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PREDICTION OF DROPLET DISPERSION AND PARTICLE REMOVAL EFFICIENCY OF A VENTURI SCRUBBER USING DISTRIBUTION FUNCTIONS*

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Abstract– A modified three-dimensional dispersion model of our previous work[5] was modified and used to investigate the effect of droplet size distribution in addition to droplet concentration distribution for the prediction of liquid droplet dispersion and particulate removal efficiency in a venturi type scrubber. For the sake of including droplet size distribution into the model properly, it was assumed that droplet size distribution obeys Rosin-Rammler distribution. The experimental data of Viswanathan *et. al.* [1] for liquid droplet dispersion and Brink and Contant [2] for particle removal efficiency were used to test the results of this new mathematical model. The results from the model show that by taking droplet size distribution into account the results of the model will be in better agreement with the experimental data.

Keywords– Venturi scrubber, droplet size distribution, droplet dispersion, Eulerian approach

1. INTRODUCTION

Venturi scrubber is a device frequently used for removing pollutant gas and particulate from a gas stream. In this device liquid is atomized into a high velocity gas stream. As a result of atomization, droplets with different diameters are formed and are then dispersed nonuniformly. Figure 1 shows the configuration of a venturi scrubber. Since droplets with different sizes have different eddy diffusivity and velocity, the size distribution of droplets varies throughout the scrubber. Furthermore, due to the nonuniform dispersion of droplets, a droplet concentration distribution (for each droplets group having the same size) exists in the scrubber.

Several attempts have been made to simulate liquid dispersion and particulate removal in a venturi scrubber. All of the models identified in the literature can be classified in one of the following groups:

1. Empirical models: these are models by which the performance of a scrubber are predicted using an empirical correlation.
2. Theoretical models: almost all of these models are based on a one, two or three-dimensional dispersion model by considering one or both of the following simplifications:
    a) Uniform droplet concentration distribution. It is assumed that transversal turbulency is high and droplets are dispersed uniformly across the scrubber.
    b) Constant mean droplet size. A mean droplet size calculated at the atomizing zone is used throughout the scrubber. This means that droplet size distribution remains constant throughout the scrubber.

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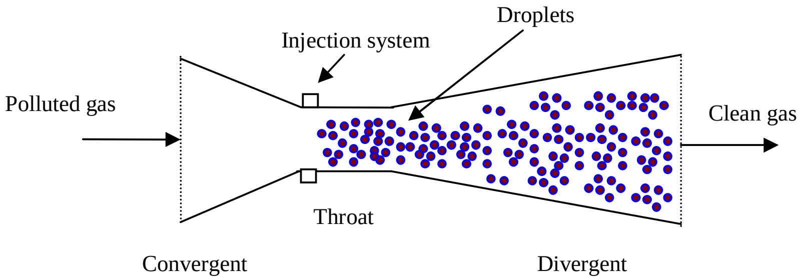

Fig. 1. Configuration of a venturi scrubber

Taheri and Shieh [3] have used a dispersion model based on mean droplet size. Placek and Peters [4] have considered the size distribution of droplets in the model, but with uniform droplet concentration through out the scrubber. Viswanathan *et. al* [1] have solved a two-dimensional dispersion equation by considering droplet size distribution and, they have used a four-point trapazoidal integration method in a simple way. Fathikalajahi *et. al*. [5] have solved a three-dimensional dispersion model based on mean droplet size by introducing a new method for the evaluation of droplet eddy diffusion. Fathikalajahi *et. al*. [6] have also studied the effect of the main operating parameters on removal efficiency by using a dispersion model based on mean droplet size. Viswanathan [7] include a term for transversal droplet velocity due to jet penetration. Ananthanarayanan and Viswanathan [8] extended their previous model to cylindrical geometries. Goncalves *et al*. [9] modified the dispersion model by modeling jet penetration through a gas stream.

In this study our previous model was modified in order to include the effect of droplet size distribution on a venturi scrubber performance. Also, by using the Gaussian quadrature method in a numerical solution, the length of the calculations and the required CPU time have been reduced significantly.

2. MATHEMATICAL MODEL

This model is based on a three-dimensional dispersion of droplets by convection and eddy diffusion. The steady state equation expressing material balance for droplets with variable diameters is as follows:

$$\frac{\partial[V_{d_x} C_d(D_d)]}{\partial x}+\frac{\partial[V_{d_y} C_d(D_d)]}{\partial y}+\frac{\partial[V_{d_z} C_d(D_d)]}{\partial z}=E_d\{\frac{\partial^2[C_d(D_d)]}{\partial y^2}+\frac{\partial^2[C_d(D_d)]}{\partial z^2}\}+SN_d(D_d) \qquad (1)$$

Where, $N_d(D_d)$ is the number frequency distribution of the drop size, ($N_d(D_d)dD_d$ is the number fraction of droplets having a size between D_d and D_d+dD_d) at the atomization zone, and $Cd(D_d)$ is the number concentration frequency distribution of the drop size (Cd(Dd)dDd is the number concentration of droplets having a diameter between Dd and Dd+dDd) at any section of the scrubber depending on x,y,z. This equation can be obtained by writing a differrential mass balance for droplets over a differential control volume.

