 Removal of SO₂ in Dry Fluidized and Fixed Bed Reactors using Granular Activated Carbon

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Received on: 8/12/2010
Accepted on: 5/5/2011

Abstract
Flue gas desulfurization process has been studied using granular activated carbon in bubbling fluidized bed reactor. For the sake of comparison, fixed bed configuration has been also studied at the same operating conditions. The effect of temperature 30 ≤ T ≤ 80 °C, inlet SO₂ concentration 500 ≤ C₀ ≤ 2000 ppm, and flue gas flow rate 2.5 ≤ Q ≤ 30 ℓ/min were investigated. The results showed that the SO₂ removal efficiency increases with increasing reaction temperature up to 80 °C. Also, it was noted that the removal efficiency decreases with increasing the inlet SO₂ concentration within the range of temperatures studied. The effect of flue gas flow rate on the desulfurization activity was in two ways; an increase in the removal efficiency with increasing gas flow rate was observed below flow rate = 7.5 ℓ/min, while a decrease in the efficiency was observed upon any increment in the gas flow rate beyond the 7.5 ℓ/min. The results of fluidized bed reactor were used to obtain an empirical correlation and the experimental results were well correlated with the proposed form with a correlation coefficient, (R) =0.989. A reaction rate equation was proposed for the oxidative desulfurization and the activation energy was obtained using differential analysis of integral reactor technique. The resulted value of apparent activation energy was 2.981 kJ/mol.

Keywords: SO₂ Removal, Activated Carbon, Fluidized & Fixed Bed Reactors

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1. Introduction

Sulfur dioxide (SO₂) is considered among the most toxic gases emitted to the atmosphere during the combustion of fossil fuels, for example, coal and diesel oil [1]. Sulfur dioxide can cause acid rain, which has acidified soils and streams, accelerated corrosion of buildings and reduced visibility. Long-term exposure to SO₂ also results in various respiratory diseases [2]. Among the famous emissions reducing methods is flue gas desulfurization (FGD). A variety of flue gas desulfurization techniques are available for SO₂ removal from flue gases. These can be broadly classified into two general categories: regenerable and non-regenerable processes. These can be subdivided into wet and dry processes [2].

It has been shown in the literature that activated carbon act simultaneously as adsorbent and as catalyst for SO₂ oxidation even at low temperatures [3]. The removal of SO₂ over activated carbon in the presence of oxygen and water vapor at relatively low temperatures (20-150 °C) involves a series of reactions that leads to the formation of sulfuric acid as the final product. The overall reaction is [4]:

\[ \text{SO}_2 + \frac{1}{2} \text{O}_2 + \text{H}_2\text{O} \rightarrow \text{H}_2\text{SO}_4 \] (1)

For flue gas desulfurization, the fluidized bed reactor has the following advantages over fixed bed reactor:

1. It can prevent the plugging phenomenon that can occur in a fixed-bed reactor as fly ash passes through the reactor.
2. High thermal /mass transfer coefficient, high gas/solid, and solid/solid contact areas are continuous.
3. It does not require a reactor shut down for the replacement of the adsorbent [5].

Lee [6] studied SO₂ removal with char from flash carbonization process as a potential adsorbent. A fluidized bed was employed. He found that increasing the velocity reduces the mass transfer resistance (both pore and external film resistance) by compression forces. He also reported that removal efficiency decreased when the velocity of gas increased.

