After determining the equation for the surface reaction rate constant, the ammonia adsorption constant is then calculated. The experiments were done at an NH_3/NO inlet molar ratio of 0.2–1.2, operating temperature range of 250°–400°C, and gas hourly space velocity of 13,500 h^{-1}. The results are shown in Fig. 2(b). As a result of linear regression based on (13), it is seen that the ammonia adsorption constant is not a function of temperature for the temperature range used in this study. The dimensionless value of the ammonia adsorption constant $K^*_{NH_3}$ is thus

$$K^*_{NH_3} = 14.58 \qquad (15)$$

The dimensionless value of $K^*_{NH_3}$ corresponds to a value of 2.85×10^5 cm^3/mol for K_{NH_3} at 350°C. This falls in the range of 10^5–10^8 cm^3/mol for values of K_{NH_3} found in the literature (Miyamoto et al. 1982; Tronconi et al. 1994; Koebel and Elsener 1998; Santos et al. 1998). As a result, the percentage coverage of ammonia at the catalyst surface $\theta^*_{NH_3}$ is close to 100%. Therefore ammonia adsorption is not a controlled factor for the catalyst used in this study.

Sensitivity Analysis of K_S

The experimental data of K_S at the temperature range of 200°–300°C were also regressed, and the regression results of K_S are listed in Table 1 together with the base case (250°–400°C) regression results. One can see from Table 1 that the regression coefficient R^2 for the base case (250°–400°C) model is 0.86 for a total of eight experimental data. But the regression coefficient is 0.997 for the 200°–300°C model.

The average experimental observation of K_S is 17.08 cm/s at 350°C. It is seen that K_S obtained from the base case (250°–400°C) model at 350°C (K_S = 14.37 cm/s) agrees well with the experimental observation. The error in K_S for the base case model is about 16%. The regression coefficient of the 200°–300°C model is very high; nevertheless, a large error was observed for its K_S at 350°C (=54.90 cm/s) as compared to the experimental observation. On the other hand, the average experimental observation of K_S is 2.54 cm/s at 250°C. And one can see that the predicted K_S at 250°C from the 200°–300°C model is more close to the average experimental observation than that from the base case (250°–400°C) model.

Comparisons of Model Results and Experimental Data

NH_3/NO Inlet Molar Ratio

The experimental results for the effect of the NH_3/NO inlet molar ratio on the NO conversion efficiency are shown in Fig. 3. The operation condition was at the temperature of 350°C, gas hourly space velocity of 13,500 h^{-1}, and NO inlet concentration of 1,030 (±6%) ppm. The predicted results from both the base case (250–400°C) model and the 200°–300°C model are also shown in Fig. 3. It is seen that both models yield similar results. And for 0.5 < NH_3/NO < 1.0, the NO conversion efficiency is larger than expected by the model predictions. That is, for every 1 mol of NH_3 injected, the NO reduction is >1 mol. The developer of the catalyst used in this study indicated that this is also a common phenomenon in his field applications of the SCR system (R.-Y. Weng, personal communication, 2000) However, it conflicts with observations in the literature of laboratory experiments (Beeckman and Hegedus 1991; Lefers et al. 1991) that there was at most 1 mol of NO reacted with 1 mol of NH_3. Measurements of NH_3 outlet concentration in this study also showed that ammonia slip is observed only for NH_3/NO > 1.0. Hence another possible reaction between NO and NH_3 may occur (Cho 1994)

$$6NO + 4NH_3 \rightarrow 5N_2 + 6H_2O \qquad (16)$$

Although (16) may occur for the range of 0.5 < NH_3/NO < 1.0, deviations of model results from the experimental data are acceptable for designing a full-scale SCR system.

Operating Temperature

The experimental results for the effect of operating temperature on the NO conversion efficiency are shown in Fig. 4.

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TABLE 1. Regression Results of Effective Rate Constant K_S Based on Data of Different Temperature Ranges

	Regression Parameters for K_S					
Temperature range	k_0 (cm/s)	Ea (KJ/mol)	R^2	Number of data	K_S at 350°C[a] (cm/s)	K_S at 250°C[b] (cm/s)
Base case (250°–400°C)	2.76×10^8	39.13	0.860	8	14.37	3.39
200°–300°C	5.47×10^4	83.47	0.997	6	54.90	2.52

[a]Average experimental value of K_S at 350°C was 17.08 cm/s.
[b]Average experimental value of K_S at 250°C was 2.54 cm/s.

