CHAPTER V

BUBBLE DYNAMICS

Bubble size distribution, swarm rise velocity, and number of bubbles were obtained using the dynamic gas disengagement (DGD) method. Theory and sample calculation for this technique are described in APPENDIX F.

A. Data Acquisition

A videocamera and a VCR unit were used to record the drop in liquid level during the disengagement process. The actual heights of liquid level were obtained from the ruler mounted adjacent to the column. After the completion of a run with a given gas velocity, the magnetic valve was shut off. The liquid level dropped sharply during the first 1 to 10 seconds, due to the disengagement of large bubbles. Thereafter, the level dropped slowly as the medium and the small bubbles disengaged.

The data analysis procedure involved replaying the video tape to obtain different values of height (H) and the time elapsed (t). The data were entered directly on a personal computer (Hp AT-compatible). During the initial period when the level dropped rapidly, data were recorded after every $\sim$ 0.05 m drop in level and about 0.005 m towards the end. This procedure was repeated three times for each velocity to reduce errors. Following this step, the normalized height was plotted on the computer display and appropriate break points were then selected similar to those shown in Figure 35. The slopes and intercepts of the straight line segments were then computed and the corresponding swarm bubble rise velocity and hold-ups fractions obtained. A computer program was used to calculate bubble

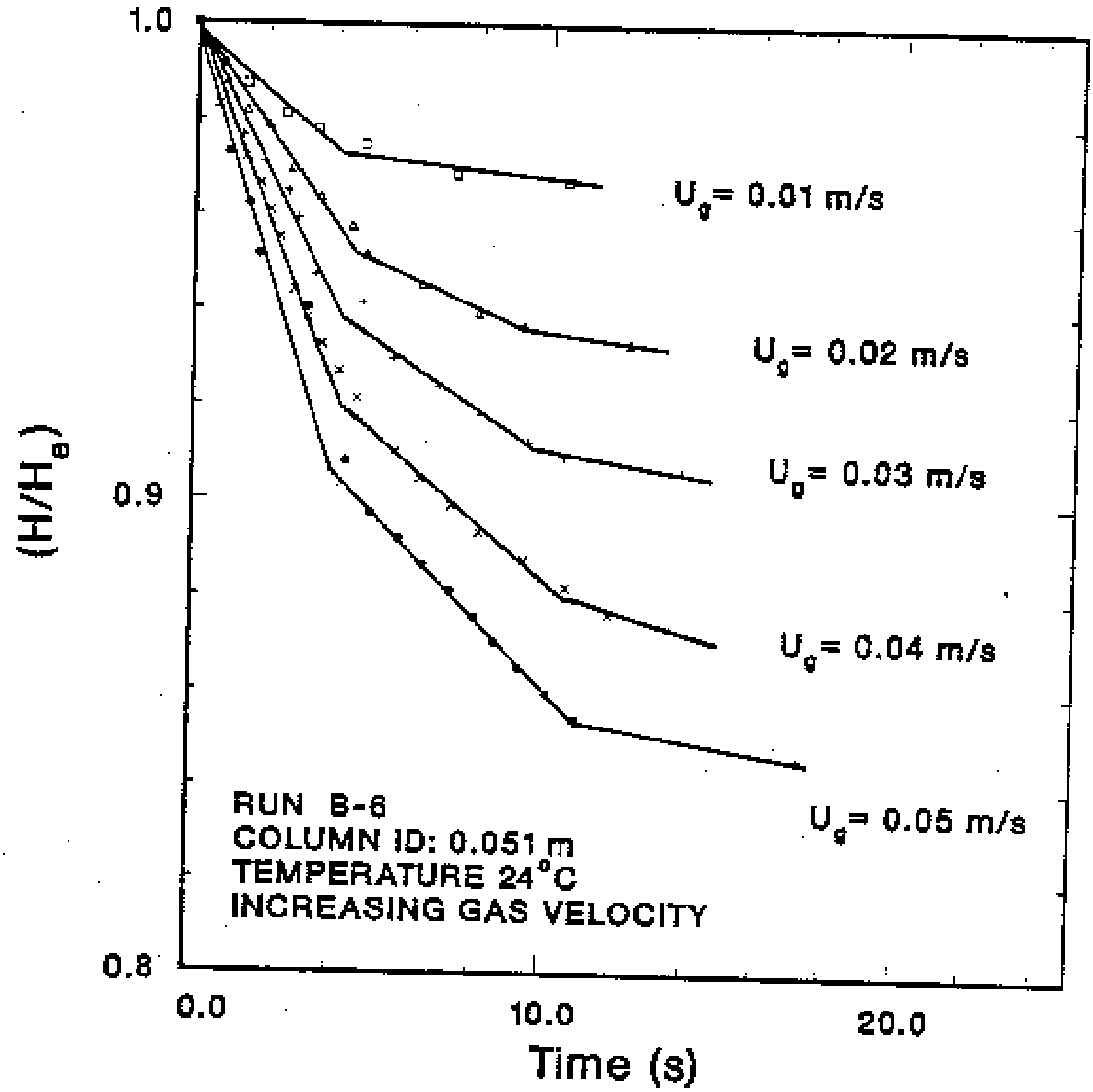

Figure 35. Change in normalized height as a function of time and velocity, example from run B-6.

sizes for each class of bubbles, the number of bubbles in each class, the Sauter mean diameter, and finally the specific gas-liquid interfacial area.

B. Procedure for Obtaining Bubble Sizes

Marrucci (1965) presented a correlation for determining terminal rise velocity of a single bubble from the swarm bubble velocity as follows:

$$u_{Bi\infty} = u_{Bi} \frac{1 - \epsilon_i^{5/3}}{(1 - \epsilon_i)^2} \tag{5.1}$$

where i is the bubble size classification, $u_{Bi\infty}$ is the terminal rise velocity of a bubble with size i, u_{Bi} is the swarm bubble velocity associated with bubbles of class i, and ϵ_i is the average gas hold-up value due to bubbles of size i, where the values of u_{Bi} and ϵ_i were estimated from the DGD technique.

The application of Marruccis's correlation is based on the assumption that, rising bubbles are spherical and have N_{Re} in the range of 1 and 300. This correlation was used to estimate the rise velocity of small and medium bubble sizes using their swarm velocities and their average gas hold-up values. No attempt was made to calculate the rise velocities of slugs and large bubbles from the swarm velocity. For slugs it is assumed that they rise at velocity equal to terminal rise velocity (Davidson and Harrison, 1966). In this study different correlations were used to obtain bubble sizes depending on the type of liquid media and group of bubble size (small, medium, large, or slug).

1. Fischer Tropsch Derived Waxes

Bubble sizes for waxes were obtained using the generalized bubble rise velocity correlation presented by Abou-el-Hassan (1983). This correlation accounts for parameters affecting bubble rise velocity as well as the interaction of bubbles. The correlation has been recommended for use with Newtonian fluids covering the following ranges:

liquid density = 710 to 1180 kg/m^3

liquid viscosity = 0.233 to 59 mPa.s

interfacial tension = 0.015 to 0.072 N/m

The correlation is given by:

$$V = 0.75 \, [\log(F)]^2 \tag{5.2}$$

where V (velocity number) and F (flow number) are defined as:

$$V = \frac{u_{Bi\infty} d_{Bi}^{2/3} \rho_l^{2/3}}{\mu_l^{1/3} \sigma_l^{1/3}} \tag{5.3}$$

$$F = \frac{g d_{Bi}^{8/3} (\rho_l - \rho_g) \rho_l^{2/3}}{\mu_l^{4/3} \sigma_l^{1/3}} \tag{5.4}$$

This correlation is valid for velocity numbers in the range of 0.1 to 40.0 and flow number in the range of 1 to 10^6. For bubble rise velocities greater than 0.15 m/s, the following correlation by Clift *et al.*, 1978 was used to estimate the bubble sizes:

$$u_{Bi\infty} = \sqrt{\frac{2.14\sigma_l}{\rho_l d_{Bi}} + 0.505 g d_{Bi}} \tag{5.5}$$

For the range of bubble rise velocities not covered by the above correlations, bubble diameters were obtained by interpolation. For example, Figure 36 presents the relation between rise velocity and bubble diameter for FT-300 wax at 265 °C.