There are several correlations in the literature to obtain droplet size distribution at the atomizing zone [10, 11]. The size distribution evaluated by these equations are significantly different. This can be attributed to the different liquid injection systems and measurement techniques used in their experiments. Goncalves *et. al.* [9] have shown that Rosin-Rammler distribution function could well fit the droplet size distribution of an atomized liquid jet. The volume frequency distribution function of Rosin-Rammler is given as follows:

$$f_d(D_d) = n\left(\frac{D_d^{n-1}}{X^n}\right) EXP\left[-(D_d/X)^n\right] \tag{2}$$

Where $f_d(D_d)$ is the volume fraction frequency distribution of the drop size, ($f_d(D_d)dD_d$ is the volume fraction of droplets having a size between D_d and D_d+dD_d), and n and X are two parameters of the above equation. The value of parameter n can be estimated equal to 2 [12]. For calculating X, the value of sauter mean diameter proposed by Boll *et al*. [13] was used. X was calculated by taking equal the sauter mean diameter evaluated from the Distribution function and the one by Boll's correlation. Goncalves *et. al*. [9] have concluded that sauter mean diameter of the droplets was well correlated by Boll's equation, which is as follows:

$$D_{32} = \frac{101300 + 23\frac{L}{G}}{V_{g_0}^{\;1.75}} \tag{3}$$

Number concentration frequency distribution function, N_d(Dd), was evaluated by the following equation:

$$N_d(D_d) = \frac{f_d(D_d)}{D_d^{\;3}\int_{D_{dmin}}^{D_{dmax}} \frac{f_d(D_d)}{D_d^{\;3}} dD_d} \tag{4}$$

Normalized flux, which is the ratio of local flux to uniform flux, can be obtained by solving the following integral:

$$f_l = \int_{D_{dmin}}^{D_{dmax}} \frac{\frac{\pi}{6} D_d^3 V_{d_x} C_d(D_d)}{\frac{L_0}{A}} dD_d \tag{5}$$

In Eq. (1) the droplets are convected in x direction, while they are dispersed in y and z direction by convection and eddy diffusion. In addition, it is assumed that droplets are generated by a point source. This point source is located by an empirical correlation for calculating liquid jet penetration length which has been obtained by Viswanathan *et. al*. [14]:

$$\frac{h^*}{D_j} = 0.1145 \frac{V_j \rho_l}{Vg \rho_g} \tag{6}$$

In order to obtain boundary conditions, a physical model must be considered for the droplets dispersion. When droplets dispersing across the cross section of a scrubber reach the walls, they collect on the wall as liquid film. At the same time, some of the liquid film formed by impacting droplets on the walls may be reatomized by the gas stream. If it is assumed that the rate of collecting droplets on the walls is low or equal to the rate of reatomization, the net flux at the wall will be approximately zero. In this model, based on the above concept, the following boundary condition is used for solving Eq. (1):

$$\left[\frac{\partial C_d(D_d)}{\partial y}\right]_{wall} = \left[\frac{\partial C_d(D_d)}{\partial z}\right]_{wall} = 0 \tag{7}$$

The velocity of droplets at y and z directions were taken az zero. The x-component droplet velocity was calculated by the following equation expressing a one-dimensional droplet momentum balance:

$$\frac{dV_{dx}}{dx} = \frac{3}{4}\frac{C_{Df}}{D_d}\frac{\rho_g}{\rho_l}\frac{(V_g - V_{dx})\left|V_g - V_{dx}\right|}{V_{dx}} \tag{8}$$

In the above equation C_{Df} is the drag coefficient, determined by the relation developed by Sartor and Abbott [15] for the accelerated motion of water drops in the air:

$$\begin{cases} C_{Df} = \dfrac{24}{Cc.Re_P} & Re_P \langle 0.1 \\ C_{Df} = \dfrac{24}{Cc.Re_P}(1 + 0.0916\,Re_P) & 0.1 \langle Re_P \langle 5.0 \\ C_{Df} = \dfrac{24}{Cc.Re_P}(1 + 0.158\,Re_P^{2/3}) & 5.0 \langle Re_P \langle 1000 \end{cases} \tag{9}$$

where Cc is the Cunningham correction factor which can be calculated using $Cc = 1 + Kn[\gamma_1 + \gamma_2.\exp(-\gamma_3 / Kn]$ [16], with Kn=2λ/Dp, γ1=1.231, γ2=0.4695, γ3=1.1783 and λ, which is the mean free path of molecules and can be taken equal to 65 nm for operational conditions. This correction factor is 1.16 and 1.08 for particles having size 1 and 2 μm respectively.