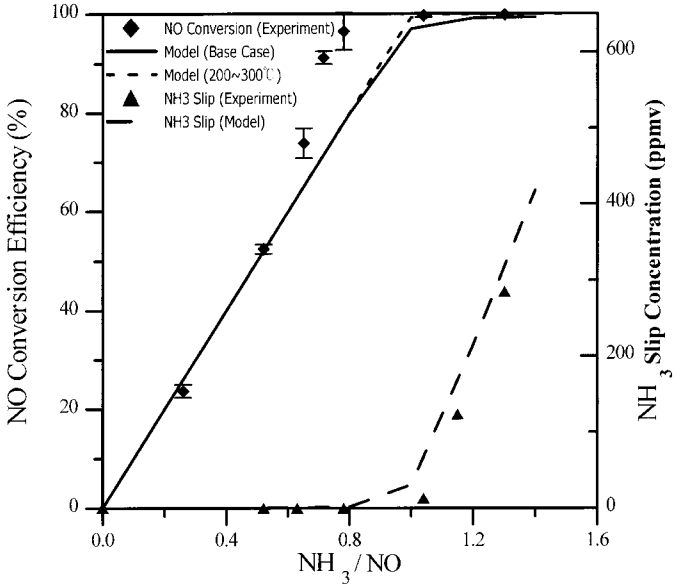

FIG. 3. Experimental Observations and Model Results on NO Conversion Efficiency as Function of NH_3/NO Inlet Ratio (Experimental Condition Was at T = 350, NO = 1,030 ppm (±6), and SV = 13,550 h^{-1})

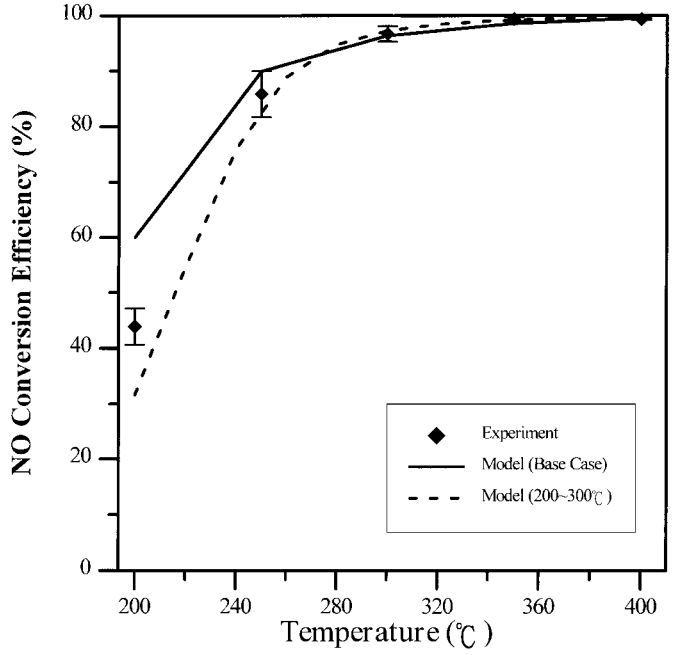

FIG. 4. Experimental Observations and Model Results on NO Conversion Efficency as Function of Temperature (Experimental Condition Was at NO = 1,030 ppm (±6), SV = 13,550 h^{-1}, and NH_3/NO = 1.0)

The operation condition was at an NH_3/NO inlet molar ratio of 1.0, gas hourly space velocity of 13,550 h^{-1}, and NO inlet ratio of 1,030 (±6%) ppm. It is seen that >95% NO conversion efficiencies were achieved for temperature ranges from 300° to 400°C. The base case model predicts the experimental data well at temperatures larger than around 250°C. And it tends to overpredict the NO conversion efficiency at temperatures below that. But this is not a concern because the SCR is usually designed at high temperature operation, where high NO_x conversion efficiency is usually encountered. On the other hand, the 200°–300°C model tends to underpredict the NO conversion efficiency at low temperatures.