Rau et al. [5] studied the removal of acid SO₂ gas by oxidation on activated carbon in fluidized bed reactor. They used activated carbon with a diameter of 1 mm classified as Geldart B powders, with constant flue gas velocity of \( \frac{u}{u_{mf}} = 1.47 \). The results of their work indicated that for activated carbon treated with HNO₃ removal efficiency reaches over 84%. They concluded that the fluidized bed adsorbent/catalyst reactor have a high potential for simultaneous removal of high concentrations of SO₂ and fly ash from flue gases. Cho and Miller [7] worked on the desulfurization by activated coal and found the temperature to influence the average removal efficiency by increasing it upon increasing the temperature. The profiles of curves of average removal efficiency against reaction temperature were exponential in shape in their work. Liu et al. [8] used waste semi-
coke as the raw material to prepare carbon catalyst for SO₂ removal. They found the dependence of sulfur dioxide retention on space velocity was of two types. Sulfur dioxide retention increased from 7.4 to 9.6 g SO₂/100 g C with increasing space velocity from 540 to 830 h⁻¹, however, further increasing the space velocity decreased the sulfur retention. Thus, they concluded that the kinetic behavior varied with space velocity and the desulfurizing property was controlled by diffusion at space velocities below 830 h⁻¹, and controlled by the oxidation step (kinetic step) at space velocities above 830 h⁻¹. Most recently Li et al. [9] have confirmed the results of Liu et al. [8], where similar results were obtained and desulfurization activity was controlled by diffusion below 800 h⁻¹ and controlled by the oxidation reaction above 800 h⁻¹. Mochida et al. [10] conducted a kinetic study of the removal of SO₂ using polyacrylonitrile-based activated carbon fibers. They developed a power-law model to describe their data and obtained the exponents of concentration terms in the model. The exponent of sulfur dioxide was between 0 and 1. Gaur et al. [11] examined the effect of concentration on the breakthrough curves. It was found that the outlet concentration of the SO₂ expectedly increased with increase in the inlet SO₂ concentrations, indicating relatively earlier saturation of the AC at higher inlet SO₂ concentrations.

Raymundo-Pinero et al. [12] Lizzio and DeBarr [13], and Otake et al. [14] have proposed the SO₂ oxidation to be the rate-limiting step. Following the assumption that the rate-determining step in the overall reaction is catalytic oxidation of SO₂ to SO₃, an overall rate expression was proposed by Lizzio and DeBarr [13]:

\[-r_{SO_2} = k \left[ C_{SO_2} \right]^n \left[ C_A \right] \]  

(2)

Where \( C_A \) denotes the concentration of the free active sites, \( k \) is rate constant, and \( n \) is the order of reaction with respect to SO₂. The rate is then only a function of the concentration of free sites and SO₂. For concentrations of SO₂ <1500 ppm, it was found that values of \( n \) between 0.5 and 1, and for concentrations of SO₂ >1500 ppm, the value of \( n \) approached 0. Up to concentration of SO₂ of approximately 500 ppm the value of \( n \) was 1. For higher concentrations, the value of \( n \) decreased [13].

The overall objective of this research is to investigate the performance of granular activated carbon for SO₂ removal from simulated flue gas in fluidized bed reactor and comparison with that of fixed bed reactor at the same operating conditions. The operating parameters are: flue gas flow rate [(2.5-12.5) l/min for fixed bed and (13.5-30) l/min for fluidized bed], reaction temperature (30-80) °C, and inlet SO₂ concentration (500-2000) ppm.

2. Experimental Work

The experimental runs were performed in a laboratory-scale apparatus. The general layout of the experimental apparatus is displayed in Figure (1). The experimental rig consists of three sections; SO₂ generation section, reaction section, and analysis section. The reaction
section is a QVF column of 1 in diameter and 50 cm long and contains activated carbon bed to a height of 12 cm and weight of 30 gm with particle diameter dp = 1.09 mm. The physical properties of the activated carbon were measured using surface area analyzer and are presented in Table (1). Heat is supplied to the reaction section via electric heater of 60 Watt power. A T-type thermocouple fitted along the axis of the reactor reaching the middle of the packing measured the temperature of the bed. The thermocouple sends its signal to a PID temperature controller. The power consumed by the heater was controlled manually by varying the voltage of the current by means of Variac transformer.