The eddy diffusivity and mixing length of droplets are correlated to the eddy diffusivity and Prandtl mixing length of gas in the scrubber. Therefore it is necessary to calculate these two parameters for the gas. The gas eddy diffusivity is obtained by multiplying Prandtl mixing length, l_g, to the mean fluctuation velocity, $\hat{V}'_g$:

$$E_g = l_g \hat{V}'_g \tag{10}$$

The mean fluctuation velocity of gas can be estimated by using shear velocity, V_0, [17] which is calculated by the following equation:

$$V_0 = \sqrt{\frac{\tau_0}{\rho_g}} \tag{11}$$

In which τ_0 is the shear stress between gas and liquid film. The value of τ_0 can be calculated from a force balance including a two-phase pressure drop and wall shear stress. For this purpose, the pressure drop was calculated using the Hagedorn and Brown [18] correlation in mist-annular two-phase flow.

It is well accepted in the literature that for various systems at high Reynold numbers, the ratio of $V_g D/E_g$ is a constant [19]. Hence, the value of gas eddy diffusivity, E_g, can be correlated to the Peclet number, which is constant for a high velocity gas stream in a tube:

$$N_{Pe} = \frac{V_g D}{E_g} \tag{12}$$

Baldwin and Walsh [19] have reported Peclet numbers for fully developed turbulent flow. By knowing E_g and $\hat{V}'_g$, the value of l_g can be found by Eq. (10).

The mixing of droplets across the scrubber is the result of the eddy diffusivity of droplets which is given by a similar equation. In order to evaluate the mixing length of the droplets, the following equation expressing force balance for drops due to gas fluctuation velocity must be solved:

$$\rho_l(\frac{\pi}{6}D_d^3)\frac{dV'_d}{dt}=\frac{1}{2}\rho_g\frac{\pi}{4}D_d^3C_{D_f}(V'_g-V'_d)^2 \tag{13}$$

The value of C_{Df} was obtained by the following linear equation which is applicable for N_{Re} between 2-20 :

$$C_{D_f}=\frac{30.2}{N_{Re}} \tag{14}$$

Replacing V'_d by dz/dt in Eq. (13) will result in:

$$\frac{d^2z}{dt^2}=A(V'_g-\frac{dz}{dt}) \tag{15}$$

In which A is:

$$A=\frac{225\mu}{\rho_l D_d^2} \tag{16}$$

The initial conditions are:

$$at \quad t=0, \quad z=0, \frac{dz}{dt}=0 \tag{17}$$

By solving the above equation the path of droplet moving due to gas velocity fluctuation can be predicted as follows:

$$z=\int_0^t e^{-At}(\int_0^t AV'_g e^{At'}dt')dt \tag{18}$$

The value of z at $t=T=l_g/\hat{V}'_g$ is equal to the droplet mixing length. T is the mean time that each individual eddy persists as an entity:

$$T=\frac{l_g}{\hat{V}'_g}=\frac{l_d}{\hat{V}'_d} \tag{19}$$

In order to solve the integral that appeared in Eq. (18), the relation of gas fluctuation velocity with time should be known. As an estimate, the following equation was used to correlate the gas fluctuation velocity with time:

$$V'_g=a\mathrm{Sin}[(\frac{2p}{T})t] \tag{20}$$

The altitude of fluctuation, a, can be obtained by the mean fluctuation velocity relation:

$$\int_0^T V_g'^2 dt=\hat{V}_g'^2 \tag{21}$$

Combining Eqs. (8) and (19), the ratio of droplet eddy diffusivity to gas eddy diffusivity can be obtained:

$$\frac{E_d}{E_g}=\frac{l_d\hat{V}'_d}{l_g\hat{V}'_g}=\frac{l_d\frac{l_d}{T}}{l_g\frac{l_g}{T}}=\frac{l_d^2}{l_g^2} \tag{22}$$

Further details of evaluating the parameters are given by [5] .