2. Pure Liquids

Peebles and Garber, (1953) presented the correlation for estimation of air bubble sizes from their rise velocity in water as follows:

$$u_{Bicc} = 0.33g^{0.76}\left(\frac{\rho_l}{\mu_l}\right)^{0.52}\left(\frac{d_{Bi}}{2}\right)^{1.28} \qquad 2 < N_{Re} < 4.02G_1^{-0.21} \qquad (5.6)$$

where

$$G_1 = \left(\frac{g\mu_l^{4}}{\rho_l\sigma_l^{3}}\right)$$

In terms of bubble rise velocity, the above equation was applicable up to bubble rise velocity of 0.20 m/s, whereas, for N_{Re} less than 2 Stokes equation (5.8) was used to estimate the bubble sizes. The correlation by Clift *et al.*, 1978 (Equation 5.5) was used to estimate the bubble sizes for bubbles with rise velocity greater than 0.20 m/s. (medium and large bubble sizes). For example, Figure 37 presents the rise velocity versus bubble diameter for n-butanol using appropriate equations.

3. Aqueous Solutions of Butanol With CMC

Small and Medium Bubble Sizes

Clift *et al.* (1978) presented a correlation for calculating the terminal rise velocity on the basis of an extensive range of data as follows:

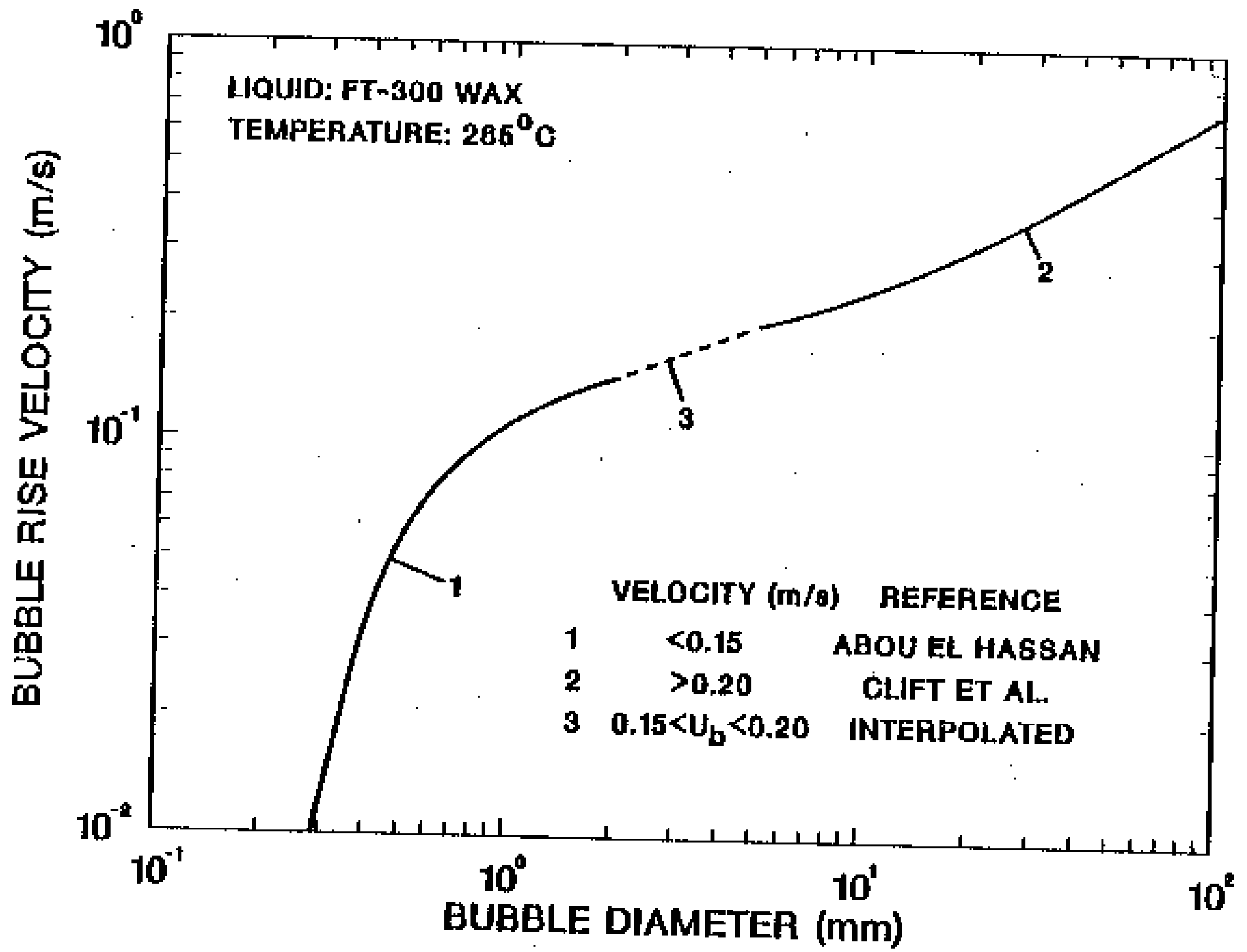

Figure 36. Bubble rise velocity versus equivalent bubble diameter for FT- 300 wax.

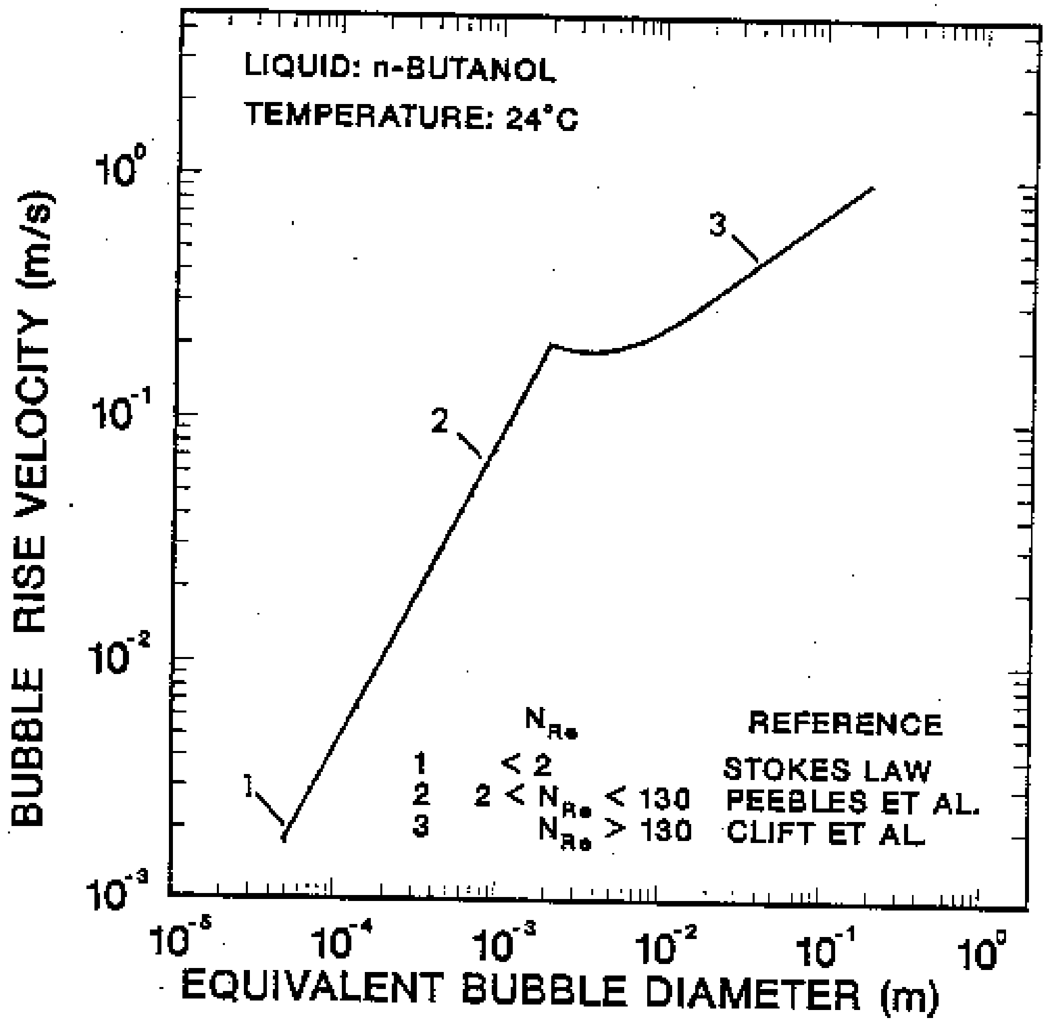

Figure 37. Bubble rise velocity versus equivalent bubble diameter for n-butanol.