2.1 Experimental Procedure

The reaction section was charged with 30 gram of activated carbon with an average particle diameter of (1.09) mm which gave a bed height of 12 cm above the distributor plate. This charge was replaced with fresh one after each test to ensure the bed activity is maintained constant. In the generation vessel, 300 gram of Sodium Sulfite powder was placed for each run. The analysis section was filled with 400 ml of standard Iodine solution (0.1 N) which was replaced at the end of each run. The controller was set to the desired temperature and the variac transformer was adjusted to control the amount of the power supplied depending on the gas flow rate and temperature required. Sulfur dioxide gas was synthesized by dropping the desired concentration of sulfuric acid solution into the SO₂ generation vessel that contains sodium sulfite powder according to the chemical reaction:

\[ \text{Na}_2\text{SO}_3 + \text{H}_2\text{SO}_4 \rightarrow \text{SO}_2 + \text{Na}_2\text{SO}_4 + \text{H}_2\text{O} \]  

Air stream at the desired flow rate carries the SO₂ to the reaction section. The outlet gas stream (from the top of the bed) was introduced into iodine solution to absorb the residual sulfur dioxide gas according to the reaction:

\[ \text{SO}_2 + \text{I}_2 + \text{H}_2\text{O} \rightarrow 2\text{HI} + \text{H}_2\text{SO}_4 \]  

In order to measure the concentration of SO₂ in the effluent gas stream, (10) ml of the iodine sample was taken from the absorption trap, titrated with standard sodium thiosulfate solution (0.1 N), in the presence of starch indicator. This step was repeated every 5 minutes till the end of experimental run time (1 hr). Each experiment was repeated twice to validate the experimental results. The removal efficiency of SO₂ was calculated as the ratio of SO₂ concentration that was removed by activated carbon to the initial concentration of SO₂ gas fed to the bed:

\[ \eta (\%) = \frac{c_0 - c}{c_0} \times 100 \]  

The average removal efficiency of SO₂ was calculated as the average of the values of removal efficiencies obtained at each 5 min throughout a run.

\[ \eta_{av} = \frac{\sum \eta}{12} \]  

3. Results and Discussion

The variation of SO₂ removal efficiency with flow rate is shown in Figure (2) at different operating temperatures and constant initial concentration of SO₂ of 500 ppm, for one hour duration time for fluidized bed operation.
The purpose of these experiments was to determine the best flue gas flow rate value in fluidized bed that gives higher removal efficiency in the range of flow rates studied. Although higher average removal efficiency of 96% was obtained at $Q = 13.5 \ l/min$, this value was slightly above the minimum fluidizing velocity. Bubbling fluidization was clearly visualized at $Q = 15 \ l/min$. This value was used throughout the ongoing sets of experiments of the effect of reaction temperature and initial $SO_2$ concentration as bubbling fluidization is the regime understudy in the present work.

As shown in Figure (2), on increasing the flue gas flow rate the removal efficiency decreases, this can be explained by the decrement of $SO_2$ residence time in bed thus decreasing the amount of $SO_2$ retained by the surface of carbon particles which depends mainly on the exposure time in the vicinity of the inner carbon surface that decreases with high velocities. These results agree with that of Lee [6].

Figure (3) shows the dependence of average removal efficiency of sulfur dioxide on reaction temperature for different $SO_2$ inlet concentrations at the constant flue gas flow rate of $15 \ l/min$ in fluidized bed. Sulfur dioxide removal efficiency significantly increased with increasing temperature and reached highest value at $T = 80 ^\circ C$ for each of the four inlet $SO_2$ concentrations examined. The slowest step in the surface steps is the $SO_2$ catalytic oxidation performed by surface groups [12,13,14]. Thus $SO_2$ interacts with surface groups due to the strong attractive forces exerted by these groups that capture $SO_3$ and catalyze its oxidation reaction to $SO_3$. Increasing the temperature will increase the rate of formation of $SO_3$ due to the faster oxidation of $SO_2$ as a consequence of rapid activation of the $SO_2$ molecule that will react in a shorter period of time at the increased temperatures. $SO_3$ formed is less volatile, strongly adsorbed on carbon surface [15], so that $SO_3$ do not require the attractive forces of the high energy adsorption sites to hold it on surface. After oxidizing an $SO_2$ molecule, the functional groups are freed to attract successive $SO_2$ molecule and oxidizing it, which increases the amount of $SO_2$ uptaked. Any increase in temperature would have its effect on the reaction rate thus increases the rate constant $k$, as it is evidenced by the likeness of profiles of the curves, which show that the average removal efficiency has an exponential-like dependence on reaction temperature emerging from obeying Arrhenius equation in its variation with temperature. These results agree with those of Rau et al. [5] who reached maximum value of removal efficiency of 84% at optimum reaction temperature of 200 $^\circ C$, and with the results of Cho and Miller [7].

