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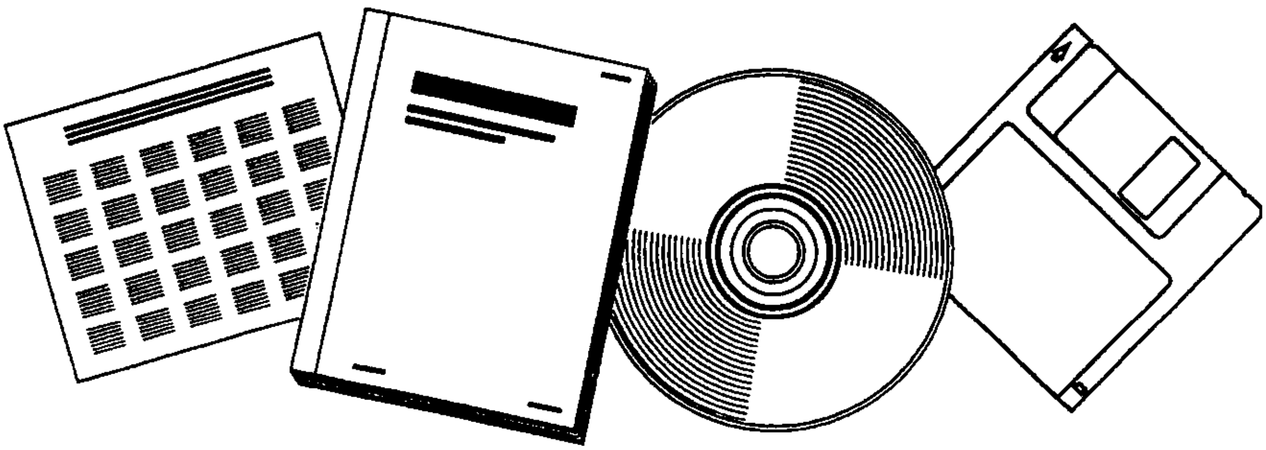

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Fluid Dynamic Behavior of Large Bubble Columns

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FLUID DYNAMIC BEHAVIOR OF LARGE BUBBLE COLUMNS

by

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ABSTRACT

A study was undertaken to investigate the fluid dynamics of large scale bubble columns. A literature search was conducted in the first phase of the study, and subsequent phases were to devise and implement an experimental program to test specific design concepts and to offer recommended design improvements for the SRC-II Demonstration Plant dissolver. The literature search included review and analysis of existing knowledge in the areas of flow regime, gas holdup and mass transfer, backmixing and solids accumulation. It was found that a vast amount of literature exists on bubble column behavior; however, most studies have dealt with small laboratory size columns, and only a few extended studies to larger sizes. This report details the findings of the literature search. The subsequent phases of the study were not undertaken due to termination of the SRC project.

FLUID DYNAMIC BEHAVIOR OF LARGE BUBBLE COLUMNS

TABLE OF CONTENTS

FLUID DYNAMIC BEHAVIOR OF LARGE BUBBLE COLUMNS

SUMMARY

There is a vast amount of literature dealing with the basic fluid dynamic behavior of multiphase flow in bubble columns. However, in most of the reported studies, either the column diameter was small or the length to diameter ratio was small. As the column size is increased, the gas, liquid, and flow patterns become more complex and it is questionable whether any of the simple models generally used to describe backmixing in small columns (such as a one-dimensional dispersion model) can adequately represent the complex flow behavior in large size columns. There is, therefore, a need to improve our understanding of the complex fluid dynamics of large scale bubble columns, and in recognition of this need, a new project was initiated in early 1981. In the first phase of this project, a detailed review and analysis of the existing knowledge in this area was undertaken. The subsequent phases of the project were not carried out because of the early termination of the project.

The various aspects of bubble column behavior, flow regime, gas holdup and mass transfer, backmixing and solids accumulation were investigated. Flow regime is the most critical consideration in evaluating the bubble column characteristics. The predominant flow regime in the large industrial columns is the churn-turbulent flow, characterized by intense churning or recirculation patterns in the liquid phase. The behavior of the bubble columns operated in this flow regime is complex and no simple and easy theoretical treatment for design and scaleup is available. The quiescent

bubble flow regime, most commonly encountered in the small laboratory columns; on the other hand, is easily represented by a simple one-dimensional dispersion model. Backmixing in these columns is characterized by a dispersion coefficient, the parameter in the axial dispersion model. There have been a number of studies relating this dispersion coefficient to system variables like flow rates, geometry of the column, and the fluid properties. It is common practice to use these same correlations to predict backmixing in large scale bubble columns. This approach suffers from two basic deficiencies, first, the extrapolation of these correlations developed by experimenting with small columns and second, the applicability and adequacy of the axial dispersion model to represent the complex flow behavior of large bubble columns. Limited work reported in the literature on the large bubble columns and work at GS&TC indicate that the literature correlations possibly overestimate backmixing in the large bubble columns. For confident scaleup, therefore, there is an urgent need to identify models which represent more closely physical phenomena in the large vessels.