$$u_{Bi\infty} = \frac{\mu_l}{\rho_l d_{Bi}} \ M^{-0.149} \ (J - 0.857) \tag{5.7}$$

where

$$M = \frac{g \ \mu_l^4 \ (\rho_l \ - \ \rho_g)}{\rho_l^2 \ \sigma^3}$$

$$N_{Re} = \frac{d_{Bi} \ u_{Bi\infty} \ \rho_l}{\mu_l}$$

$$E_o = \frac{g \ (\rho_l \ - \ \rho_g) \ d_{Bi}^2}{\sigma}$$

$$J = 0.94 \ H^{0.747} \qquad (2 \ < \ H \ \geq \ 59.3)$$

$$J = 3.42 \ H^{0.441} \qquad (H \ > \ 59.3)$$

and

$$H = \frac{4}{3} \ E_o \ M^{-0.149} \ (\frac{\mu_l}{\mu_w})^{-0.14}$$

for the range $M \ < \ 10^{-3}$, $E_o \ < \ 40$, and $N_{Re} \ > 0.1$.

This correlation was developed for predicting rise velocities of bubbles and drops for contaminated mediums, and it has been recommended by Shah *et al.* (1982) and by Haque *et al.* (1987) for use in highly viscous non-Newtonian liquids as long as the above conditions are satisfied. With the known values of bubble rise velocities, the corresponding bubble sizes were calculated.

For the range of fl < 2 (small bubbles), Stokes law was assumed to be valid, that is:

$$u_{Bi\infty} = \frac{g\, d_{Bi}^{\,2}\, (\rho_l - \rho_g)}{18\mu_l} \qquad N_{Re} \leq 2 \tag{5.8}$$

In this case the small bubbles are considered to be rigid spheres of the same size and low rise velocity.

Large Bubbles and Slugs

The correlation developed by Mendelson (1967) using wave analogy to predict bubble rise velocity for large bubbles has been used by Acharya and Ulbrecht (1978) and by Schügerl (1981) to describe terminal rise velocities of large bubbles and slugs at large Reynold numbers. The correlation has been recommended by the above workers for applications in Newtonian and non-Newtonian aqueous solutions. Mendelson's correlation is given as:

$$u_{Bi\infty} = \sqrt{\frac{2\sigma_l}{d_{Bi}\rho_l} + \frac{g d_{Bi}}{2}} \tag{5.9}$$

This correlation is applicable to bubbles with rise velocities greater than 0.25 m/s.

Figure 38 shows equivalent bubble diameters as a function of bubble rise velocities using the above correlations. At low terminal velocities (< 0.25 m/s), three different curves were obtained representing different rheological properties of the mixtures (Newtonian solutions, 1.0 and 0.5 wt % aqueous solutions of butanol; and the non-Newtonian solutions, 1.0 wt % butanol with 0.1 wt % CMC solution and 1.0 wt % butanol with 0.5 wt % CMC solution) used in the present study (the presence of CMC results in decrease in rise velocity of bubbles due to increased

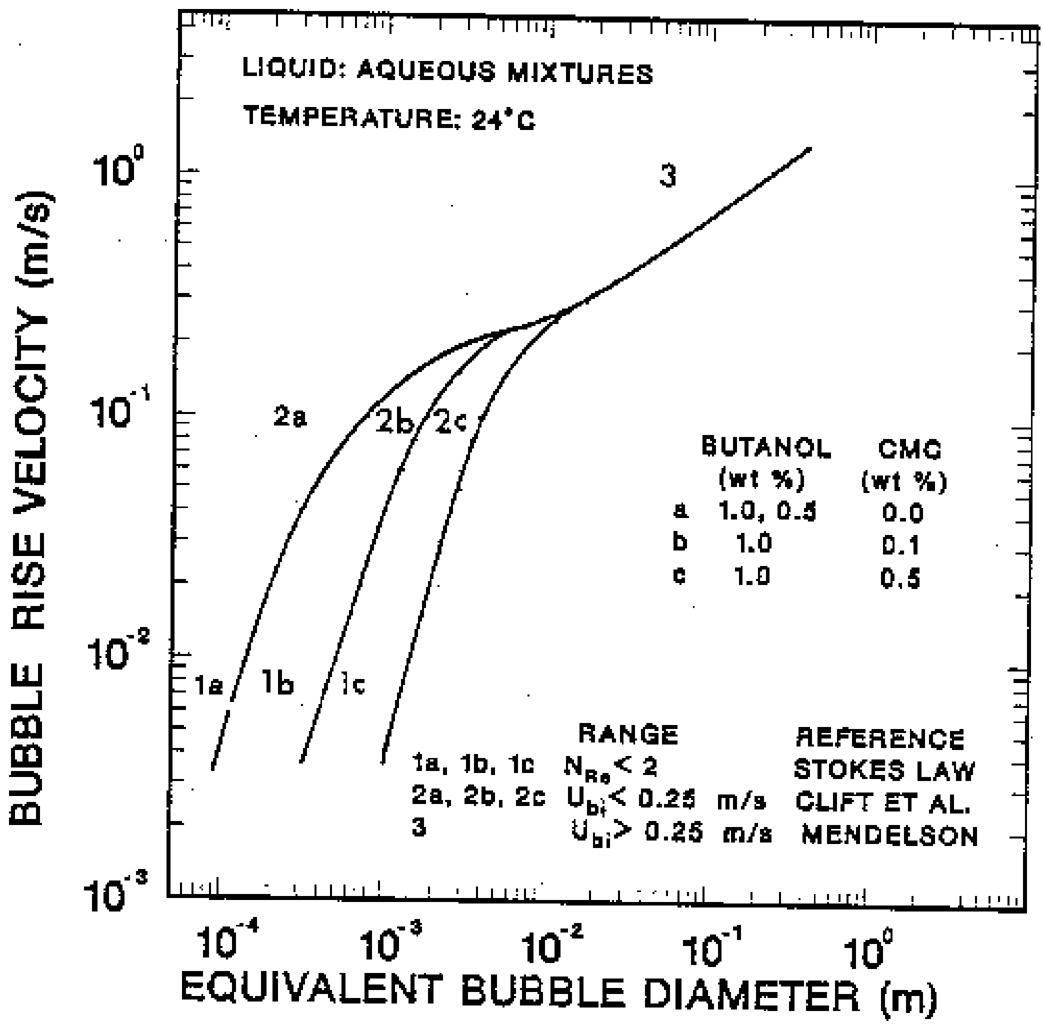

Figure 38. Bubble rise velocity versus equivalent bubble diameter for aqueous solutions.

drag friction on the ascending bubbles). At high bubble rise velocities (> 0.25 m/s) there is no effect of drag friction on the ascending bubbles, because the motion of the bubbles and the coalescence rate are much higher due to turbulent flow in the column. For the small column, large bubbles develop into slugs due to coalescence; whereas, in the large column, the motion of the bubbles becomes much more violent (no slugs).