Effect of inlet $SO_2$ concentrations on $SO_2$ removal efficiencies for various reaction temperatures is shown in Figures (4) to (6) for fluidized bed operation. These figures show that $SO_2$ removal efficiency decreased as the concentration of $SO_2$ increased at different operating temperatures. This is attributed to the order of oxidation reaction ($n$), which is the rate determining step, being less than 1 with
respect to SO$_2$ (when the initial concentration roughly exceeds 500 ppm according to Lizzio and DeBarr [13] so that the gas removal efficiency is decreased with increased concentration. Thus it can be concluded from these results along with the findings of Lizzio and DeBarr [13] that the desulfurization efficiency as well as the order of dependence of reaction rate ($-\dot{r}_{SO_2}$) on SO$_2$ concentration, are both functions of inlet SO$_2$ concentration. The removal efficiency is inversely proportional to the inlet SO$_2$ concentration, but directly proportional to the weight of carbon bed [6].

The effect of flue gas flow rate on SO$_2$ removal efficiency at constant inlet concentration of 500 ppm and different temperatures for fixed bed reactor (combined with that of fluidized bed reactor) is shown in Figure (7). The results show that as the gas flow rate increases the removal efficiency increases up to a value of $Q = 7.5$ l/min at which maximum removal efficiency is reached, thus flow rate of 7.5 l/min is the optimum flue gas flow rate. Beyond this value any increase in gas flow rate will lower the removal efficiency.

At lower flue gas flow rates (below the optimum flow rate), the desulfurizing behavior is controlled by diffusion of reactant SO$_2$ into the inner surface of the sorbent. In this way, increasing the gas flow rate will increase the diffusion rate as a result of the increased forces of convection. The retention times at these flow rates in the vicinity of active surfaces are long enough to convert adsorbed SO$_2$ to H$_2$SO$_4$, due to the minimized diffusion resistance. Hence, sulfur dioxide retention increases with flow rate up to the optimum flow rate (7.5 l/min), where starting from this value the desulfurizing behavior is possibly controlled by oxidation step (i.e. kinetic step) because of completely minimized diffusion resistance. Due to the reduction of retention time at higher velocities (at or above the optimum flow rate), reactants cannot have the sufficient time for oxidation of large amount of SO$_2$ to form SO$_3$ to take place, which leads to the decrease of sulfur dioxide retention[8].

Referring to Figure (7), it can be recognized two regions of effect of gas flow rate on removal efficiency suggesting that there are two different ways by which gas flow rate can change the efficiency. The experimental runs can be divided according to either of the two regions of Figure (7) they belong to, either into the region below the optimum flue gas flow rate of 7.5 l/min, which is the case of fixed bed reactor, or the region at or above the optimum flow rate in which all mass transfer resistance due to the accessibility obstacles to the porous structure is well overcome (because of high velocities) and gas molecules is flushed through the carbon, which is the case of the highest velocities in fixed bed reactor, and fluidized bed reactor. According to Gauthier [16] the elimination of diffusional limitations (internal diffusion limits inside particles), gives rise to an effectiveness factor very close to unity. This makes the oxidation step controls at or above the
7.5 l/min, where the global rate will be the same as reaction rate. These results are consistent with those of Li et al. [9], Liu et al. [8].

Figure (8) shows the effect of temperature on the average removal efficiency of SO\(_2\) at the optimum flue gas flow rate of 7.5 l/min for different inlet SO\(_2\) concentrations in fixed bed conditions. As shown in the figure, increasing temperature increased the removal efficiency and approached 100% removal at 80 °C and 500 ppm. Thus temperature has positive effect on the desulfurization process in that it increases the reaction rate. An increase in removal efficiency on increasing reaction temperature was also observed by many investigators [7,8,9,14].

The trend of the results shows a similar pattern as that obtained in fluidized bed, thus temperature affects in the same manner in both reactors emerging from same reaction steps in both reactors. As in fluidized bed results, the exponential-like behavior is also observed in Figure (8) for fixed bed operation.