Gas Hourly Space Velocity

The gas hourly space velocity SV is defined

$$SV = Q/V \tag{17}$$

where Q = gas flow rate (m^3/h); and V = volume of the catalyst bed (m^3). It is seen from (17) that increasing the gas flow rate or reducing the catalyst volume tends to increase the space velocity and thus reduce the residence time of waste gas. The typical operating range of the gas hourly space velocity for a honeycomb catalytic reactor is 2,000–7,000 h^{-1} (standard temperature and pressure) (Beeckman and Hegedus 1991).

Fig. 5 shows the experimental results as well as the model predictions for the effects of space velocity on the NO conversion efficiency. The operation condition was at an NH_3/NO inlet molar ratio of 1.0, NO inlet concentration of 1,030 (±6%) ppm, and reaction temperature of 350°C. One can see from the experimental results that for space velocity of <13,550 h^{-1}, the NO conversion is nearly completed. A further increase in the space velocity leads to a slow decrease in the NO conversion efficiency. The value of the space velocity for a high NO conversion efficiency is much larger than typical values used in the field. That is, the catalyst volume required

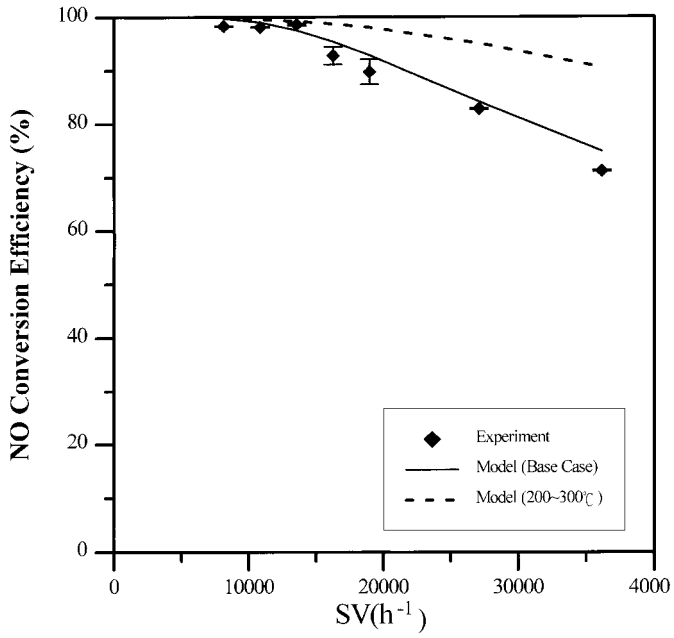

FIG. 5. Experimental Results and Model Results on NO Conversion Efficiency as Function of Gas Hourly Space Velocity (Experimental Condition Was at T = 350, NO = 1,030 ppm (±6), and NH_3/NO = 1.0)

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for a high NO conversion efficiency is much smaller without considering any safety factor. If the safety factor is considered, the catalyst volume will have to be increased to ensure a long, useful life for the SCR catalyst. This will be discussed later.

One also can see that as the space velocity increases, deviations between model predictions and experimental data are increased. A maximum deviation of <5% in terms of NO conversion efficiency occurs at a space velocity of 36,000 h^{-1} for the base case (250°–400°C) model, but it is 20% different in the NO conversion efficiency using the 200–300°C model. Because space velocity is the dominant factor for the design of the catalyst volume, errors encountered in the prediction of the space velocity must be considered very carefully. So, if the base case model was used, the safety factor is around 1.10 to account for the difference in the deviation between predicted and actual space velocity for achieving the same NO conversion efficiency. But if the 200°–300°C model was used in the design of the catalyst size, a safety factor of around 2.5 must be considered to account for the deviation. Hence, it is obvious that, although the linear correlation for the 200°–300°C model is very high, the prediction at a typical operating temperature of 350°C is not very good. Therefore it is concluded from previous discussions that a good design model depends not only on a high linear correlation of the rate constant K_S equation but also on an accurate prediction of the K_S value at the specified temperature where the SCR system is to be operated.