CHAPTER VI

SAUTER MEAN BUBBLE DIAMETER

A. FT Derived Waxes

1. Effect of Operating Temperature

Figure 39 presents the effect of temperature and superficial gas velocity on d_s. The Sauter values decreased as the temperature increased. Similar trends were observed for all types of waxes. Moreover, the Sauter mean diameter increased as the gas velocity increased, therefore results for Sasol wax are discussed below. At low superficial gas velocities (u_g between 0.01 and 0.03 m/s) the Sauter mean diameters for Sasol wax at 200 °C and at 265 °C were about 3.0 mm and 1.0 mm respectively. However, the difference increased as the superficial gas velocity increased and at 0.09 m/s the Sauter mean diameters at 200 °C were about twice those at 265 °C (4.7 mm compared to 2.0 mm).

The significant difference in d_s at 200 and 265 °C for these waxes can be due to the difference in viscosities at the two temperatures. At 200 °C, the viscosities of these waxes were about 50 to 65 % higher than viscosities at 265 °C. These observations are in agreement with results reported by Schurgel (1981) and results obtained in this study with non-Newtonian liquid mediums, where it was observed that increasing liquid viscosity makes the liquid medium more coalescing, therefore large bubbles are formed in this process resulting in high Sauter mean bubble diameters.

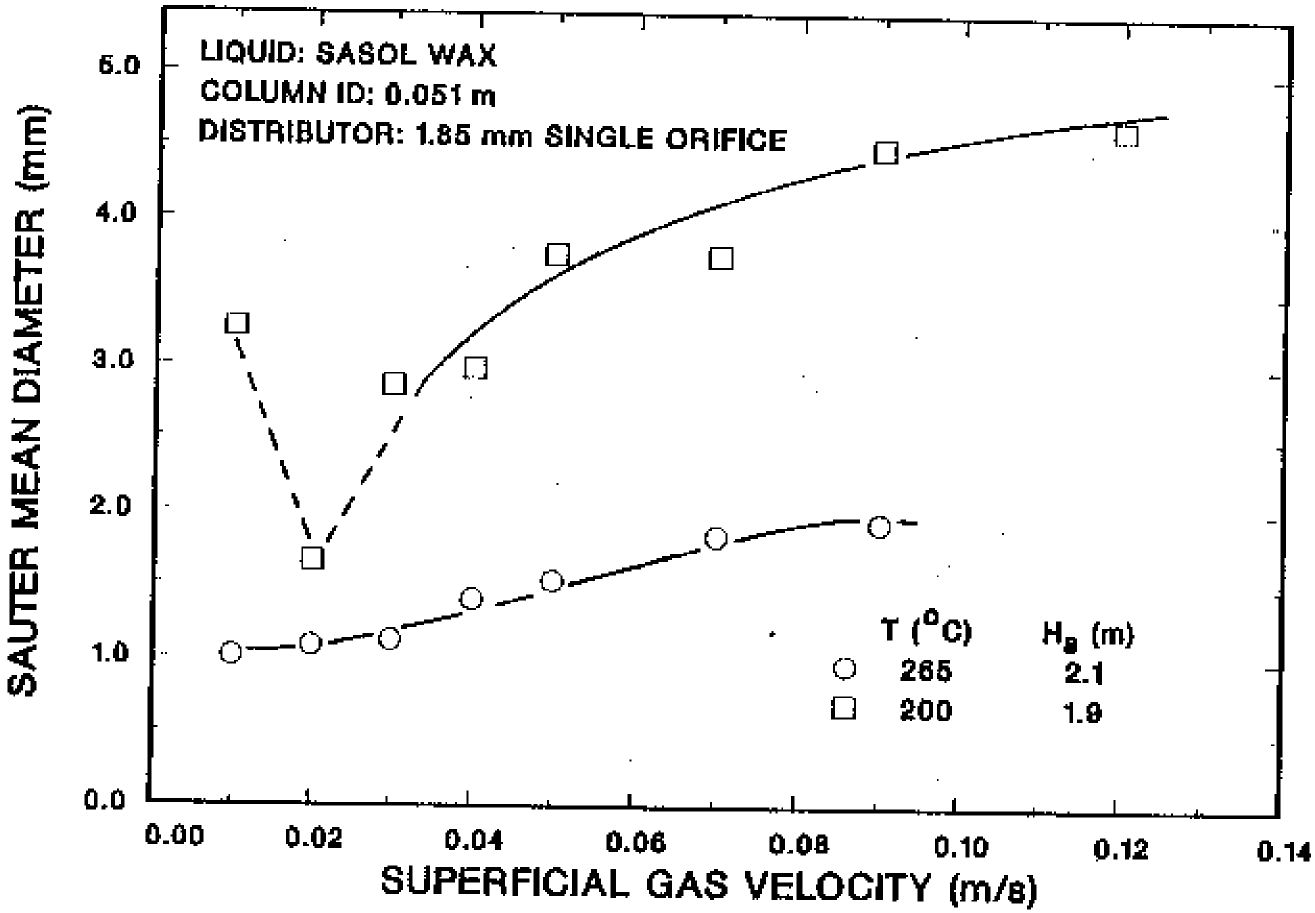

Figure 39. Effect of temperature and superficial gas velocity on Sauter mean bubble diameter (□ -Run 8-3; ○ -Run 8-4).

2. Effect of Distributor Type

Figure 40 shows results from experiments conducted with mobil reactor wax using 1.85 mm orifice plate and SMP distributors at 265 °C. The sauter mean diameter increased with superficial gas velocity from 1 mm at 0.01 m/s to about 5 mm at 0.12 m/s using the 1.85 mm orifice plate distributor. These values are about 20 % higher than those obtained with the SMP distributor at most velocities. Similarly with Sasol reactor wax, lower values of Sauter bubble diameters were obtained with SMP distributor compared to those from the 1.85 mm orifice plate distributor.

3. Effect of Wax Type

Figure 41 compares Sauter mean diameters for three different waxes: FT-300 wax, Sasol reactor wax, and Mobil reactor wax. Runs were made in the 0.051 m ID column using a 1.85 mm orifice plate distributor at 265 °C. The decrease in d_s for FT-300 as u_g increases from 0.01 to 0.02 m/s may be due to change in flow regime, from bubbly flow to the foamy regime. However, at high gas velocities, d_s is almost constant, about 1 mm. Whereas, d_s for reactor waxes increased as the gas velocity increased. Sauter mean diameters for Sasol wax increased from about 1 mm at 0.01 m/s to about 2 mm at high gas velocities, and those of Mobil wax increased from about 1.5 mm to about 5.5 mm at high gas velocities. The steady increase in d_s for reactor waxes as the gas velocity is increased is due to the absence of foam and increase in coalescence (concentration of large bubbles is increased).

