SIMULATION OF SULFUR DIOXIDE REMOVAL FROM A GAS STREAM IN A FLUIDIZED-BED REACTOR

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ABSTRACT

The object of this work is to model and simulate a fluidized bed desulfurization reactor by coupling a reactor kinetic model with a fluidized bed model based on two –phase theory(bubbling-bed model) proposed by **Kunni and Levenspiel(1968)**.This model is useful for analysis of a reaction involving gas and solid. It generates the conversion data with respect to reaction time of both reacting gas and solids in a continuous flow(for both gas and solids) in fluidized bed. The flue gas or stack gas from a combustor which normally contains sulfur dioxide mixed with excess air is used as the fluidizing gas with calcium oxide as the fluidized solids . Calcium oxide is quite capable of reacting with SO_2 to effect its removal from the gas phase according to the exothermic reaction :

$$CaO_{(s)} + SO_{2(g)} + \frac{1}{2} O_{2(g)} \longrightarrow CaSO_{4(s)} + Heat$$

The effects of the solid feed rate, SO_2 concentration in the flue gas, bed height, bed diameter, particle size, fluidizing velocity and operating temperature on the extent of conversion of both gas and solid were investigated, for a fixed feed rate of flue gas and SO_2concentration (50.000 cm^3/sec ,3-5% by volume)at temperature range of (950-1000°C)and pressure 1 *atm*.

الخلاصة

الهدف من الدراسة هو انموذج رياضي ومحاكاة ازالة غاز ثاني اوكسيد الكبريت في مفاعل الطبقة المميعة وذلك بدمج حركة التفاعل الكيمياوي وموديل مفاعل الطبقة المميعة اعتمادا على نظرية الطورين (طور الفقاعة وطور الاستحلاب). الغاز المستخدم هو غاز خارج من مدخنة او محرقة تحوي غاز ثاني اوكسيد الكبريت . الغاز المستخدم للتمييع يكون قابل للتفاعل مع المادة الصلبة المستخدمة(اوكسيد الكالسيوم) المتوفرة موقعيا و القادرة على ازالتة .

تم في البحث دراسة تاثير معدل جريان المادة الصلبة ,تركيز الغاز ,قطر الطبقة و ارتفاعها ,معدل حجم الجزيئات للمادة الصلبة ,سرعة الغاز ,وحرارة الغاز داخل الطبقة بعد تثبيت معدل جريان الغاز ومعدل التركيز (50000 سم3/ثانية,5-3% نسبة حجمية)

النتائج النهائية لعملية المحاكاة اوضحت التصميم الامثل للمفاعل للحصول على اعلى تركيز لكل من الغاز والصلب ضمن الظروف السابقة.

INTRODUCTION

Sulfur dioxide, $S0_2$ originating from many sources, is a pollutant can result from the burning of coal, oils, and gases; refining of petroleum; smelting of ores containing sulfur, manufacture of sulfuric acid(1,2),paper,and pulp mills. It is produced by volcanoes and in various industrial processes. During the manufacture of sulfuric acid by rather the old chamber process or the newer contact process, some sulfur dioxide is emitted to the atmosphere (3).

The amount of this pollutant depends upon the size, type of plant and the efficiency of conversion of sulfur dioxide to sulfur trioxide (2).

Almost every sulfur dioxide control system in operation or under construction for full-scale utility or industrial boilers involves systems that produce a "throwaway" or disposable waste form of sulfur rather than a recovered product for sale. Most of these throwaway systems used direct lime or limestone a few others involve sodium-based sulfur dioxide absorption(4).Oxidation of SO_2 usually in the presence of acatalyst such as (VO_2), form(H_2SO_4),and thus acid rain.

Although a wide range of toxic gases and fumes can be released in accidents in chemical plants or stores, there are several gases and vapors which are commonaly encountered in many pollution situation, including contaminated land waste disposal and fuel combustion. It is interesting to note that the most hazardous of these, when judged by the concentration causing toxicity, is sulfur dioxide(5). Sulfur dioxide has the following four adverse effects:

i. Toxicity to humans.
ii. Acidification of lakes and surface waters.
iii. Damage to trees and crops.
iv. Damage to buildings).

Fluidization and Fluidized-Bed Reactors

The fluidized bed is one of the best known contacting methods used in the processing industry, for instance in oil refinery plants. Among its chief advantages are that the particles are well mixed leading to low temperature gradients, they are suitable for both small and large scale operations and they allow continuous processing. The fluid used to fluidize the solid particles can be either liquid or gas. Gas-solid fluidization is considered. A one-parameter model, termed the bubbling-bed model, is described by Kunii and Levenspiel (1991). It is used to calculate the hydrodynamic parameters. The one parameter is the size of bubbles. This model endeavors to account for different bubble velocities and the different flow patterns of fluid and solid that result.Compared with the two-region model, the Kunii-levenspiel (KL) model introduces two additional regions. The model establishes expressions for the distribution of the fluidized bed and of the solid particles in the various regions. These, together with expressions for coefficients for the exchange of gas between pairs of regions, form the hydrodynamic + mass transfer basis for a reactor model.

Simulation Procedure

The following assumptions were made in the simulation :

- The reactor is under steady state operation.
- The lime particles are of uniform size and are completely mixed in the bed.
- 3.No elutriation of particles occurs.
- Following these assumptions,an iterative computational algorithm was set up.The calculation procedure is described below and to start the computation a set of input data is required .These are as follows:
    - Mean particles size(0.1,0.075,0.05,0.025 *cm*).
    - Bed diameter range(50,75,100,150,200,250,300 *cm*).
    - Bed height range(50,75,100,125 *cm*).
    - Temperature of flue gase entering reactor (750,850,950,1000 ^{o}C) accordingly the viscosity as a function of temperature is determined (6)**,** a nomograph to determine absolute viscosity of gas as a function of temperature). Table (1) gives the viscosity as a function of temperature.

Table (1) viscosity as a function of temperature

Temp. (^{o}C)	750	800	850	900	950	1000
Viscosity(gm/cm.sec) $\times 10^{-4}$	4.5	4.7	4.9	5	5.2	5.4

- T_o (referans temperature) =273 *K*, P=1 *atm*, Mwt.*CaO* =56 *gm/mol* ,Mwt. of inlet gas mixture(fluidizing gas) =30.75 *gm/mol* consists mainly of SO_2=5% and air 95%(20%excess) as reactants..
- Mass flow rate of *CaO* (280,560, 840,1120 *gm/s*ec).
- Density of solid *CaO*(7) =2.8 gm/cm^3.