In order to find the variation of particulate concentration throughout the scrubber the following differential mass balance for particles must be solved:

$$\frac{\partial[V_{g_x} C_p(D_p)]}{\partial x} = E_g\{\frac{\partial^2[C_p(D_p)]}{\partial y^2} + \frac{\partial^2[C_p(D_p)]}{\partial z^2}\} + \int_{D_{d\min}}^{D_{d\max}} \frac{\pi}{4} D_d^2 \eta_t (V_g - V_d) C_p(D_p) C_d(D_d) dD_d \quad (23)$$

Where $C_P(D_P)$ is the number concentration frequency distribution of particulate, ($C_P(D_P)dD_P$ is the concentration of particles having a size between D_P and D_P+dD_P), and η_t is the removal efficiency of a single droplet. The value of η_t can be found by the following equation given by Calvert *et. al* [20]:

$$\eta_t = (\frac{\Psi}{\Psi + 0.7})^2 \quad (24)$$

In the above equation ψ is the inertia impaction parameter and is given by the following equation:

$$\Psi = \frac{\rho_p D_p^2 |V_g - V_d|}{9\mu_g D_d} \quad (25)$$

The total concentration of particles with different sizes can be obtained by solving the following integral:

$$C_p = \int_{D_{p\min}}^{D_{p\max}} C_p(D_p) dD_p \quad (26)$$

In order to find the particle size cumulative curve the following equation was used:

$$C_P^c(D_P) = \frac{\int_{D_{P\min}}^{D_P} C_P(D'_P) dD'_P}{\int_{D_{P\min}}^{D_{P\max}} C_P(D'_P) dD'_P} \quad (27)$$

3. METHOD OF SOLUTION

The upwind control volume method was used to solve Eqs. (1) and (22). The details of using this method are given by Patanakar [21]. The main problem of using this method is false diffusion. False diffusion error can be found only at a diverging section where the direction of the velocity vector is not perpendicular to the surface of the control volume. In order to reduce this error, finer grids were used at diverging sections.

The integrals in Eqs. (6), (22), (25) and (26) should be calculated numerically. There are several numerical integration methods, namely Gaussian quadrature method, Sympson's rule and trapazoidal rule. The gaussian quadrature with n points provides the same general order of accuracy as does Sympson's rule with 2n points and the trapazoidal rule with 4n points. For this reason Gaussian quadrature method was used as a numerical integration method. By using this method the number of iterations in the calculation procedure, and hence CPU computer time, are reduced to about 25%. The general form of this method can be presented mathematically as follows:

$$\int_{D_{\min}}^{D_{\max}} g(D) = \frac{D_{\max} - D_{\min}}{2} \sum_{i=1}^{n} A_i\, g\left(\frac{D_{\max} - D_{\min}}{2} U_i + \frac{D_{\max} + D_{\min}}{2}\right) \tag{28}$$

Where n is the number of points at which the function of g should be calculated, and Ai and Ui are the specified values which can be found in the literature [22]. These points are determined by the following equation:

$$D_{d_i} = \frac{D_{\max} - D_{\min}}{2} U_i + \frac{D_{\max} + D_{\min}}{2} \tag{29}$$

4. RESULTS

Since, in the real situation the formed droplets have different sizes and the size distribution of droplets does not remain constant throughout the scrubber, it is clear that droplet size distribution must be taken into account in realstic models. Figures 1-5 show the comparison between experimental data of Viswanathan *et al*. [1] and the results of the mathematical model based on both drop size distribution and mean drop size for various operating conditions. The experimental data reported by Viswanathan *et al*. [1] belongs to an air water system taken in a venturi scrubber having a throat with a dimension of 7.5×15 cm.. Parameter n provides a measure of the spread of the droplet size. The higher the value of n, the more uniform the droplet sizes. The results provided in Figs. 2-6 do not show any significant advantage of using droplet size distribution in the drop dispersion prediction. However, considering that drop size distribution can influence particle removal calculations, the results of the mathematical model for predicting particulate removal efficiency are compared with the Brink and Contant [2] experimental data in Figs. 7-9. Apparently, the results of the model are in better agreement with the experimental data when droplet size distribution is considered.