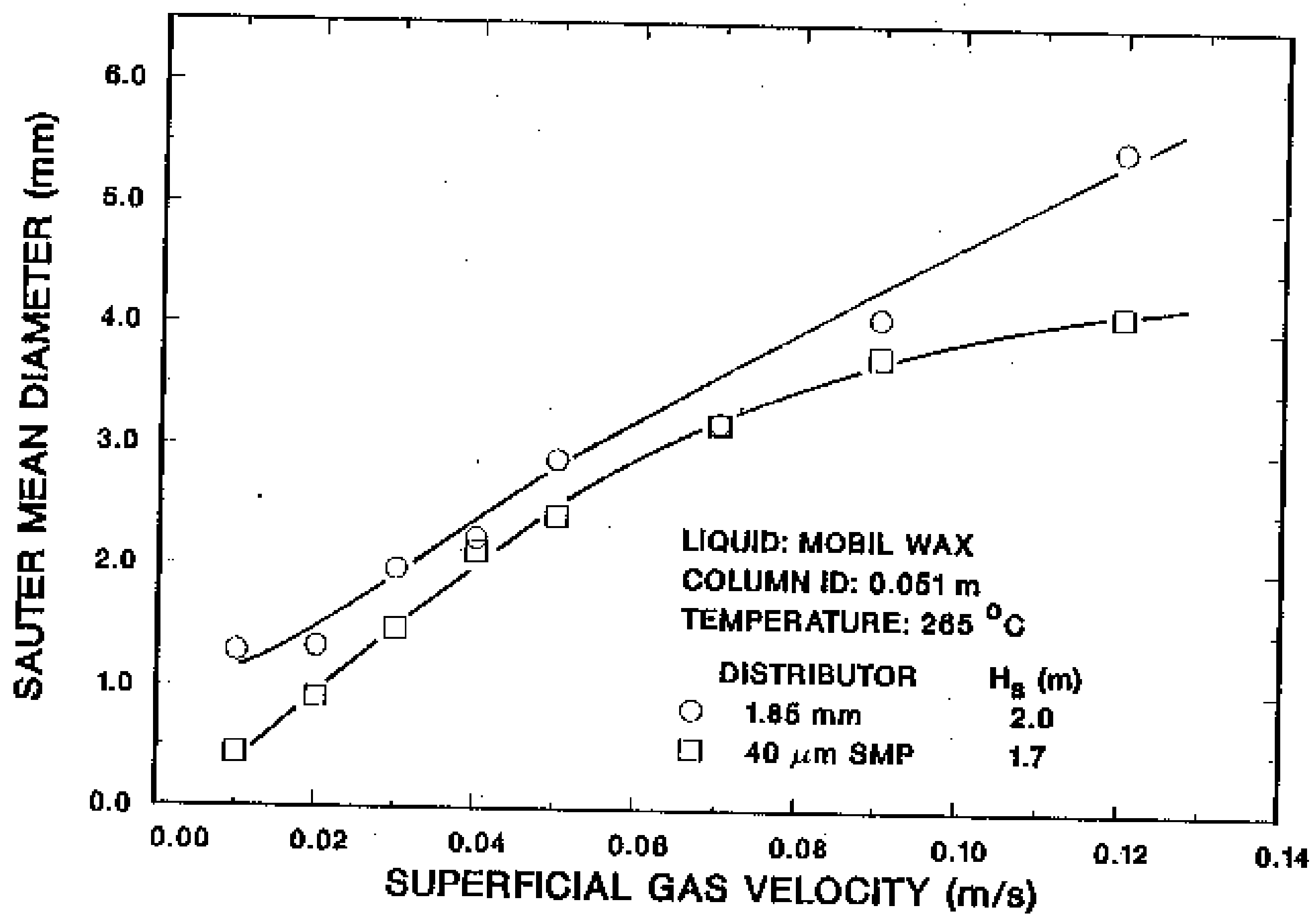

Figure 40. Effect of distributor type and superficial gas velocity on Sauter mean bubble diameter (○ -Run 9-3; □ -Run 9-4).

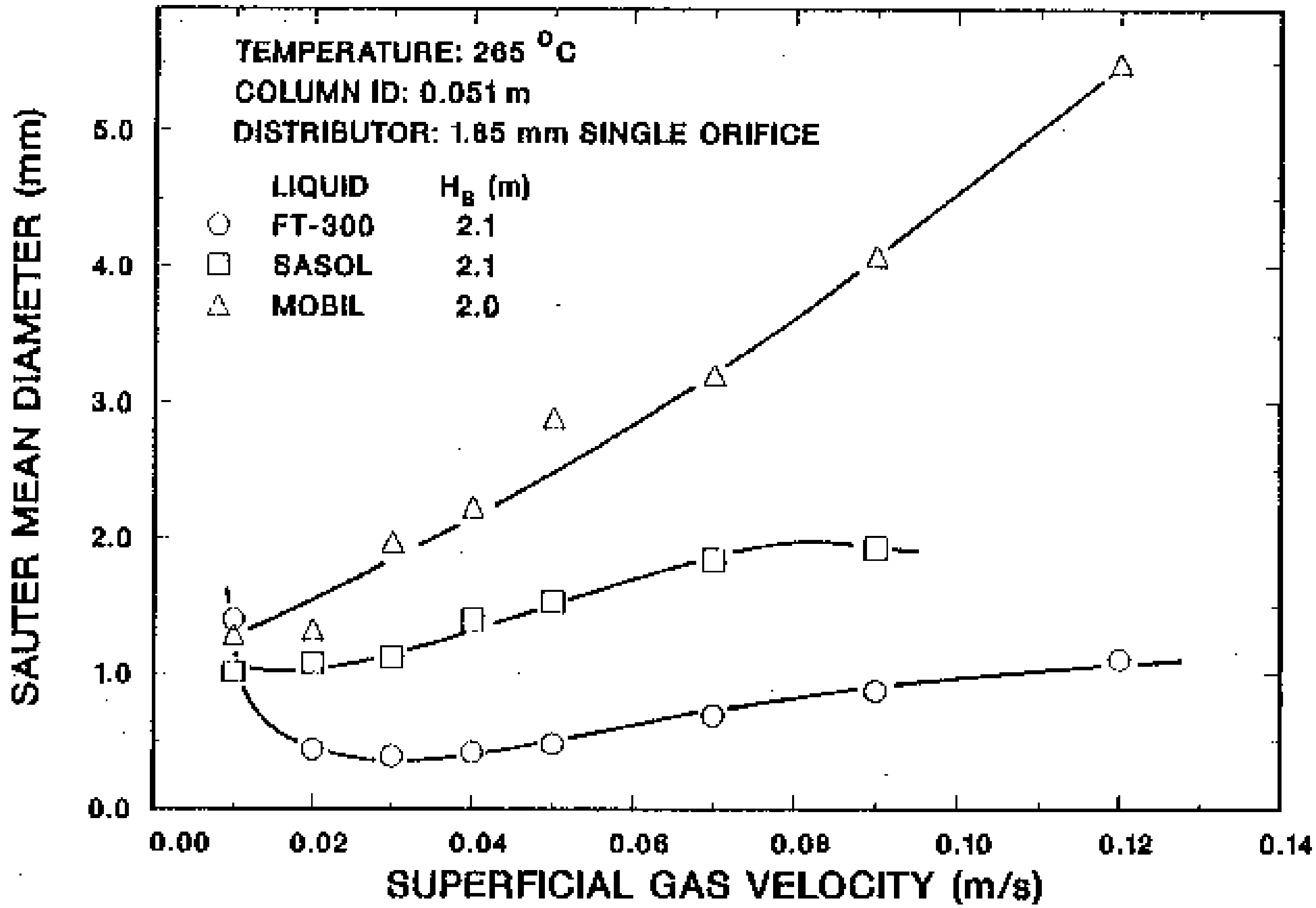

Figure 41. Effect of wax type and superficial gas velocity on Sauter mean bubble diameter (○ -Run 13-3; □ -Run 8-4; △ -Run 9-3).

B. Pure Liquids

1. Effect of Distributor Type

Figure 42 compares Sauter mean bubble diameters for n-butanol in the 0.051 m ID column (data for distilled water are not available). d_s for the 1.85 mm orifice distributor decreased from about 9 mm at 0.01 m/s to a constant value of about 6 mm at high superficial gas velocities (> 0.05 m/s). Whereas, for the SMP distributor d_s increased from about 5 mm at 0.01 m/s to a constant value of about 7 mm at high superficial gas velocities.

These results show that, in the fully developed slug flow regime, there is no significant effect of distributor type on d_s. Similar trend was observed with reactor waxes (e.g. Figure 40), where in the slug flow regime the d_s for the SMP and the 1.85 orifice plate were about 5 to 6 mm.