=0.5 ε_{mf} =0.4, ε_m *

* g=980 gm*/sec*^{2},

- 9- Density of fluidizing gas = (M.wt. p)/(RT)

The sequence of steps of calculations are as follows:

- The minimum fluidization velocity, u_{mf} , is evaluated by Ergun equation as:

$$u^2_{mf} + \frac{150(1-\varepsilon_{mf})\mu_g}{1.75\,\rho_g\,d_p}u_{mf} - \frac{g(\rho_s-\rho_g)\varepsilon^3{}_m\,d_p}{1.75\rho_g} = 0 \tag{1}$$

- Determine Re_p and calculate u_t through equations:

$$\mathrm{Re}_p = \frac{d_p\rho_g u_{mf}}{\mu_g} \tag{2}$$

$$u_t = \frac{(\rho_s - \rho_g) g d_p{}^2}{18\mu_g} \quad \text{for} \quad \mathrm{Re}_p \ \langle \ 0.4 \qquad (2.a)$$

$$u_t = \left[\frac{3d_p\, g(\rho_s - \rho_g)}{\rho_g}\right]^{0.5} \quad \text{for } 0.4 < \ \mathrm{Re}_p < 500 \qquad (2.b)$$

$$u_t = \left[\frac{3.1g(\rho_s - \rho_g)d_p}{\rho_g}\right]^{\frac{1}{2}} \quad \text{for } 500 < \ \mathrm{Re}_p < 2\times 10^5 \qquad (2.c)$$

- Predication of u_o (i,e. $u_o = \mathit{factor} \times u_{mf}$),the factor is between (3-6).
- Calculate d_p by equation (3):

$$d_b = 0.00376\times(u_o - u_{mf})^2 \qquad (3)$$

- Substitute d_p in equation (4) to calculate u_{br}.

$$u_{br} = 0.711\ (g\, d_b)^{1/2} \qquad (4)$$

- Calculate u_b by equation (5):

$$u_b = u_o - u_{mf} + u_{br} \qquad (5)$$

- The bed fraction of bubble phase , δ , is determined by equation (6), $\gamma_b = 0.0055$, $\alpha = 0.3$, $D_O = 2.6\times 10^{-4}\, cm^2/sec$

$$\delta = (u_o - u_{mf}) \ / \ u_b \qquad (6)$$

- Calculate m_b by equation (7):

$$m_b = (\delta\gamma_b)/((1-\varepsilon_{mf})(1-\delta)) \qquad (7\ )$$

- Calculate γ_c by equation (8).

$$\gamma_c = (1-\varepsilon\,\mathrm{mf})\left(\frac{3u_{mf}/\varepsilon_{mf}}{u_{br} - u_{mf}/\varepsilon_{mf}} + \alpha\right) \qquad (8)$$

- Calculate m_c by equation (9).

$$m_c = (\delta\ \gamma_c)/((1-\varepsilon_{mf})\ (1\ldots-\delta)) \qquad (9)$$

- Calculate γ_e by equation (10).

$$\gamma_e = ((1-\varepsilon_{\ mf})\ (1-\delta))/\delta) + (\gamma_b + \gamma_c) \qquad (10)$$

- Calculate m_e by equation (11).

$$m_e = (\delta\ \gamma_e)/((1-\varepsilon_{mf})\ (1-\delta)) \qquad (11)$$

- Checking for $m_b + m_c + m_e = 1$
- Calculate the diffusivity by equation (12), effective diffusivity by equation (13), and D_s by equation (14).

$$D = D_O\left(\frac{T}{T_O}\right)^{1.83} \qquad (12)$$

$$D_e = D\varepsilon_{mf} \qquad (SO_2\ through\ flue\ gas) \qquad (13)$$

$$D_s = D0.15 \qquad (SO_2\ through\ solid\ ) \qquad (14\ )$$

- From equation (15) and (16) Calculate K_{bc}, K_{ce}.

$$K_{bc} = 4.5\left(\frac{u_{mf}}{d_b}\right) + 5.85\left(\frac{D^{\frac{1}{2}}g^{\frac{1}{4}}}{d_b^{\frac{5}{4}}}\right) \qquad (15)$$

$$K_{ce} = 6.78\left(\frac{\varepsilon_{mf}\ D_e\ u_b}{d_b^{\ 3}}\right)^{1/2} \qquad (16)$$

- Select values of Kr between l and 1000 sec^{-1} and for each value of Kr :-

a. Calculate Kf by equation (17) and substitute in to equation (18) to find XA.

$$Kf = (\gamma_b\ Kr\ ) + \frac{1}{\dfrac{1}{K_{bc}} + \dfrac{1}{(\gamma_c Kr) + \dfrac{1}{\dfrac{1}{k_{ce}} + \dfrac{1}{\gamma_e kr}}}} \qquad (17)$$

$$1 - XA = \frac{C_{Ao}}{C_{Ai}} = e^{-kf} \qquad (18)$$

b. Calculate τ by equation (19).

$$\tau = \frac{\rho_B R}{bk_c C_A} = \frac{\rho_B d_p}{2bk_c C_A} \qquad (19)$$

and

$\rho_B =$ molar density of CaO particle(density of CaO /M.wt. of CaO) , and Kc fromAreniesequation(i.e $Kc = Ae^{\frac{-Ea}{RT}}$)
(19.a)

c. Calculate the over all rate constant(K_m)by equation (20).

$$\frac{1}{K_m} = \frac{1}{kc} + \frac{d_p}{12D_s} \qquad (20)$$

d. Determine L_f from eq.(21) and L_{mf} from the correlation between them (i.e. eqn 22).

$$L_f = -(\ln\ (1 - XA)\, u_b)/\, Kf \qquad (21)$$

The bed height can be related as follows:

$$\frac{L_f}{L_{mf}} = \frac{1-\varepsilon_m}{1-\varepsilon_{mf}} \tag{22}$$

e. Determine the mean residence time of particles in the bed ($\bar{t}$) by equation (23).

$$\bar{t} = \frac{W}{F_1} = \frac{A_t\, L_{mf}\left(1-\varepsilon_m\right)}{F_1} \qquad (\ 23)$$

Where

W=weight of bed(gm), F_1 = out flow rate of solid(gm/sec)

A_t is the cross-sectional area of bed needed and obtained by mass balance.

$A_t = ((F_o*CAI)/mwts)/(((u_o*(T_o./(T_o+T)))*(((0.21-0.1)/22400)+((CAI-(CAI*0.01))/22400)))$

f. Substitute t and $\bar{\tau}$ in equation (24) to find $\overline{XB}$.

$$\overline{XB} = 1-3(\frac{\bar{t}}{\tau})+6(\frac{\bar{t}}{\tau})^2-6(\frac{\bar{t}}{\tau})^3 \left[1-\exp(-\tau/\bar{t})\right] \qquad (24\ )$$

g. Calculate ϕ and $\overline{CA}$ by equation (25) and (26)respectively.

$$\phi = \left[\frac{1}{1+y_e\,\frac{kr}{k_{ce}}}\right]\left[\frac{1}{y_c+\frac{1}{\frac{kr}{k_{ce}}+\frac{1}{y_e}}}\right] \qquad (\ 25)$$

$$\overline{C_A} = \frac{(1+m_c\gamma_e\,\frac{kr}{kce})C_{Ai}X_i\phi}{(L_{mf}kr/u_b)} \qquad (\ 26)$$

h. Calculate XA by equation (27).

$$\frac{1}{b}\left(\frac{F_0}{M_b}\right)\overline{XB} = A_t\, u_o\, C_{Ai}\ XA \tag{27}$$

i . Compare XA calculated from equations (18) with XA calculated from equation (27).If they do not match, assign new trial value for Kr and repeat steps a through i.

k . The calculation is terminated when the error criterion $\left|XA(18)-XA(27)\right|<10^{-3}$ is satisfied.

Results and discussion

The effects of the fluidizing velocity,solid feed rate,mean particle size of the solids,operating temperature,SO_2 concentration of feed gas and bed geometry on the conversion efficiency of both gas and solids are discussed separately. Fig.s(1)-(19) show these effects on plots as % conversion versus reaction time.It should be noted that in this study the feed gas flow rate is set within the range of 49580-51600 cm^3/sec (average 50000 cm^3/sec) as a design parameter.