Practical Application—A Case Study

Redesign of Catalyst Volume

An industrial plant that had an SCR system installed was used as a study case. The plant's SCR reactor uses catalysts imported from Germany. It is intended to see if redesigning the SCR system with a locally developed catalyst can significantly reduce the catalyst cost. The original design was at a space velocity of 3,392 h^{-1} under a gas flow rate of 32,220 m^3/h and NO inlet concentration of 1,067 ppm. The corresponding catalyst volume was 9.5 m^3 with a target NO_x conversion efficiency of 82% for 3 years of operation. Because one is not able to obtain the original catalyst's performance data, it cannot be known if the original catalyst reactor would give NO conversions of >82%.

By using the two designed-parameter values obtained from the base case (250°–400°C) model, the new catalyst volume and its NO outlet concentration were calculated. The results are shown in Fig. 6. It is observed that, if the new SCR reactor were operated at a space velocity of 3,392 h^{-1} (i.e., the same catalyst volume as the original reactor), the NO conversion efficiency can be as high as 99%. And if the design is to achieve an NO conversion efficiency of only 82%, then the space velocity can be as high as 28,000 h^{-1}. In other words, the SCR reactor can be reduced to only 12% of its original size without considering any safety factor. If considering the safety factor of 2.0 that is typically used in the field, the size of the SCR reactor is still significantly lower.

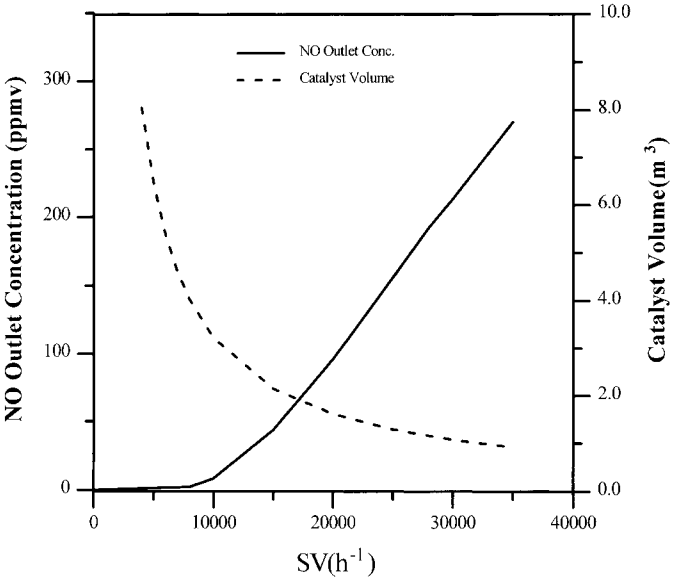

FIG. 6. Design of SCR Reactor Volume and Outlet NO Concentration under Different Values of Space Velocity

Catalyst Cost

Table 2 lists the comparison of the catalyst cost of the original design versus the new design using the base case (250°–400°C) model. The safety factors of 1.0–3.0 for the catalyst bed size were employed in the new design. The original catalyst cost was \$17,400/m^3. Thus the total cost of a 9.5-m^3 catalyst bed was \$165,000. On the other hand, the catalyst cost is \$13,600/m^3 for the new catalyst, so the total catalyst cost for the safety factor of 1.0 (corresponding to a 1.15-m^3 catalyst bed size) is only \$15,600. This is only one-tenth of the original price. And for the safety factor of 3.0, the difference in the total cost of the catalyst bed is still significant.

The disadvantage of the locally developed catalyst is that its first field application started about 2 years ago. Although the first SCR application still functions well without any SCR catalyst replacement, it cannot be known if the new catalyst works as long as the original catalyst. Hence, a safety factor of 5.0 is considered to ensure a long useful life for the new catalyst. And it is seen from Table 2 that, even for the safety factor of 5.0, the catalyst cost is only one-half the original catalyst cost.