C. Aqueous Solutions of n-Butanol

1. Effect of Distributor Type

Figures 43 and 44 show Sauter mean bubble diameters for runs in 0.051 m ID column with 0.5 wt. % and 1.0 wt % butanol in distilled water (non-coalescing media) using different kinds of gas distributors: SMP, 1.0 mm orifice plate (not used for 0.5 wt % butanol solution), and 1.85 mm orifice plate. In both cases Sauter mean diameters decreased as the superficial gas velocity increased, the d_s values obtained from the 0.5 wt % butanol solution (see Figure 43) showed no significant effect of distributor type on d_s. The d_s values decreased from about 5 mm at 0.01 m/s to about 1 mm at at 0.12 m/s, whereas the d_s values obtained from the 1.0 wt % n-butanol solution decreased sharply at low superficial gas velocities (0.01

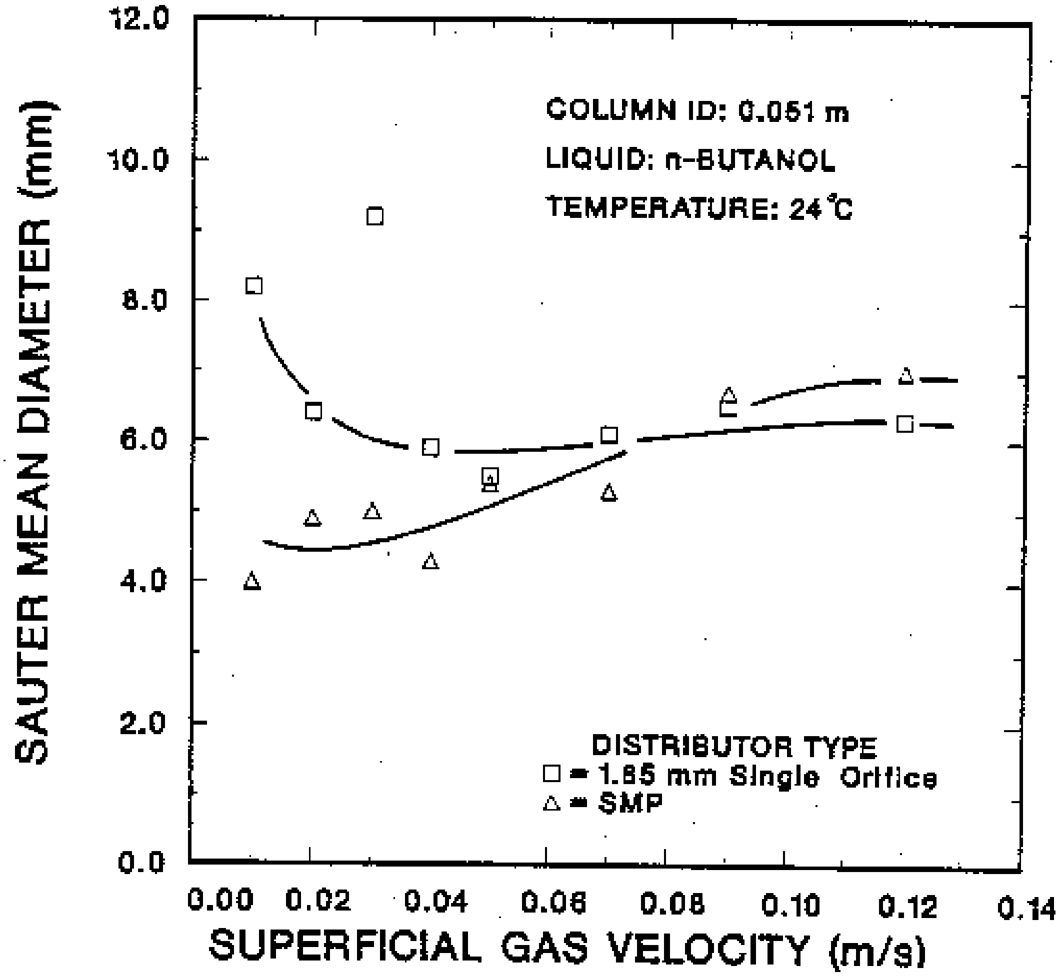

Figure 42. Effect of distributor type and superficial gas velocity on Sauter mean bubble diameter (□ -Run B-1; △ -Run B-2).

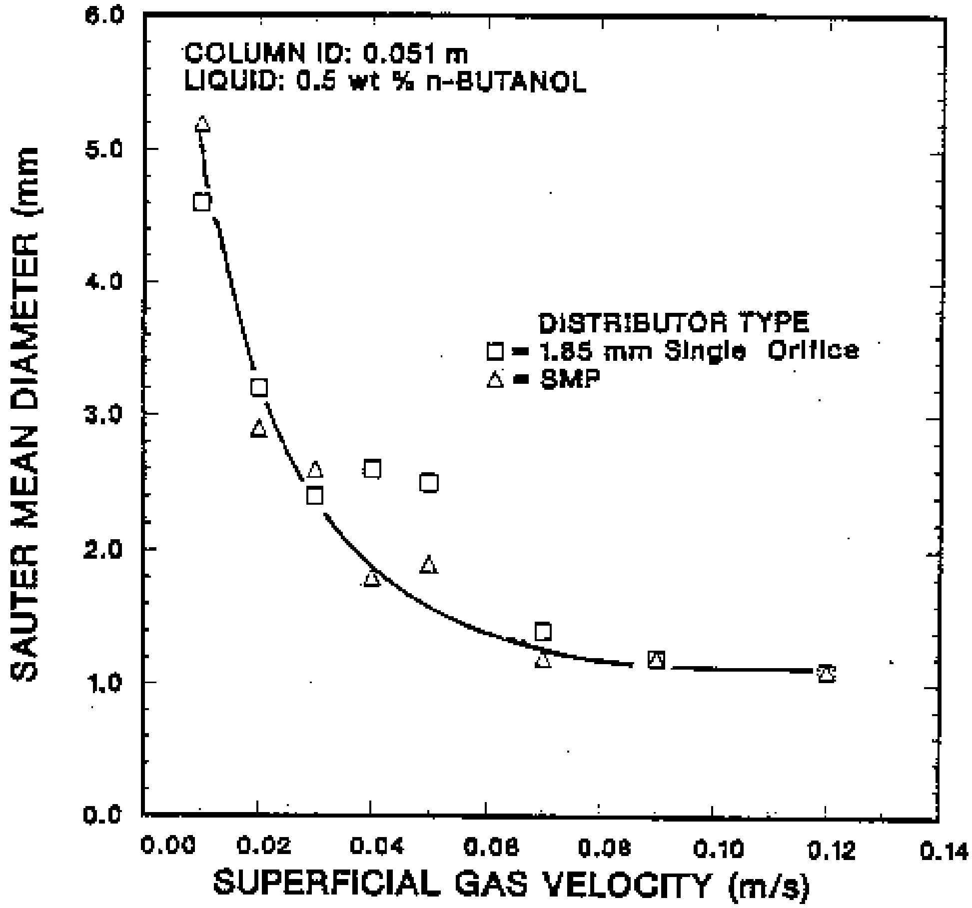

Figure 43. Effect of distributor type and superficial gas velocity on Sauter mean bubble diameter (□ -Run B-7, △ -Run B-8).