Effect of Operating Temperature

The effect of temperature on the conversion efficiency of the gas and solids are presented in Fig.s (1) and (2) respectively. The temperature range covered in this study is 750 ^{o}C-1000^{o}C.At relatively lower temperatures (<500^{o}C) calcium sulfite($CaSO_4$) an unstable product of the reaction can be formed and may react reversibly to give back SO_2.It is therefore important that temperatures higher than 650^{o}c should be used to avoid the formation of $CaSO3$. For this purpose, temperatures above 650 ^{o}C were used in this investigation(8). As it is seen maximum conversions are obtained within the temperature range(950 ^{o}C-1000 ^{o}C) , shorter reaction times are obtained as the temperature is increased because the rate of reaction increases with temperature as well as the solute mass diffusivity (9) .From these plots led to the conclusion that the desirable range of temperature is 950 ^{o}C-1000 ^{o}C ,preferably 950 ^{o}C .Consequently, the operation temperature chosen is 950^{o}C and used throughout the calculations of the simulation program.

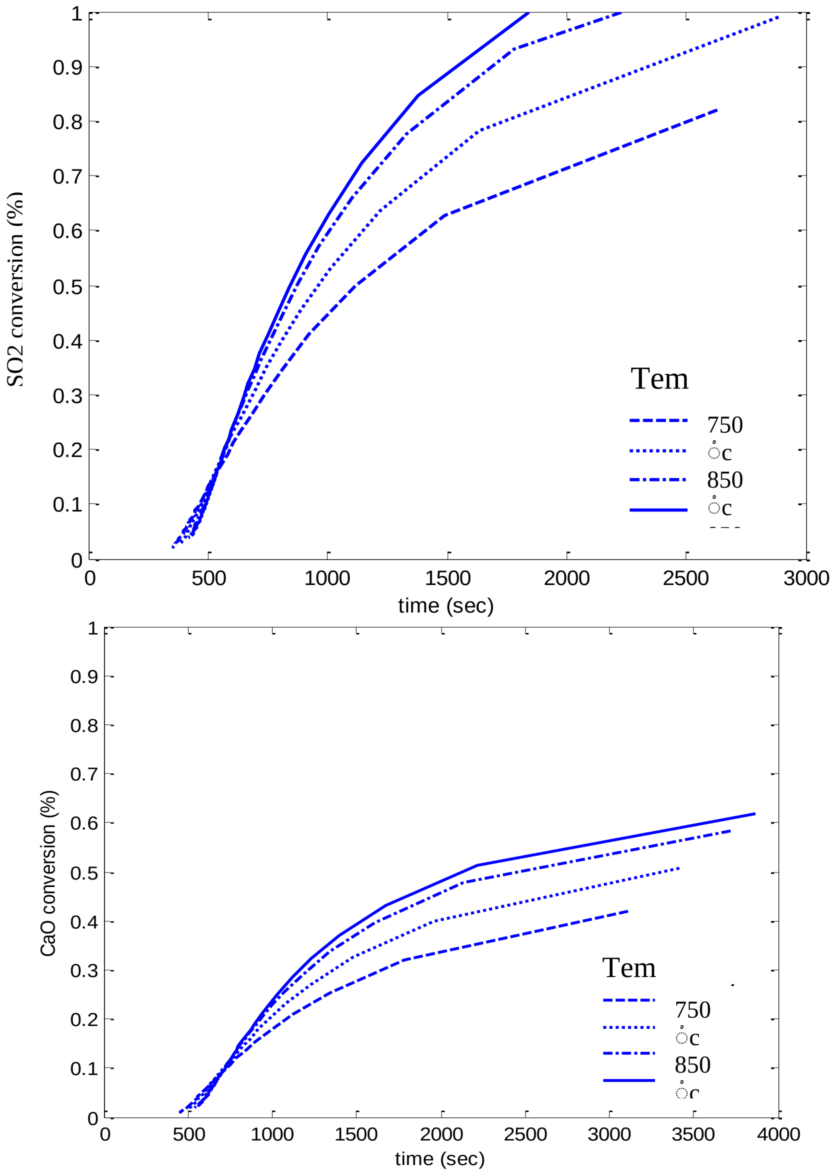

Fig.2.- Fo=280 *g/s*, dp=0.1*cm*, uo/ umf =3

- EFFECT OF MEAN PARTICLES SIZE

The effect of mean particle size on the conversion efficiency is shown in Figs (3)and(4) for the gas and solids. These data are obtained for a fixed feed of solids F_o =280 gm/sec and fluidizing velocity uo=3umf cm/sec. For the range of particle size used in this study(0.1-0.025*cm*) and as it is seen from the plots,a decrease in particle size is to increase the conversion efficiency of both gas and solids for all temperatures used.For instance ,at 750 ^oC the conversion of the gas increased from 62-99% over a particle size decrease from 0.1*cm* down to 0.025*cm*. On the other hand,this increase in efficiencies is accompanied by rapid decrease in the reaction time over the residence time of the particles in the bed. But this residence time showed little change with decreasing particle size and increasing temperature. The heighest conversion efficiencies were obtained with the small size particles(i.e =0.025 *cm*).This is expected,because small size particles offer larger surface area for reaction and solute diffusion than particles of larger size.However,very small size particles are not suitable for fluidized bed reactors due to carry over problems and therfore are not considered in this study .