Figures (9) to (11) illustrate the effect of inlet SO\(_2\) concentrations on its removal efficiencies at various reaction temperatures for fixed bed operation, with the gas flow rate kept constant at its optimum value of 7.5 l/min. The removal of SO\(_2\) decreased with increasing the inlet SO\(_2\) concentrations, indicating the suitability of the current amount of sorbent to raise the value of SO\(_2\) conversion with rising inlet concentrations only for low levels of inlet concentration [11]. At 500 ppm inlet gas concentration, nearly complete removal was observed, the removal efficiency decreased to 97% at the inlet concentration of 1000 ppm both at the best oxidative reaction temperature obtained for this work of 80 °C (Figure (11)). Removal efficiency was observed to decrease to its lowest values at all temperatures tested corresponding to the inlet gas concentration of 2000 ppm.

Gaur et al. [11] showed that the exit gas concentration of SO\(_2\) gradually increased at higher inlet SO\(_2\) concentration suggesting inadequacy of bed for high concentrations. The amount of sorbent existing is fixed while the inlet SO\(_2\) concentration is continuously increasing. This is the reason of why the reaction order is less than 1 at high concentrations. This idea of the change of the order of dependence of reaction rate on SO\(_2\) concentration, with increased initial concentration of SO\(_2\) due to consumption or decaying of free sites was first suggested by Lizzio and DeBarr [13]. A decrease in removal capacity upon increasing SO\(_2\) concentration was observed by Mochida et al. [10].

### 3.1 Empirical Correlation

Based on the experimental results obtained in the fluidized bed reactor, an empirical correlation that includes the effect of operating variables on the removal efficiency was obtained in this study. The correlation incorporates the effect of temperature in terms of exponential-like dependence, and the gas velocity which is the crucial variable in fluidized beds. Gas velocity has effects on the removal process through affecting residence time in the
bed as well as the efficiency of contact between particles and gas during the residence time. The first term is well described by a dimensionless velocity. It is well proven that the particle Reynolds number \( \text{Re}_p \) is a good characteristic dimensionless group for fluidized beds representing gas velocity. The second effect is due to formation of gas voids in different forms such as channels, bubbles, and slugs by the excess gas \( (u - u_{mf}) \). Hence gas holdup or bed voidage, a critical hydrodynamic parameter, is used as the convenient dimensionless expression of the inefficient contact component of the gas velocity effect on \( \eta \). At first, bed voidage was correlated with the velocity using a form similar to that first suggested by Richardson and Zaki [17] for liquid fluidized beds, and used by Davies and Richardson [18] for gas-solid systems:

\[
\frac{\dot{u}}{u_0} = a(\varepsilon)^n \tag{7}
\]

The STATISTICA software is applied to evaluate the empirical constants, \( a \), and \( n \), and were found to be 0.4607 and 2.3827 respectively with correlation coefficient \( (R) = 0.973 \).

The second step a correlation form is suggested for the dependence of removal efficiency on the operating variables as mentioned earlier and as follows:

\[
\eta_m = 1 - Y = 1 - \frac{C}{C_0} = A \left( \exp \left( \frac{B}{T} \right) \right)(\varepsilon)^{N_1}(\text{Re}_p)^{N_2} \tag{8}
\]

Where \( A, B, N_1, \) and \( N_2 \) are constants. The constants were calculated using STATISTICA software. The final empirical formula is:

\[
\eta_m = 1 - Y = 1 - \frac{C}{C_0} = 179.7090 \left( \exp \left( -\frac{3.3804}{T} \right) \right)(\varepsilon)^{-0.0691}(\text{Re}_p)^{-0.1864} \tag{9}
\]

with correlation coefficient \( (R) = 0.989 \). Figure (12) gives the experimental \( \eta_m \) against that predicted by the correlation Eq. (9).