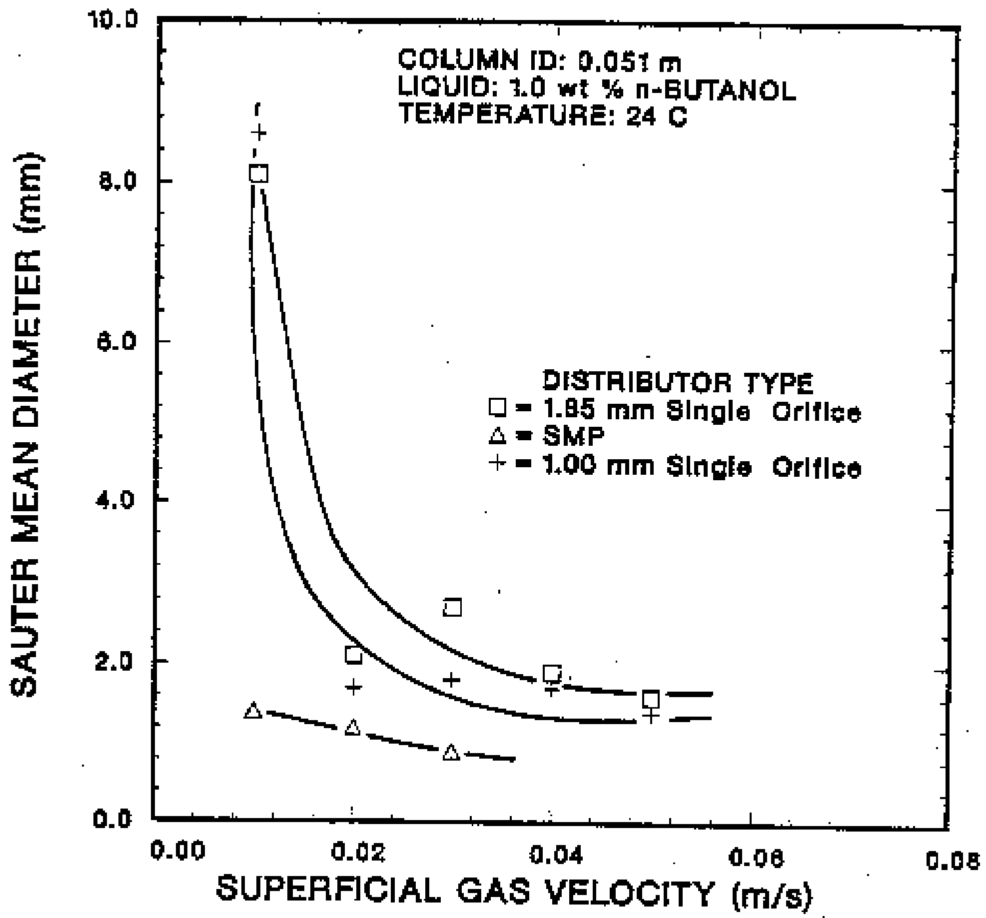

Figure 44. Effect of distributor type and superficial gas velocity on Sauter mean bubble diameter (□ -Run B-3, △ -Run B-5, + -Run B-6).

to 0.03 m/s) from about 8 mm to about 2 mm for the orifice plate distributors (see Figure 44), while the Sauter mean diameters for the SMP dropped a little, from 1.5 mm to about 1 mm. It was not possible to obtain Sauter mean diameters for superficial gas velocities higher than 0.05 m/s due to formation of foam which made the DGD method impossible to use. Orifice plates produced Sauter mean diameters which were higher than those produced by SMP. the difference was more pronounced in the bubbly flow regime. However, upon entering the transition to foamy regime (0.03–0.04 m/s) d_s values for all distributors were about 1 to 2 mm. These values of d_s (near the foamy regime) were higher than those observed with FT-300 wax (at 265 °C), where d_s in the foamy regime (0.02–0.04 m/s) were about 0.5 mm. This difference is probably due to different physical properties of each system.

D. Aqueous Solutions of n-Butanol With CMC

1. Effect of Distributor Type

Figure 45 shows Sauter mean diameters for a strong coalescing media (aqueous mixture of 1 wt % butanol and 0.5 wt % CMC) in the 0.051 m ID column using three different distributors: SMP, 1.0 mm orifice plate, and 1.85 mm orifice plate. At low gas velocities SMP distributor produced higher Sauter mean diameters than the orifice plates. The Sauter mean diameters for the SMP distributor were about 20 mm at 0.01 m/s; whereas, for the orifice plates Sauter mean diameters were about 18 mm and 15 mm for the 1 mm and 1.85 mm orifice plate distributors, respectively. At a superficial gas velocity of 0.02 m/s all distributors gave the same Sauter mean diameter of about 15 mm. At high superficial gas velocities (> 0.07 m/s) d_s values were fairly constant, the 1.85 mm orifice plate distributor produced

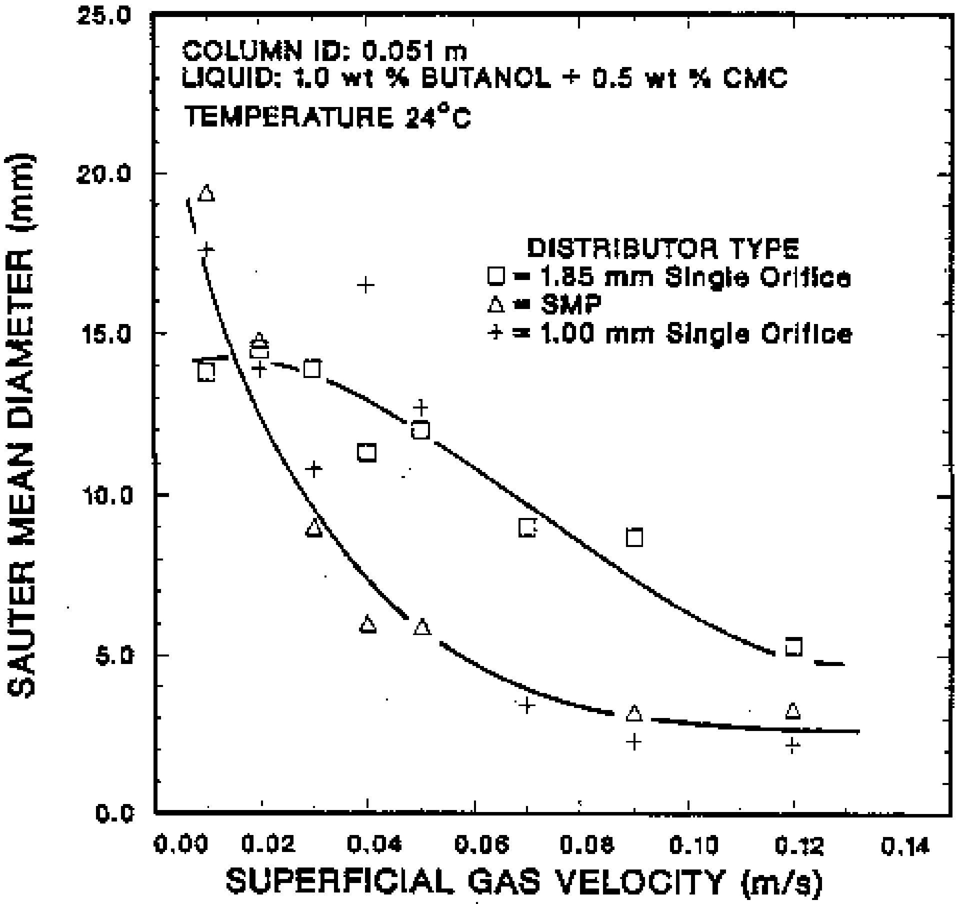

Figure 45. Effect of distributor type and superficial gas velocity on Sauter mean bubble diameter (□ -Run B-13, Δ -Run B-14, + -Run B-15).

Sauter values of about 5 mm, whereas, Sauter mean diameters for the SMP and the 1.0 mm orifice plate were about 3 mm.

The Sauter mean bubble diameters from the solutions containing CMC were higher than the d_s values obtained in a 1.0 wt % n-butanol solution (for the same type of distributor), the d_s values were about 3 mm for the viscous mixture and about 1 mm for the foamy mixture. These results illustrate that increasing the viscosity of the liquid renders the liquid coalescing, thus larger bubbles are formed in the process.

2. Effect of Column Diameter

Effect of column ID on Sauter mean diameters was studied using a mixture of 1.0 wt % butanol with 0.5 wt % CMC (strongly coalescing mixture). Runs were conducted in the 0.051 m ID column and the 0.229 m ID column using 1 mm and 1.85 mm holes diameter orifices, results of this investigation are presented in Figure 46. At low gas velocities (< 0.03 m/s) the d_s values from the small column were slightly higher than those from the large column. The d_s values in the small column were about 14 mm and 11 mm (at u_g=0.03 m/s) for the runs with 1.85mm orifice and 1.0 mm orifice plates, respectively. Whereas, the corresponding Sauter diameters in the large column were about 9 mm and 3 mm, respectively. This might be due to high rate of bubble coalescence in the small column (limited flow area, resulting in slug formation) than in the large column. At high gas velocity (u_g=0.07 m/s), no significant effect of column diameter was observed, the Sauter mean diameters for the 1.85 mm distributor were about 5 mm and about 2 mm for the 1.00 mm distributor.