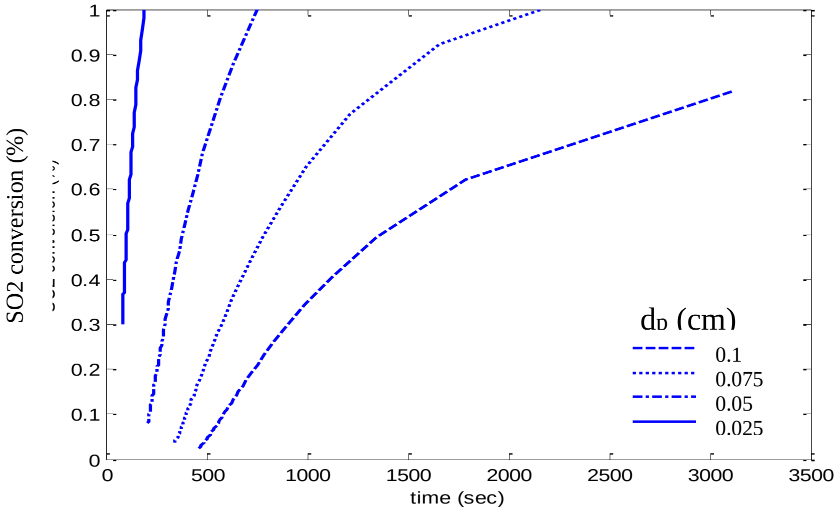

Fig.3.- Fo=280 *g/s*, T= 750 ^{o}c, uo/ umf =3

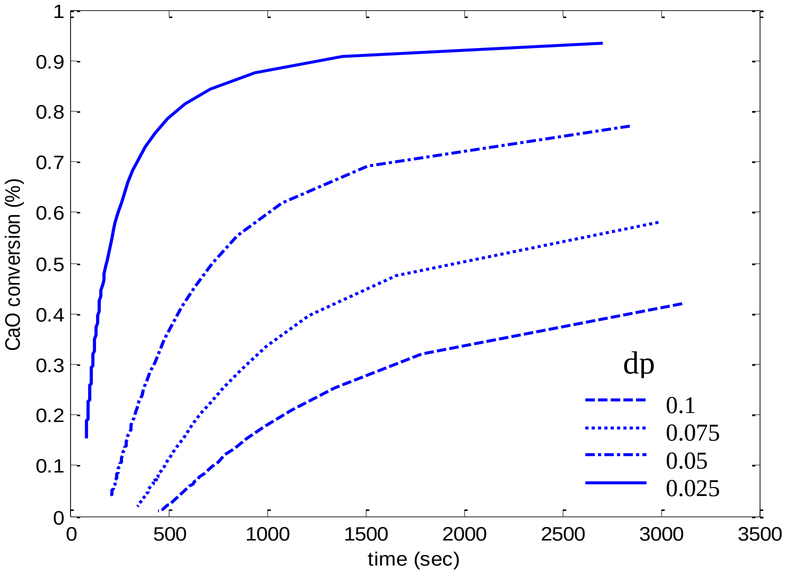

Fig.4.- Fo=280 *g/s*, T= 750 °c, uo/ umf =3

EFFECT OF GAS FLUIDIZING VELOCITY

As stated in Kunni and Levenspiel(1968) recommend a fluidizing velocity in fluidized bed reactors to be greater than $2u_{mf}$ (i,e .u_o >=$2u_{mf}$) for proper reaction and stable operation. However, higher values of uo are not recommended in order to avoid solids carry over.Solids carry over can be prevented if uo is chosen to be well below the terminal velocity of the solid particles in the bed. Accordingly, the range of uo values coverd in this study is taken to be (3-6)u_{mf} cm/sec. Figs.(5) and (6) depict the effect of changing uo on the conversion efficiency of the gas and solid with the operation temperature , solid feed rate and size of particles are kept constant at 950 ^{o}C , 280 *gm/s*ec and 0.025*cm* respectively. An increase in uo leads to a decrease in %conversion,for solid only (gas conversion remains unchanged). However, both gas and solids almost attain the same conversion(95-96% for the solid and 99% for the gas) within the same range of reaction time of 360-475 *sec*.

This range of reaction time is well below the residence time of the solid particles in the reactor. For proper operation of the reactor, it is essential that the reaction should go to completion before any of the solid particles leave the reactor. In other words, it is essential that the reaction time should , always, be less than the residence time of the solid particles in the bed in order to achieve complete reaction.

From these results, it appears that a fluidizing velocity of 3umf is our choice as an optimum value.The use of fluidizing velocity higher than 3umf does not affect the conversion efficiency of the gas but more reaction time is needed. This would exceed the residence time of the solids in the bed. On the other hand an increase of uo above 3umf entails a decrease in solid conversion. Thus higher values of uo are not desirable. Consequently, the fluidizing velocity is fixed at 3umf and used throughout the calculations of the simulation program.

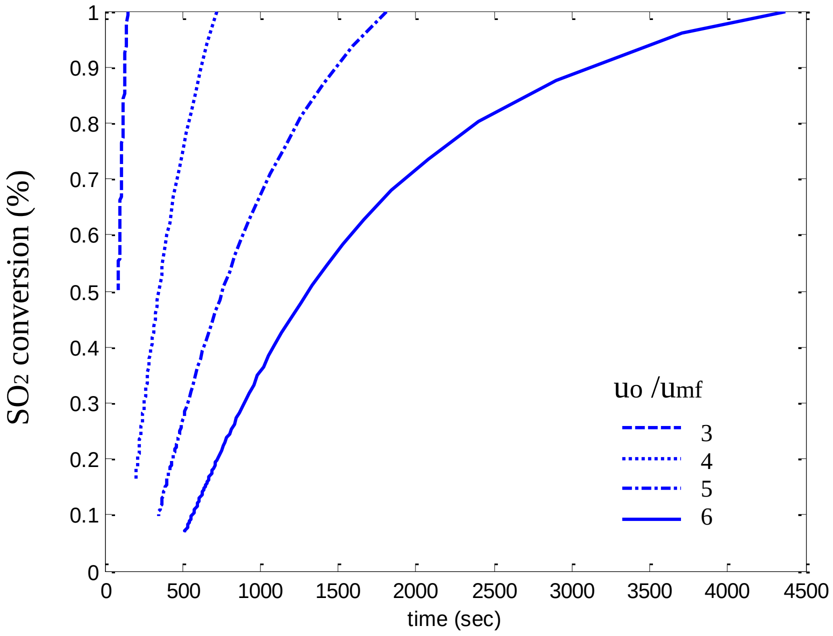

Fig.5.- dp=0.025*cm*, T= 950 °c, Fo=280 *g/s*

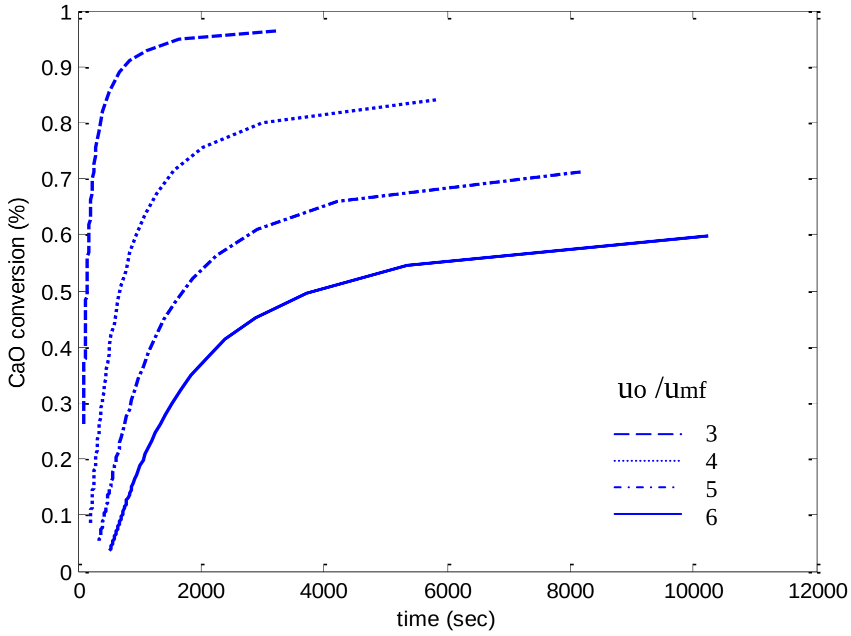

Fig.6.- dp=0.025 *cm*, T= 950 °c, Fo=280 *g/s*

EFFECT OF INLET *SO*2 CONCENTRATION

The concentration of SO_2 in the feed gas entering the reactor is fixed within the range 3-5% by volume.This concentration originates in the heavy fuel oil used in the combustion. Davidson and Harrison(1968) state that commercial fuel oils used for combustion usually contain sulfur that is finally converted to SO_2 within the range 3-5%v/v in the flue gas.

Figs (7-12) show the effect of varying SO_2 concentration on the conversion efficiency of the gas and solid respectively. As it is shown, with the other parameters fixed, the reaction time increases as SO_2 concenteration is decreased. It should be noted that the inlet concenteration of SO_2 is usually fixed as the concenteration of the feed gas to be desulfurized. However, some variations in the quality of fuel used in the combuster could change this composition, but still maximum conversions are obtained. For the simulation programe, the inlet SO_2 concenteration is set at 5%v/v in a feed gas of 50000 cm^3/sec. It can be noted from the results the conversion of the gas and solid remaine constant regardless of SO_2 conversion. However,if the concentration of gas is increased ,the material balance requires a decrases of the gas flow rate and hence a decrase of reaction time and residence time.