4. Kinetic Analysis

Kinetic analysis of the experimental results was conducted to obtain the kinetic parameters of the catalytic oxidation of \( \text{SO}_2 \), since oxidation reaction is the rate limiting step for the range of flue gas flow rates at or above the optimum flow rate of 7.5 \( \ell/min \) as shown in the region of decaying efficiency in Figure (7). The suggested rate expression assumes first order reaction with respect to \( \text{SO}_2 \) as reported by Otake et al. [14], and facilitating the findings of Lizzio and DeBarr [13] that the reaction order = 1 up to approximately 500 ppm inlet concentration of \( \text{SO}_2 \), which is the concentration used in the experiments of varying \( W/F_{\text{SO}_2,o} \) values in this work. The rate expression given in Eq. (10) is based on the rate expression of Lizzio and DeBarr [13] (Eq. (2)) with simplification made by dropping the term of active sites decaying, since \( \eta \) decreased a little during the run (1 hour), thus:

\[
-r_{\text{SO}_2} = k C_{\text{SO}_2} \tag{10}
\]

Differential method of kinetic analysis stated by Levenspiel [19] was applied.

At first the conversion (which is taken as the average removal efficiency \( \eta_m \)) Vs. \( W/F_{\text{SO}_2,o} \) curves are plotted as shown in Figure (13) for \( C_o = 500 \text{ ppm} = 0.0203 \text{ mol/m}^3 \). Then the slopes \( \left( \frac{dW}{dx} \right) \) at selected points of \( W/F_{\text{SO}_2,o} \) on the conversion vs. \( W/F_{\text{SO}_2,o} \) curve which equal to the reaction rate \(-r_{\text{SO}_2}\),
were plotted against $C_{SO_2}$. The linear relation between $-r_{SO_2}$ and $C_{SO_2}$ (as shown in Figure (14) for $T = 80 \, ^\circ C$ as an example) gives the rate equation:

$$-r_{SO_2} = \left( \frac{0.882 \, m^3}{Kg \, \text{cat.\,min}} \right) \left( \frac{C_{SO_2} \, \text{mol}}{m^3} \right)$$

(11)

The results of the rate constant at the three tested temperatures are given in Table (2). The apparent activation energy for the oxidative desulfurization reaction is calculated by using the Arrhenius equation

The plot of $(\ln k)$ vs. $(1/T)$ shown in Figure (15) gives a straight line with slope $=-E/R$ and intersection $=\ln k_0$, so that $E=2.981 \, \text{kJ/mol}$ & $k_0 = 2.2 \, m^3/Kg \, \text{cat.\,min}$

Thus the reaction rate constant will be:

$$k = 2.2 \cdot \exp \left( \frac{-2981}{8.314 \cdot T} \right), \quad T \text{ in } K$$

(12)

And the oxidation reaction rate equation is

$$-r_{SO_2} = 2.2 \cdot \exp \left( \frac{-2981}{8.314 \cdot T} \right) C_{SO_2}, \quad T \text{ in } K$$

(13)

5. Conclusions

For both fixed and fluidized bed reactors an increase in the SO$_2$ removal efficiency was observed with an increase in the reactor temperature and a decrease of the SO$_2$ initial concentration in the flue gas stream. Curves of $\eta_{av}$ Vs. T showed an exponential like behavior for both reactors. The bed voidage in fluidized bed reactor was correlated with $u/u_t$ with $(R) = 0.973$. The effect of fluidized bed hydrodynamics on the removal efficiency was correlated with $(R) = 0.989$. A rate expression was suggested for the oxidation reaction. This rate equation was found to be accurately describes the reaction kinetics and the activation energy of the reaction was found to be 2.981 kJ/mole.

References


**Nomenclature**

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<thead>
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<th>Symbol</th>
<th>Description</th>
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<tr>
<td>a</td>
<td>Constant in Eq.(8) [-]</td>
</tr>
<tr>
<td>A</td>
<td>Constant in Eq.(9) [-]</td>
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<tr>
<td>B</td>
<td>Constant in Eq.(9) [-]</td>
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<tr>
<td>C₀</td>
<td>Inlet concentration of Sulfur Dioxide ppm</td>
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<tr>
<td>C_SO₂</td>
<td>Concentration of Sulfur Dioxide in the reactor mol/m³ or mg/m³</td>
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<td>C</td>
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<td>E</td>
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<td>F_SO₂</td>
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<td>N₁</td>
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### Table of Symbols and Units