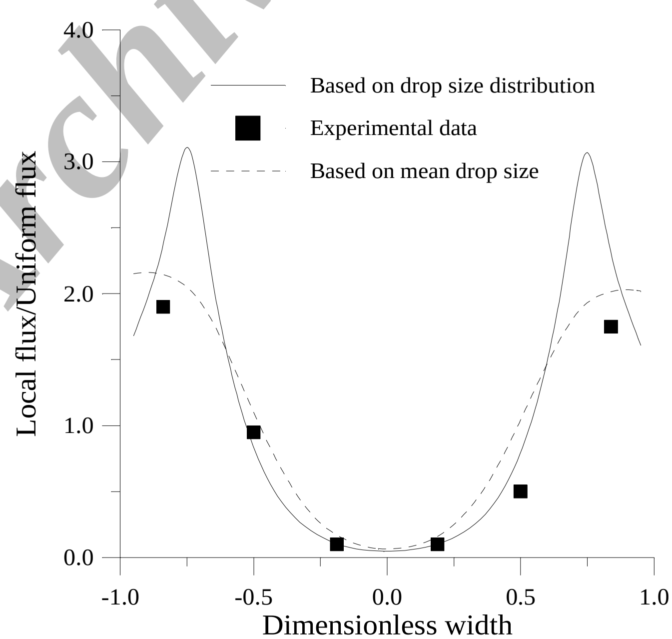

Fig. 2. The comparison between calculated result based on both considering drop size distribution and mean drop size and experimental data of Viswanathan *et al*. (1984), L/G=0.4

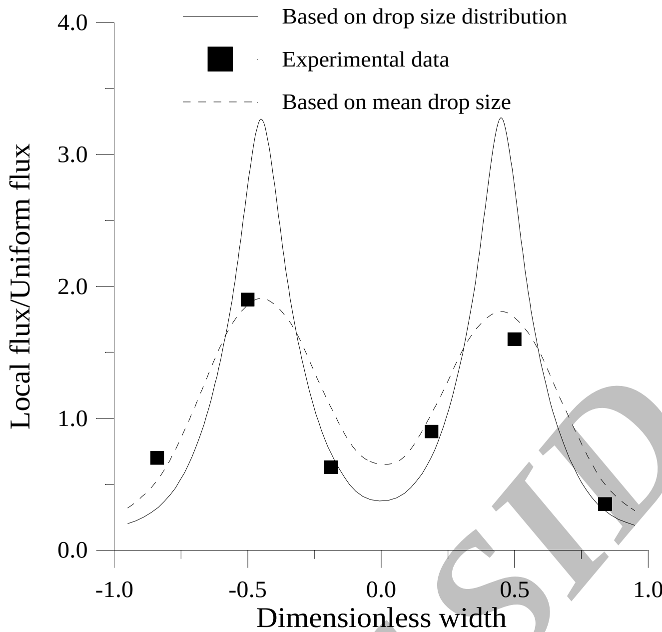

Fig. 3. The comparison between calculated result based on both considering drop size distribution and mean drop size and experimental data of Viswanathan *et al*. (1984), L/G=0.93

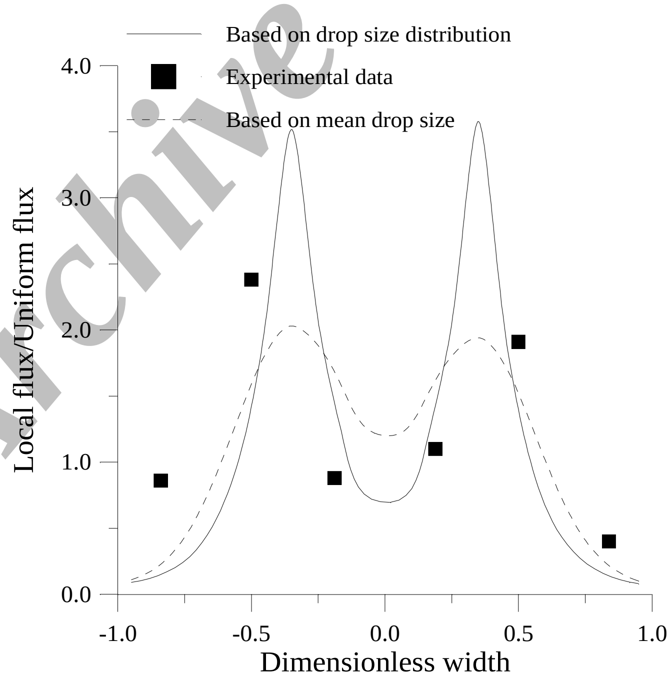

Fig. 4. The comparison between calculated result based on both considering drop size distribution and mean drop size and experimental data of Viswanathan *et al*. (1984), L/G=1.2