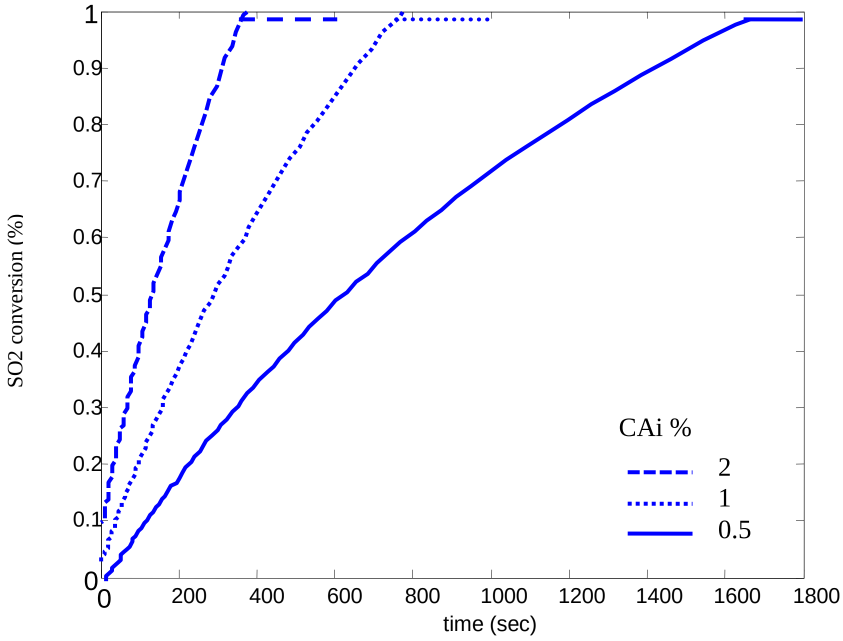

Fig.7.- dp=0.025 *cm*, T= 950 °c, uo/ umf =3, Fo=280 *g/s*

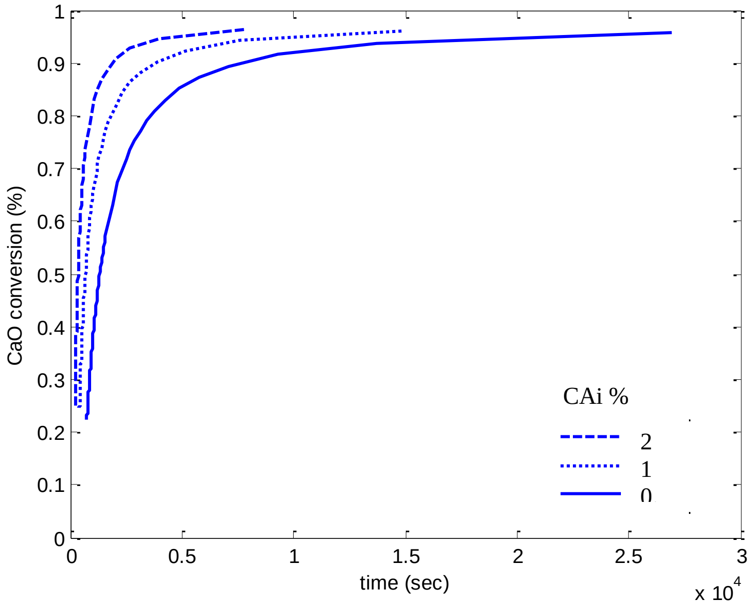

Fig.8.- dp=0.025 *cm*, T= 950 ̊c, uo/ umf =3, Fo=280 *g/s*

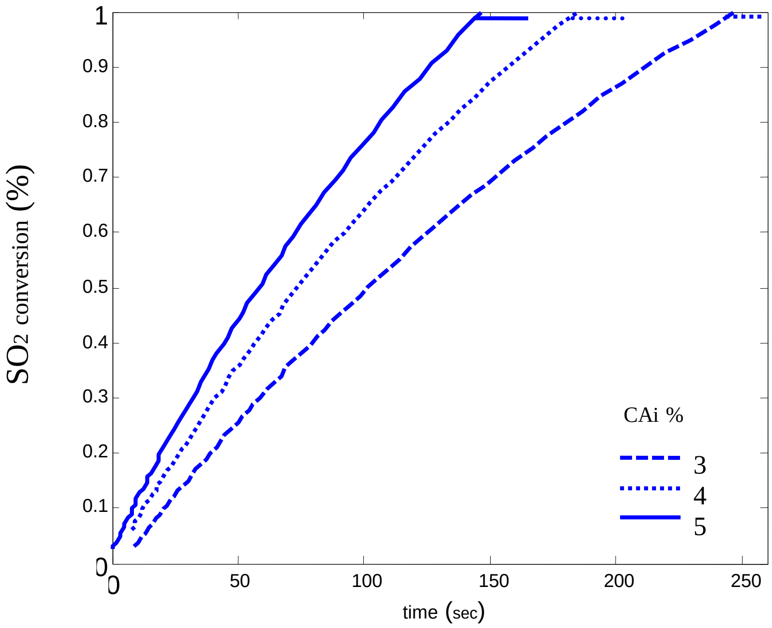

Fig.9.- dp=0.025 *cm*, T= 950 ̊c, uo/ umf =3, Fo=280 *g/s*

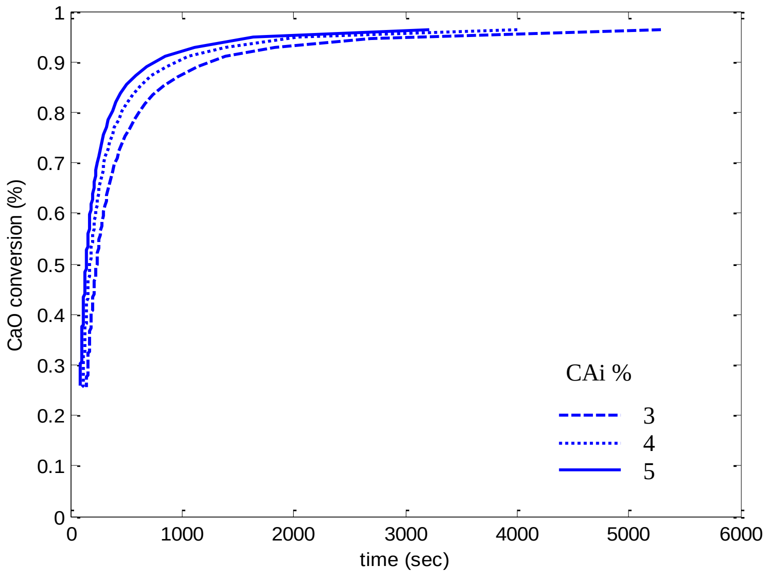

Fig.10.- dp=0.025 *cm* T= 950c, uo/ umf =3, Fo=280 *g/s*

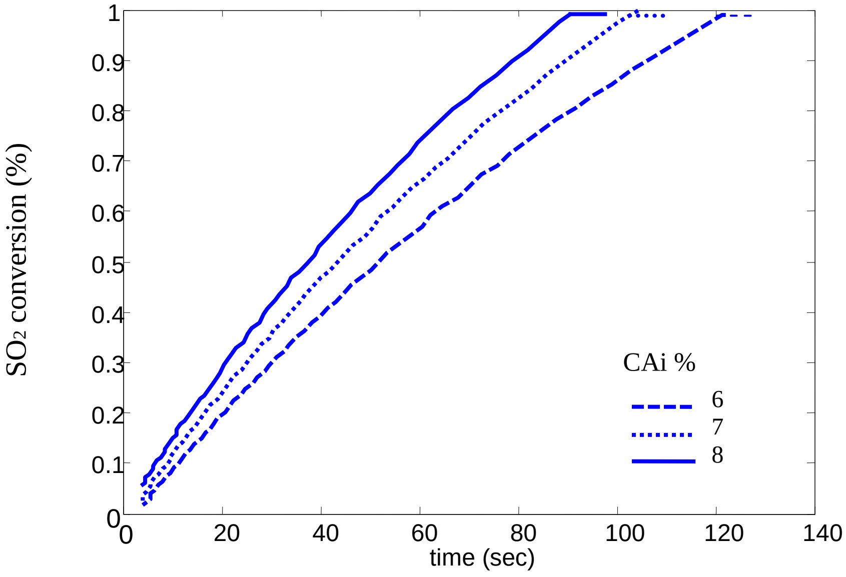

Fig.11.- dp=0.025 *cm*, T= 950 ̊c, uo/ umf =3, Fo=280 *g/s*

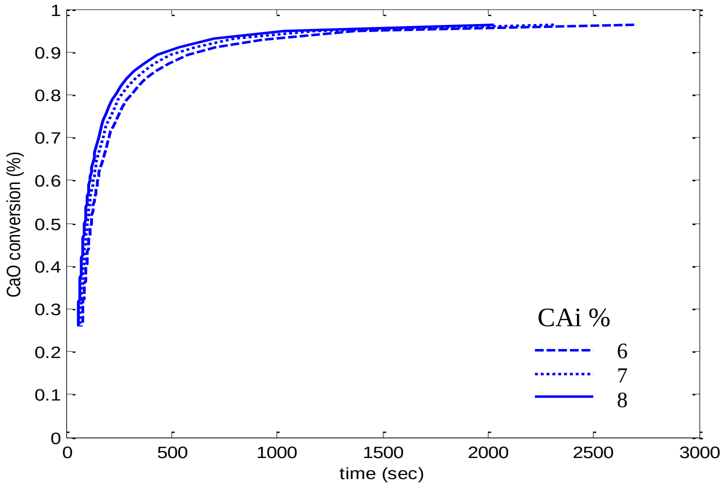

Fig.12.- T=950^{O}C, dp=0.025 *cm*, uo/ umf =3, Fo=280 *g/s*

EFFECT OF SOLID FEED RATE

The effect of solid feed rate has been investigated for a reaction temperature of 950 ^{o}C and mean particle size=0.025cm with the other parameters are fixed at their desired values.

Increasing the rate of solids entering the reactor has no effect on the conversion efficiency of both gas and solid and the reaction time. As can be seen from Figs(13)and(14), the conversion of the gas and solid rise rapidly to their maximum values within a reaction time of 360-475 *s*ec and this reaction time appears to be well below the