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<td>$r_{SO_2}$</td>
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<td>t</td>
<td>Time min</td>
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<tr>
<td>u</td>
<td>Superficial velocity of gas</td>
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<td>$u_{mf}$</td>
<td>Minimum fluidization velocity</td>
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<td>$u_t$</td>
<td>Terminal (maximum) velocity</td>
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<td>W</td>
<td>Weight of activated carbon in the bed kg</td>
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<td>x</td>
<td>Sulfur dioxide conversion to sulfur trioxide</td>
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<tr>
<td>Y</td>
<td>Dimensionless SO$_2$ concentration = $\frac{C}{C_0}$</td>
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<td>Voidage of the expanded bed</td>
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<td>$\eta$</td>
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<td>$\eta_{av}$</td>
<td>Average removal efficiency of SO$_2$</td>
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<td>AC</td>
<td>Activated Carbon</td>
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### Constants and Units

- $r_{SO_2}$: mol/kg cat.min
- $\rho_g$: kg/m$^3$
- $dp$: m
- $u$: m/s
- $\mu_g$: kg/m$^3$.s
- $C$: mol/m$^3$
- $C_0$: mol/m$^3$
- $\varepsilon$: -
- $\eta$: -
- $\eta_{av}$: -
Table (1) Physical Properties of Activated Carbon

<table>
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<th>Property</th>
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<tr>
<td>Bulk density, (kg/m$^3$)</td>
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<td>Surface area, (m$^2$/g)</td>
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<td>Bed voidage, $\varepsilon_m$, (-)</td>
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<td>Particle density, (kg/m$^3$)</td>
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Table (2) The Values of Rate Constant for the 1st Order Reaction
Rate at Different Temperatures.

<table>
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<th>$k$ (m$^3$/min. Kg cat)</th>
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<td>0.76</td>
</tr>
<tr>
<td>323</td>
<td>0.8</td>
</tr>
<tr>
<td>353</td>
<td>0.882</td>
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Figure (1) Schematic Diagram of the Experimental Setup.
Figure (2): Effect of flue gas flow rate on the average removal efficiency of SO$_2$ at $C_o=500$ and different temperatures in fluidized bed for one hour duration.

Figure (3) the effect of temperature on the average removal efficiency of sulfur dioxide at $Q=15$ L/min and different inlet SO$_2$ concentrations in fluidized bed.
Figure (4) The effect of initial SO$_2$ concentration on SO$_2$ removal efficiency (T= 30 °C) in fluidized bed.

Figure (5) The effect of initial SO$_2$ concentration on SO$_2$ removal efficiency (T= 50 °C) in fluidized bed.
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Figure (6) The effect of initial SO$_2$ concentration on SO$_2$ removal efficiency ($T=80^\circ C$) in fluidized bed.

Figure (7) The effect of flow gas flow rate on the average removal efficiency of SO$_2$ at different temperatures in fixed and fluidized bed for one hour duration.
Figure (8) The effect of temperature on the average removal efficiency of sulfur dioxide at $Q=7.5 \text{ l/min}$ and different inlet $\text{SO}_2$ concentrations in fixed bed.

Figure (9) The effect of $\text{SO}_2$ initial concentration on removal efficiency ($T=30^\circ \text{C}$).
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Figure (10) The effect of SO$_2$ initial concentration on removal efficiency (T= 50 $^\circ$C).

Figure (11) The effect of SO$_2$ initial concentration on removal efficiency (T= 80 $^\circ$C).
Figure (12) The observed Vs. predicted values of average removal efficiency. \( C_0 = 500 \text{ ppm}, T = (30-80) \, ^\circ\text{C}, \text{ and } Q = (13.5-30) \, \ell/\text{min} \) for one hour duration.

Figure (13) Conversion of \( \text{SO}_2 \) as a function of \( W/F_{\text{SO}_2,0} \) at different reaction temperature.
Removal of SO$_2$ in Dry Fluidized and Fixed Bed Reactors using Granular Activated Carbon

Figure (14) Relation between $-r_{SO_2}$ and $C_{SO_2}$ for T = 80°C

Figure (15) Plot of ln $k$ Vs. 1/T.