Flow Rate Effect

The flow rate of a plant's waste gas emission may vary from time to time. Therefore once the SCR is installed, it is necessary to evaluate in advance the variation of NO conversion efficiency when the flow rate changes. Fig. 7 shows the effect of flow rate variation on the NO conversion efficiency at different values of the safety factor. It is seen that for a safety factor of 1.0, the NO conversion efficiency changes from >95% to around 70% as flow rate changes from 50 to 150% of its original value Q_0. And for a safety factor of 2.0, the

TABLE 2. Comparison of Catalyst Costs between Original Design and New Design

Design	Safety factor	Bed volume (m^3)	Space velocity (h^{-1})	Catalyst cost (dollars/m^3)	Total catalyst cost (dollars)
Original	—	9.5	3,392	17,400	165,000
New	1.0	1.15	28,000	13,600	15,600
	2.0	2.30	14,000	13,600	31,300
	3.0	3.45	9,333	13,600	46,900
	5.0	5.75	5,600	13,600	78,000

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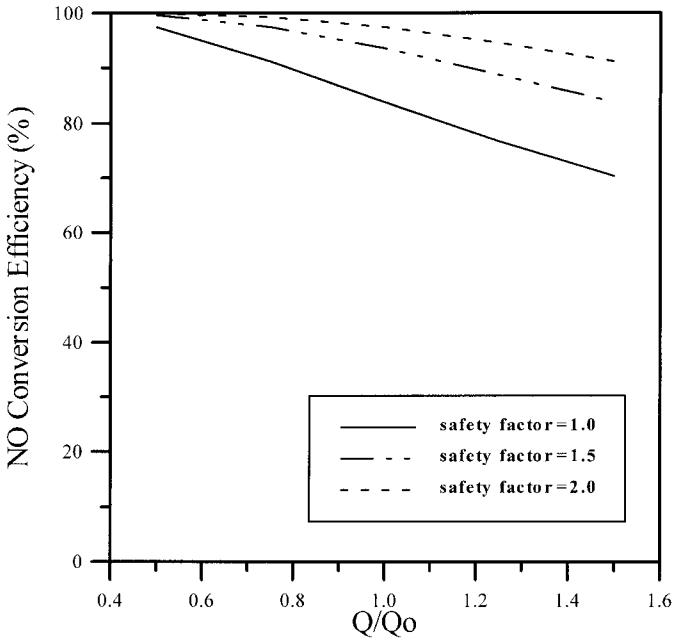

FIG. 7. Effect of Gas Flow Rate on NO Conversion Efficiency

deviation in NO conversion efficiency is <10%. The NO conversion efficiency can be well above 90% for flow rate changes from 50 to 150% of its original value.

CONCLUSIONS

This paper presented results on a pilot study of an SCR system installed in a glass manufacturing factory. It showed that the NO conversion efficiency is >95% for a gas hourly space velocity of <13,550 h^{-1}, which is much larger than typical values used in the field. And for an NH_3/NO inlet molar ratio of >0.5, >1 mol of NO is reacted for every 1 mol of NH_3 injected. This may be due to another NO—NH_3 reaction that occurred. The values of design parameters were obtained using the experimental data. The theoretical model was then used to redesign an SCR reactor installed in an industrial plant. The results showed that the catalyst used in this study has a better performance than the original catalyst. Thus the catalyst bed size as well as the total catalyst cost can be significantly reduced.

However, the question arises, does the catalyst used in this study really perform so superior to the original catalyst? This may be somewhat true because some progress on the SCR catalyst has been made by the local supplier. But the chemical composition and the physical structure of the SCR catalyst cannot be provided because of a confidential agreement with the local supplier. Nevertheless, it is believed that, because of a lack of a proper design model, the SCR system suppliers would rather design the catalyst volume with a large safety factor to ensure its performance. This is observed from the case studied. The plant's SCR catalyst was to be replaced every 3 years, but in fact it has been operated >5 years without the need for any catalyst replacement. As a result, the annualized SCR cost can be reduced because the catalyst accounts for a significant portion of the overall SCR cost. So the SCR system may be thought to be more expensive than it should be.

ACKNOWLEDGMENTS

Support from National Science Council (Taipei, Taiwan) of the Republic of China through grant NSC-89-2211-E-009-047 is gratefully appreciated. Partial support from the Industrial Technology Research Institute is also acknowledged. The writers would like to thank Dr. R.-Y. Weng at the KJ Environmental Technology Co., Ltd., for providing the SCR catalysts.

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J. Environ. Eng. 2001.127:735-740.