residance time of solids in the reactor. However, an increase in solid flow rate involves marked increase in the volumetric flow rate of the input gas to the reactor. Since the range of the input flow of gas is already fixed at 50000 *cm*^{3}*/s*ec an increase in solid flow rate above 280*gm/sec* will require higher gas flow rate. Therefore, the recommended solid feed rate is 280*gm /sec*.

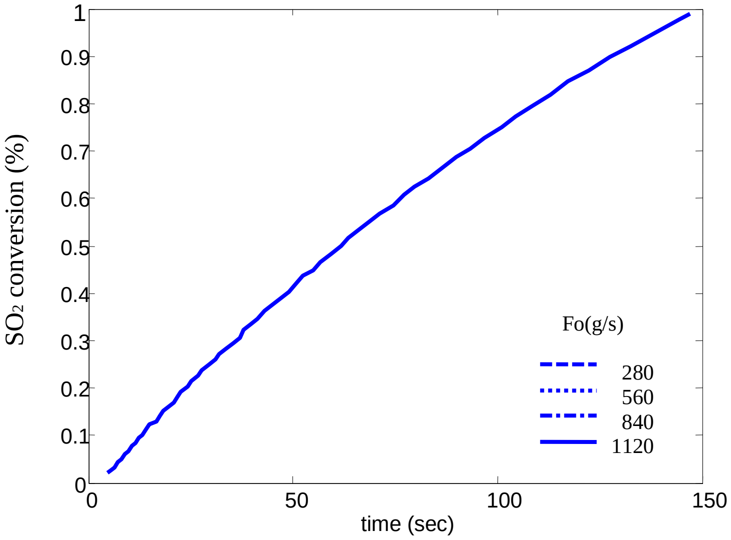

Fig.13.- dp=0.025 *cm*, T= 950 °c, uo/ umf =3

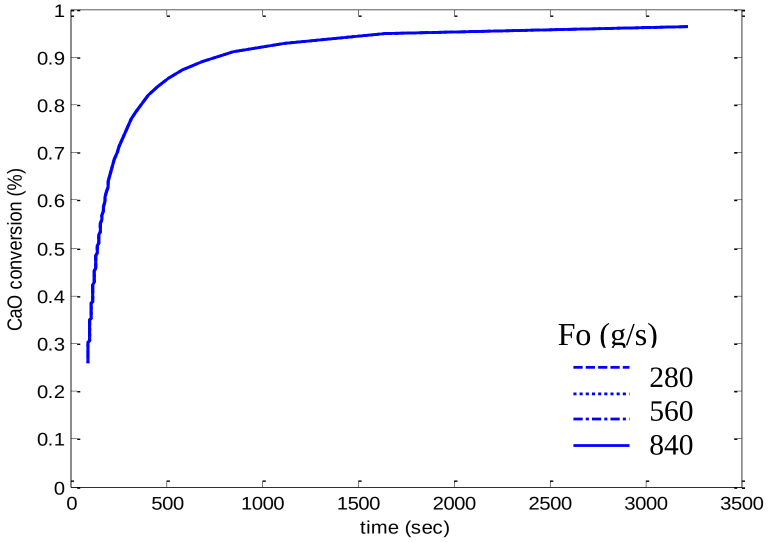

Fig.14.- dp=0.025 *cm*, T= 950 °c, uo/ umf =3

Effect of Bed Height

Fig(15) depicts the effect of increasing bed height on the reaction time to achieve maximum conversion for the gas and solid at a given bed diameter of 60*cm*. As it is shown the reaction time increases with bed height as well as the residence time. This effect is indirectly linked with equation (26),(19),(23) for the gas concenteration inside the reactor, reaction time and residence time respectively.The results show that the optimum reaction time of 369-475 *sec* corresponds to a fixed bed height of 75-150*cm*. Using higher beds will lead to higher reaction times.