In the large bubble column reactors, there is a strong potential for bulk gas/slurry separation (streaming or channeling) and reduced bubble surface area available for mass transfer at increased heights. Additional work is required to better understand the phenomena of bubble coalesence and gas/slurry separation leading to potentially reduced mass transfer rates as a function of column height.

There is a reasonable level of confidence in the existing correlations for predicting phase holdups in the large scale bubble column reactors.

INTRODUCTION

A number of investigations dealing with the basic fluid dynamics of two- and three-phase flow in bubble columns have been reported. However, almost in all published literature, either the column diameter was small or the column length-to-diameter ratio was small. As column size is increased, it is believed that the gas, liquid, and solid flow patterns in the column become more complex and it is questionable whether any of the simple models proposed to describe the bubble flow dynamics, based on observations in small columns, are applicable to large scale systems. A project was therefore initiated to address the development of a better understanding of the fluid dynamics of large scale bubble columns of the type which are used in the SRC-II coal liquefaction process. In the first phase of this project, a detailed review of the existing knowledge in this area of large scale bubble column design was undertaken and the results of this review are discussed in this report.

The review and analysis completed was to lead to the development of a detailed experimental program to test some specific design concepts and offer recommendations for improvements in the design of SRC-II Demonstration Plant dissolver. Some of the specific design concepts planned for testing were column internals, such as downcomer, draft tubes, gas distribution/ redistribution devices, multiple feed entries, solids withdrawal techniques, and gas addition at intermediate locations. Before work began on these later phases of work, the project was terminated.

DISCUSSION

1. General

It is generally recognized that there are inherent differences in the fluid dynamic behavior of small and large size bubble column reactors. One reason for this is the range of gas and slurry velocities at which the two scales of equipment are generally operated. The bubble flow dynamics are strong functions of the magnitudes of these velocities. The large size bubble column reactors for the SRC-II Demonstration Plant operate at high enough velocities compared to the various pilot plant reactors, that significant differences in the column hydrodynamics between the two scales of equipment should be expected. For a confident scaleup of the process, therefore, a better or more fundamental understanding of the basic fluid dynamic behavior of these large scale systems is required.

2. Fluid Flow Patterns

There is a vast amount of literature on bubble column behavior, in general. Most of the studies, however, deal with small laboratory size columns and only a limited few extended studies to larger sizes. In general terms, bubble column reactors are those in which a discontinuous gas phase in the form of bubbles move relative to a continuous phase. The continuous phase can be a liquid or a homogeneous slurry. In describing flow through these types of multiphase flow reactors, three flow regimes are generally recognized, namely quiescent bubble flow, churn or recirculation flow, and slug flow. Quiescent bubble flow regime is characterized by bubbles rising uniformly in the continuous slurry phase without any appreciable interaction. At some higher velocities, however, the bubble flow becomes nonuniform and unstable, marked by significant interaction between rising gas bubbles with intense recirculation patterns or 'churning' established in the slurry phase. At still higher gas velocities, the bubble interactions become very intense and the pseudo gas-in-liquid dispersion of bubble flow cannot be maintained. The result is a heterogeneous flow regime characterized by bubble coalescence and, as the large coalesced bubbles move with high rise velocities, unsteady flow and channeling. The movement of coalesced bubbles at high velocity produces turbulence in the slurry phase and when this turbulent field becomes sufficiently intense, bubble breakage occurs as a result of the local pressure fluctuations to which the coalesced bubble is subjected. Some of these small bubbles thus produced tend to coalesce and form large bubbles again. The turbulent breakup process can prevent this reagglomeration of bubbles only if the bubble size produced is small enough to cause the bubbles to remain spherical. Unless the intensity of turbulence is sufficient to produce a bubble size less than a critical value where the bubble shape is spherical, some of the smaller bubbles produced by turbulence in the slurry phase coalesce to form larger bubbles again and an equilibrium size distribution will eventually be reached for low L/D tank-type reactors and mass contactors.