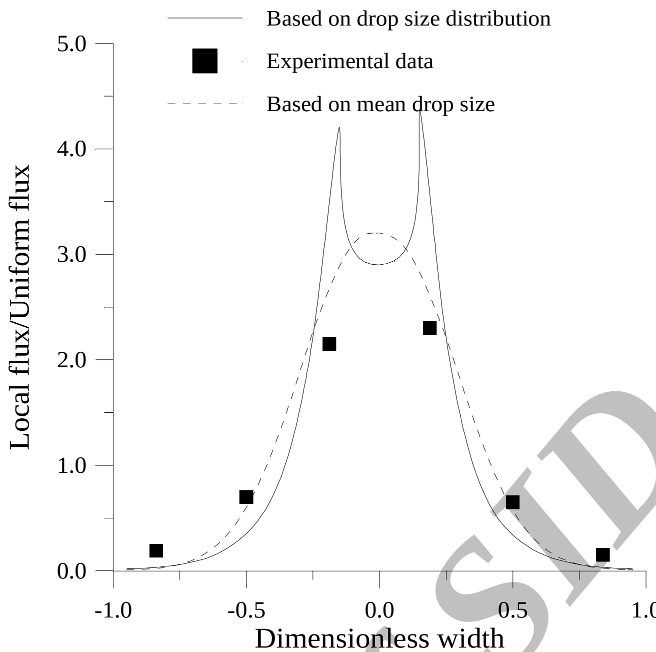

Fig. 5. The comparison between calculated result based on both considering drop size distribution and mean drop size and experimental data of Viswanathan *et al*. (1984), L/G=1.47

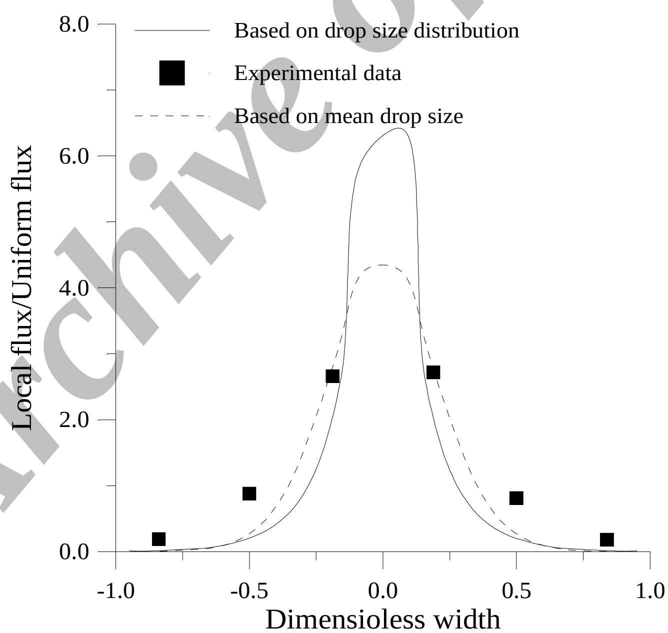

Fig. 6. The comparison between calculated result based on both considering drop size distribution and mean drop size and experimental data of Viswanathan *et al*. (1984), L/G=1.79

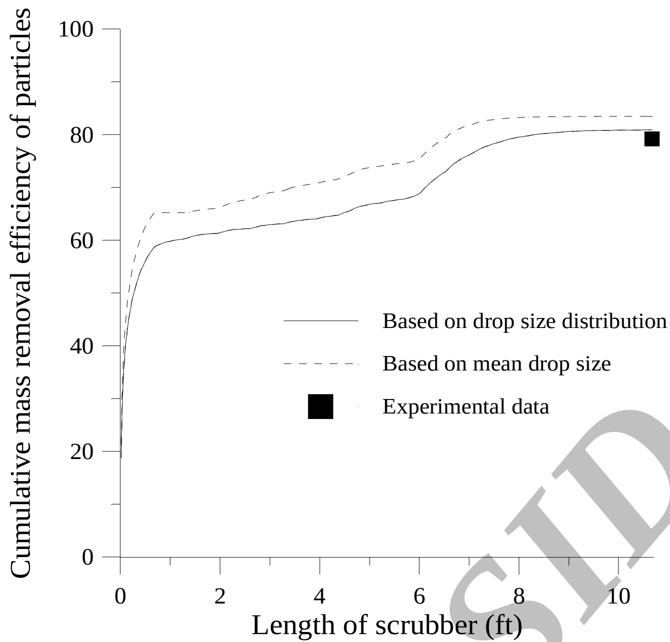

Fig. 7. The comparison between cumulative mass removal efficiency calculated based on drop size distribution and mean drop size and experimental data of Brink and Contant [1], for particle diameter of 0.5 μm