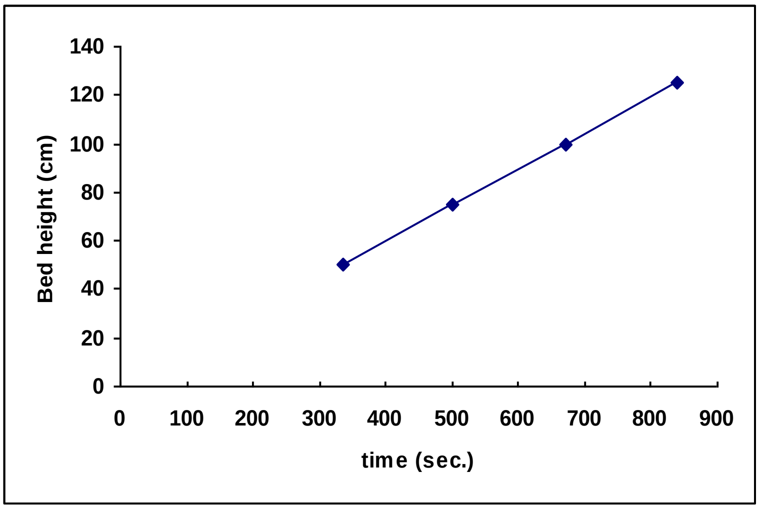

Fig. 15. Fo= 280 *g/s*., D_B= 50-75 *cm*, T= 950 ^oC, u_o/u_{mf}=3,d_p=0.025*cm*

EFFECT OF BED DIAMETER

The effect of bed diameter on the conversion efficiency of the gas and solids has been investigated for the range of bed diameters of 50-75*cm* at constant temperature of 950°C,particle size of 0.025*cm*, fluidizing velocity of 3umf cm/sec and solid feed rate of 280*gm/sec* .The inlet concentration of SO_2 is taken to be the heighest allowable value of 5% by volume.The results are plot in Figs(16)-(19).The plots show that solid conversion is ranging between 92-97% for bed diameters between 50-75*cm*.This conversion increases until it reaches 99% for beds between 150-300*cm* in diameter.On the other hand the gas conversion exhibits a rapid decrease with increasing bed diameter(i,e. 8% for beds close to 300*cm* in diameter).It is also noted that inorder to achieve a fluidizing velocity of 3umf cm/sec. The feed flow rate of the gas should increase,and as can be seen this flow rate becomes exceedingly high at bed diameters close to 300*cm*. Such high flow rates do not match the design feed rate of 50000cm^3/sec .The reaction time remains unchanged at 360-475 sec,but the residence time increases with bed diameter. From these results it is clear that the increase in bed diameter adversely affects the efficiency of gas conversion, but it has little effect on solid conversion.It appears that bed diameters between 50-75*cm* would be the best choice for operation.