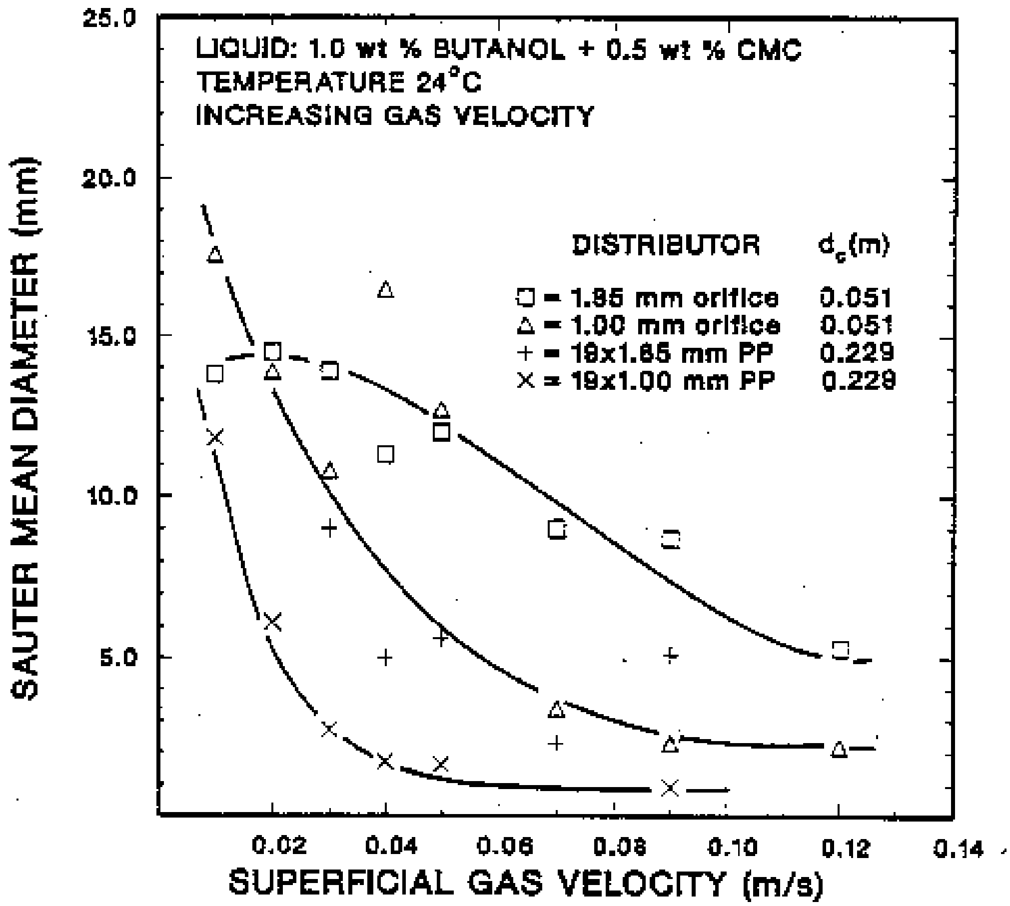

Figure 46. Effect of column diameter and superficial gas velocity on Sauter mean bubble diameter (□ -Run B-13, △ -Run B-15, + -Run B-20, × -Run B-21).

In summary, the effect of column diameter on d_s is more significant at low superficial gas velocity (bubbly flow regime), once in churn-turbulent flow regime or slug flow regime there is no significant effect of column diameter on d_s.

3. Effect of Physical Properties

Figure 47 shows the Sauter mean diameters for various aqueous solutions: 0.5 wt % butanol, 1.0 wt % butanol, and 1.0 wt % n-butanol with 0.5 wt % CMC, using 1.85 mm orifice plate distributor in the 0.051 m ID column.

The Sauter mean diameters increased with increase in liquid viscosity. That is, as the viscosity of the liquid increased, the more coalescencing liquid mixture became and larger bubbles were formed. At superficial gas velocity of about 0.03 m/s the foaming mixture had d_s of about 2 mm, whereas the highly viscous mixture had d_s of about 10 mm. Similarly, at high gas velocity (0.12 m/s) the foaming mixture had d_s of about 1 mm and the viscous mixture had d_s of about 7 mm.

In summary, This confirms that increasing the viscosity of the liquid results in increase in d_s. Similar observation were observed with FT derived waxes when the temperature was decreased from 265 °C to 200 °C higher values of d_s were observed (viscosity of wax increases with decrease in temperature).

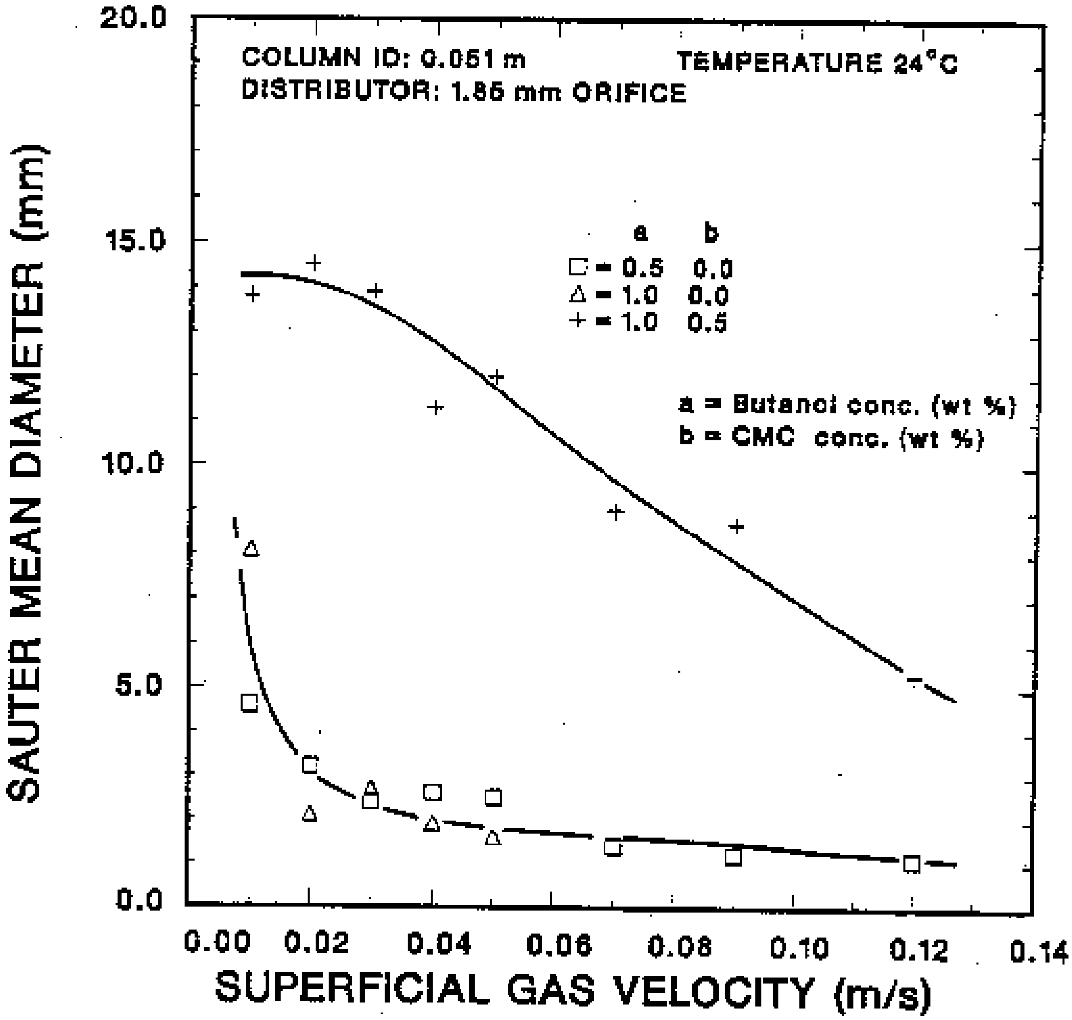

Figure 47. Effect of butanol and CMC concentration on Sauter mean bubble diameter (□ -Run B-3, △ -Run B-7, + -Run B-13).