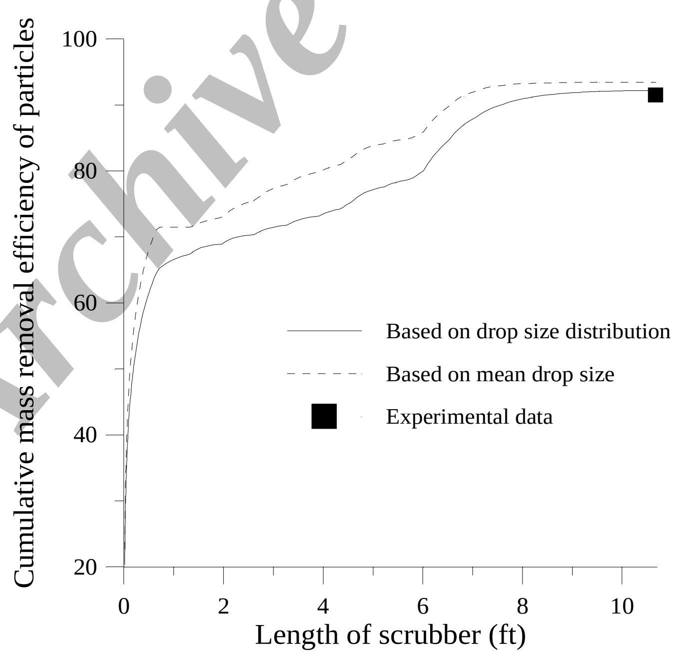

Fig. 8. The comparison between cumulative mass removal efficiency calculated based on drop size distribution and mean drop size and experimental data of Brink and Contant [1], for particle diameter of 0.65 μm

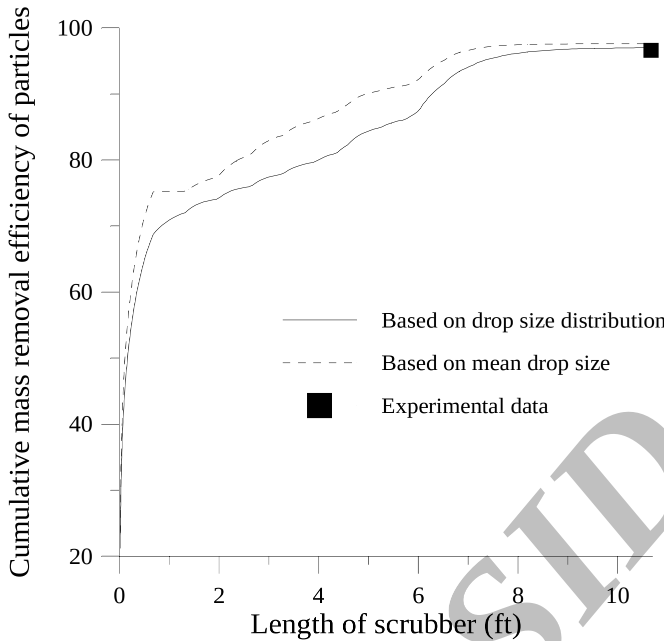

Fig. 9. The comparison between cumulative mass removal efficiency calculated based on drop size distribution and mean drop size and experimental data of Brink and Contant [1], for particle diameter of 0.81 μm

Figure 10 shows the comparison between the calculated cummulative particle size distribution and the experimental data at the outlet of the scrubber. As can be seen in these figures, the results are in good agreement with the experimental data.

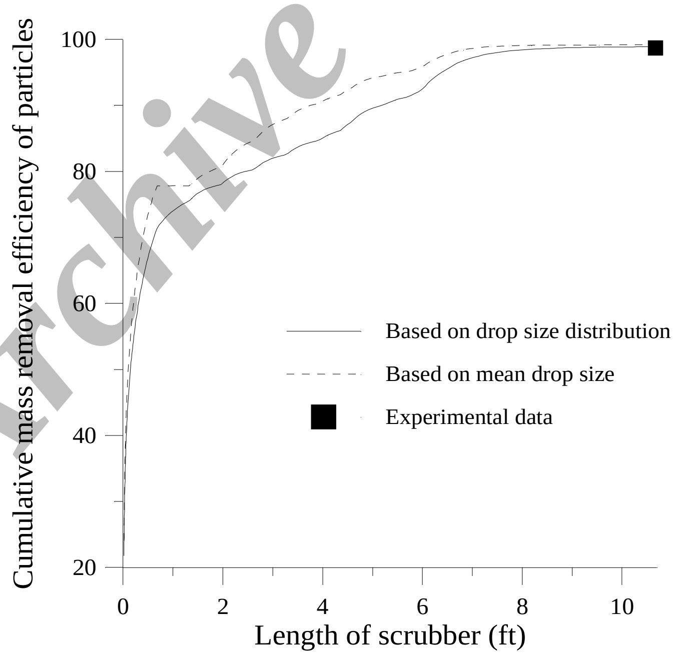

Fig. 10. The comparison between cumulative mass removal efficiency calculated based on drop size distribution and mean drop size and experimental data of Brink and Contant [1], for particle diameter of 1 μm