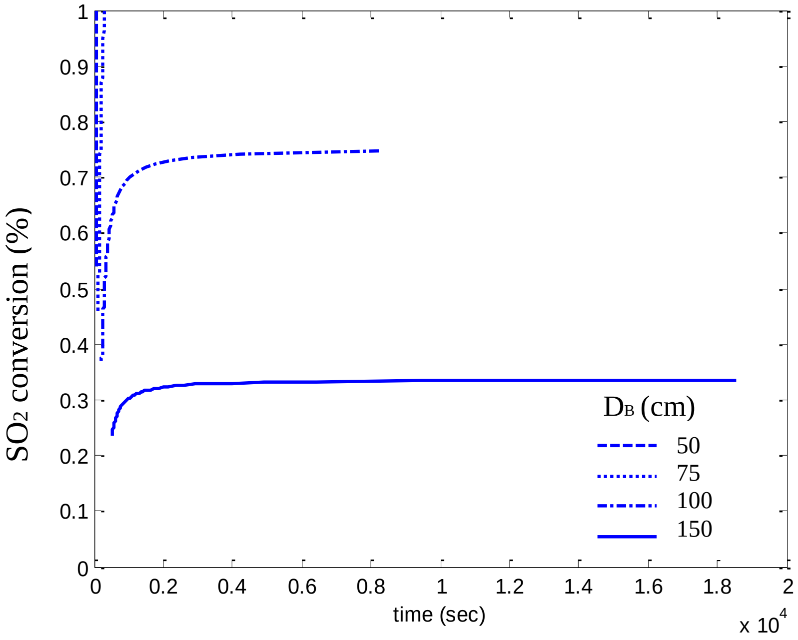

Fig.16. Fo=280 *g/s*, T= 950 °c, uo/ umf =3, dp=0.025

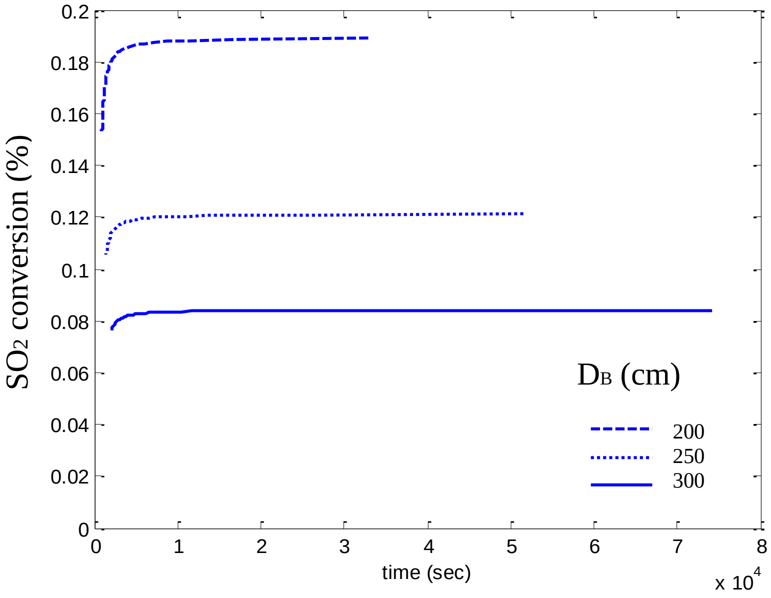

Fig .17. d_p=0.025 *cm*, T= 950 $^{\circ}$c, u_o/u_{mf} =3, F_o=280 *g/s*

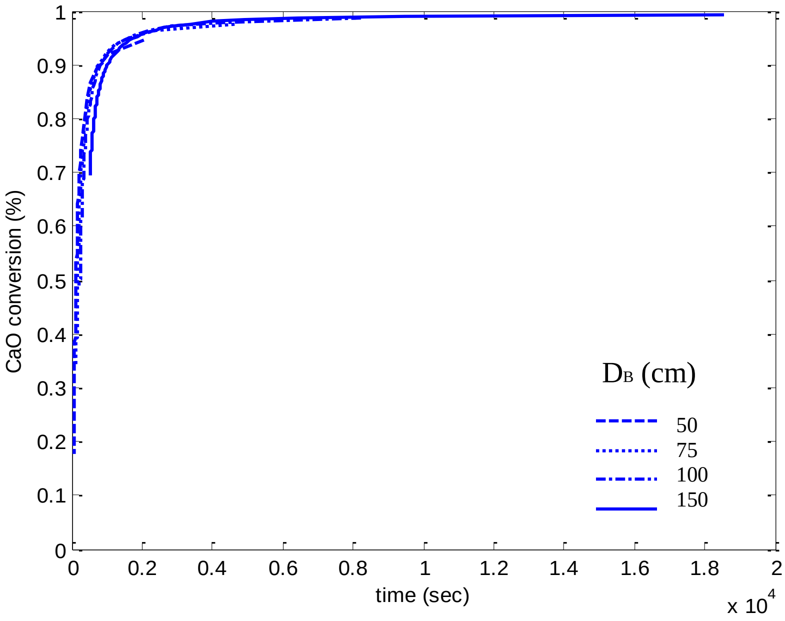

Fig.18. F_o=280 *g/s*, T= 950 $^{\circ}$c, u_o/u_{mf} =3, d_p=0.025 *cm*

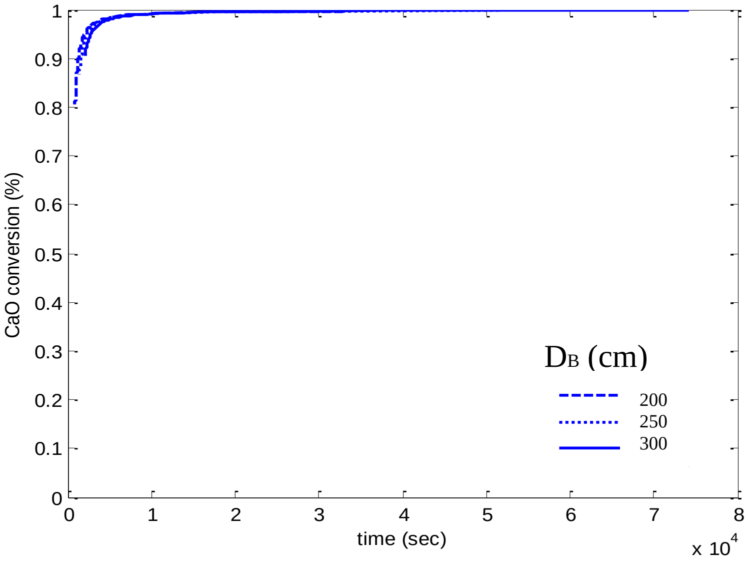

Fig.19.dp=0.025 *cm*, T=950 c°, uo/ umf =3.,Fo=280 *g/s*

CONCLUSIONS

Recently the bubbling-bed model was applied to the prediction of gas conversion in solid catalysed gas-phase reaction. In the present work this model is used to determine the extent of chemical conversion of both reacting gas and solids in a continuoas flow(for both gas and solid) fluidized bed. This procedure is applied for removal of SO_2 from flue gas using CaO as solid reactant in a fluidized bed reactor based on a feed rate of 50000*cm*^{3}*/sec* of flue gas.The results of the simulation programe for maximum conversion of gas and solid are summarized as follows:

-The feed rate of CaO is to be 280gm/sec(about 24.192*Kg/day*) producing 3.617 gm/sec (about 27 *Kg/day*) of $CaSO_4$(99% conversion) as by product,a useful building material.

-The reactor will operate a diabatically at temperature within 950-1000 °C.

-The mean particle size of *CaO* is 0.025*cm*.

- The fluidizing velocity of the gas(flue gas) should be 3 times the minimum fluidizing velocity of the solid particles.
- The flow rate of the solid feed depends on the flow rate of the feed gas.
- To a chieve the same conversion of both gas and solid an increase in inlet *SO2* concentration requires a decrease in gas flow rate.
- The best ratio of bed height to bed diameter for maximum conversion should not exceed 2.

NOMENCLATURE

A_t = cross-sectional area of bed (cm^2)

A = pre-exponential factor for solid diffusion of SO_2 through $CaSO_4$ (cm^2/s).

b = stoichiometeric coefficient in gas/solid reaction equation.

C_{Ai} = concentration of gas in the inlet gase stream $(gmol/cm^3)$

C_{Ao} = concentration of gas (or bubble gas)leaving the bed $(gmol/cm^3)$

$\left.\begin{matrix} C_{Ab} \\ C_{Ac} \\ C_{Ae} \end{matrix}\right\} =$ concentration of gas A in bubble,cloud-wake region,and emulsion respectivlely $(gmol/cm^3)$

$\overline{C}_A =$ average concentration of reactant A encountered by solid particles in the bed $(gmol/cm^3)$

$\left.\begin{matrix} D \\ D_e \\ D_s \end{matrix}\right\} =$ diffusion coefficient, effective diffusion coefficient in emulsion, and diffusion coefficient through layer of solid product or ash $(cm^2/\sec)$

D_O = diffusivity at standard conditions $(T = 273K, P = 1atm)(cm^2/\sec)$.

D_B = bed diameter (cm).

$d_p =$ particle diameter (cm).

$d_b =$ bubble diameter (cm).

E_a = activation energy for solid diffusion of CaO through $CaSO_4$ (J/mol).

$F_o =$ feed flow rate of solids $(gm/\sec)$.

$F_1 =$ out flow rate of solid $(gm/\sec)$.

$g =$ acceleration of gravity $(gm/\sec)$.

$\left.\begin{matrix} K_{bc} \\ K_{ce} \end{matrix}\right\} =$ interchange coefficients between bubble and cloud wake region, and between cloud – wake region and emulsion, respectively $(1/\sec)$.

$Kr =$ reaction rate constant defined in equation (17) $(1/\sec)$.

$K_c =$ rate constant for surface defined in equation (19.a) $(cm/\sec)$.

$k_f =$ reaction rate group defined in equation (17) dimensionless.

K_m = overall rate constant $(cm/\sec)$.

L= height of the reactor (cm).

$\left.\begin{matrix} L_f \\ L_{mf} \end{matrix}\right\} =$ height of fluidized bed and static bed respectively (cm).

$M_B =$ molecular weight of solid material (gm/mol).

$\left.\begin{matrix} m_b \\ m_c \\ m_e \end{matrix}\right\} =$ weight fraction of solids defined by equation (7,9,11).

N_A = gramme moles of A.

rc = radius of solid unreacted core (cm).

R= gas constant $(atm.cm^3/mol.K)$.

mean residence time of particles in the bed (sec) $\bar{t} =$

t = residence time for particle in the bed (sec).

$\left.\begin{matrix}u_b\\ u_{mf}\\ u_o\end{matrix}\right\} =$ velocity of rising bubble, minimum fluidizing velocity, and superficial gas velocity, respectively $(cm/\sec)$.

u_t = terminal or free-falling velocity of the particles $(cm/\sec)$.

$\left.\begin{matrix}V_b\\ V_w\end{matrix}\right\} =$ volume of bubble and of wake, respectively (cm^3)..

V_s = volume of a solid particle (cm^3).

(cm^3) V_T= total volume of the particles

$W =$ weight of bed (gm).

$X_A =$ fractional conversion of reactant gas (A).

$\overline{X}_B =$ mean fractional conversion of solid (B) material in the exit stream.

$\left.\begin{matrix}y_b\\ y_c\\ y_e\end{matrix}\right\} =$ volume fraction of solids by equations (8,10) , dimensionless.

Greek Letters

$\left.\begin{matrix}\varepsilon_m\\ \varepsilon_{mf}\end{matrix}\right\} =$ void fraction in a fixed bed and in a bed at minimum fluidization, respectively , dimensionless.

$\tau =$ time for complete conversion of a single particle (sec).

δ = bed fraction of bubble phase .

ρ_s = density of solid (gm/cm^3) .

ρ_g = density of gas (gm/cm^3)

μ_g = viscosity of gas $(gm/cm.\sec)$

ϕ = sphericity of a particle , dimensionless .

$\alpha =$ ratio of wake volume to bubble volume .

$\rho_B =$ molar density of solid material $(g\,mol/cm^3)$

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