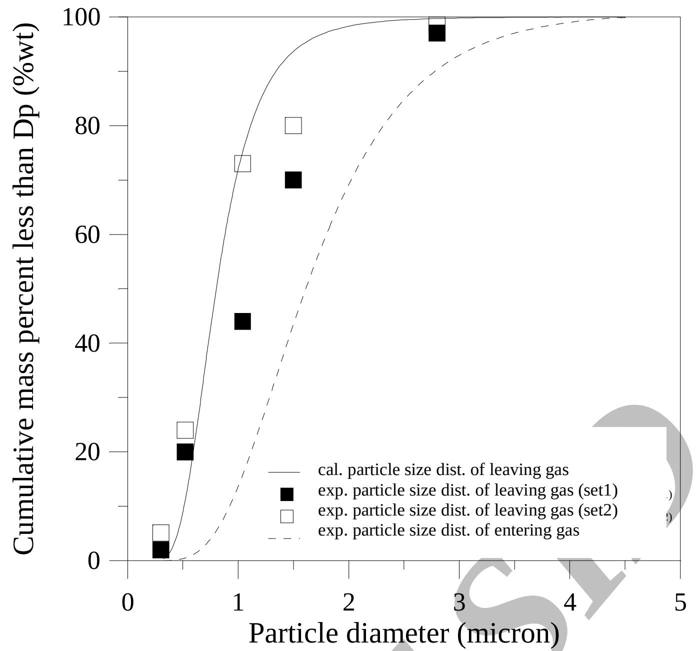

Fig. 11. The comparison between calculated pareticle size distribution at the outlet of the scrubber and experimental data of Brink and Contant (1958)

Figure 8 shows the variation of particle removal efficiency with liquid jet velocity. The dimension of the scrubber used to obtain these results was considered the same as that used by Brink and Contant. The penetration of the liquid jet increases as liquid jet velocity increases. At low liquid jet velocity the penetration is low and droplets cannot be uniformly distributed over the scrubber. By increasing this velocity, to some extent, droplets are dispersed more uniformly, so particle removal efficiency increases. As can be seen in this figure, particle removal efficiency, calculated based on droplet size distribution, lays under that based on mean drop size.

5. CONCLUSION

A three-dimensional dispersion model was developed to include the size distribution of droplets into the mathematical model of venturi srubber performance. By using this model the effect of droplet size distribution on particle removal efficiency of venturi scrubbers was investigated. The results of the model reveal that drop size distribution causes a reduction in liquid dispersion and uniformity of droplet concentration distribution. Consequently, it can be concluded that considering drop size distribution does decrease particle removal efficiency in venturi scrubbers.

NOMENCLATURE

A throat cross sectional area of scrubber (m^2)

Cd total droplets concentration ($No./m^3$)

Cd(Dd) number concentration frequency distribution of drop size ($No./m^3/m$)

Cp total particle concentration (g/m^3)

Cp(Dp) number concentration frequency distribution of particle size ($g/m^3/m$)

$C^c p(Dp)$ concentration comulative distribution of particle size (g/m^3)

C_{Df} drag coefficient of drop (dimensionless)

D_{32} mean diameter of droplets (m)

D_j diameter of nozzle (m)

E_d eddy diffusivity of droplets (m^2/s)

E_g eddy diffusivity of gas (m^2/s)

$f_d(D_d)$ volume fraction frequency distribution of drop size (m^{-1})

G_0 gas flow rate (m^3/s)

L_0 liquid flow rate (m^3/s)

f_l local flux to uniform flux (dimensionless)

h* vertical penetration length (m)

L_{0T} total liquid flow rate (m^3/s)

l_g Prandtl mixing length of gas (m)

l_d Prandtl mixing length of droplets (m)

$N_d(D_d)$ number fraction frequency distribution of drop size (No./m^3/m)

N_{Pe} Peclet number (V_g D/E_g) (dimensionless)

N_{Re} Reynold number (VD/μ)(dimensionless)

S source strength (No./m^3.s)

t time (s)

T mean time that each individual eddy persists as an entity (s)

V_g gas velocity (m/s)

V_d droplets velocity (m/s)

V_0 shear velocity (m/s)

$\hat{V}_g'$ mean fluctuation velocity of gas (m/s)

$\hat{V}_g'$ fluctuation velocity of droplets (m/s)

V_j liquid jet velocity (m/s)

x length (m)

y height (m)

z width (m)

Greek symbols

ρ_g density of gas (Kg/m^3)

ρ_l density of liquid (Kg/m^3)

ρ_j density of liquid jet (Kg/m^3)

μ_g viscosity of gas (Kg/m.s)

μ_l viscosity of liquid (Kg/m.s)

τ_0 shear stress between liquid film and gas (P0a)

ν kinematic viscosity (m^2/s)

Subscripts

d droplets

g gas phase

x x direction

y y direction

z z direction

0 throat section

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