Since the coalescence and breakage is a random process, some of the coalesced bubbles do escape the turbulent field and do not undergo breakage. With increased column heights, the number of these bubble clusters or agglomerates that go unaffected by the turbulence in the slurry phase increase, and lead to streaming or channeling in the upper part of the column. As a result, more and more gas is trapped in this large bubble phase or bubble clusters and the interfacial area for mass transfer of hydrogen to the slurry is drastically reduced.

The reduction in the bubble surface area at increased heights of the column, discussed in the previous paragraphs is only relevant to large scale bubble column reactors, and therefore cannot be characterized or even identified by experimenting with short columns.

Considering the quiescent bubble flow regime, the slurry phase moves through the column in plug flow with a small degree of backmixing caused by the presence of gas and the gas phase - uniformly distributed as bubbles - flows through in a near perfect plug flow. The fluid dynamic behavior of the bubble column reactor operating in this flow regime can therefore be reasonably described by a simple dispersion model for the slurry phase and plug flow model for the distributed gas phase. Outside the quiescent bubble flow, however, a recirculation or churn flow model is more appropriate. In this model, intense recirculation patterns or churning in the liquid phase are visualized. Each circulation pattern is known as a cell and visual observations indicate the presence of multiple circulation cells in the axial direction. A prerequisite for the establishment of a recirculation is the existence of the cross-sectional nonuniformity of gas holdup. Because of this nonuniformity, a density difference is established between those parts which are either rich or poor in the dispersed phase and circulation is initiated. As a result, a radial slurry velocity gradient is established. The existence of the radial gas holdup and slurry velocity distribution has been confirmed by experimentation.[1,2,3] The one-dimensional dispersion model, on the other hand, does not account for or recognize any radial variations in gas holdup or velocities. The churn-turbulent flow regime is the flow regime most commonly encountered in large industrial bubble columns.[4] A number of investigators attempted to model this flow regime. Most recently, Joshi and Sharma[5] provided a sound theoretical model based on the energy balance approach of Davidson.[6]

As can be inferred from the foregoing discussion, it is the flow regime definition which is the most important consideration in evaluating the characteristics of these types of multiphase flow reactors. The criteria for transition from one flow regime to another are not easily defined, but are expected to depend on column dimensions in addition to flow rates and fluid properties. In a 15 cm column, Braulick[7] observed transition from quiescent bubble flow at a superficial gas velocity of 2.7 cm/sec. and slugging was observed at a velocity of 6.1 cm/sec. Deckwer,[8] in his work on a bubble column of 20 cm, did not observe any significant bubble coalescence or slug flow until a gas velocity of 6 cm/sec. is reached. Slug flow itself is important only for small diameter columns, perhaps no larger than 20 cm. Darton and Harrison[9] proposed a flow regime chart based on the slip velocity, U_S, defined as

$$U_S = \frac{U_G}{E_g} - \frac{U_L}{1-E_g} \qquad (1)$$

where,

U_L, U_G = superficial velocities of liquid and gas

E_g = gas holdup

and the drift flux of gas, W_{CD}, which is defined by Wallis[10] as the volumetric flux of gas relative to a surface moving at the average velocity, i.e.

$$W_{CD} = U_S \, E_g \, (1 - E_g) \tag{2}$$

If the drift flux is plotted as a function of gas holdup, the change in slope indicates the transition between quiescent bubble flow and the heterogeneous churn-turbulent flow.

In the discussion thus far, no reference was made to the presence of solids. What effect they have on the flow regimes in the gas-liquid-solid operation as in SRC-II reactors is unknown.

3. Gas Holdup and Mass Transfer

Gas holdup is a key variable in evaluating the performance of a bubble column reactor. Apart from influencing the slurry residence time in the bubble column, it affects the interfacial mass transfer rate. There have been numerous studies attempting to characterize the effects of several particle and fluid parameters on gas holdup. Earlier work in this area was reviewed by Ostergaard.[11] A considerable number of additional studies were reported since that time.

In three-phase flow, the particle and flow parameters directly or indirectly affecting the gas holdup are: (1) superficial velocities of gas and liquid phases, (2) their physical properties, (3) particle size distribution and their average density, (4) concentration of solids in the inlet stream, and (5) diameter of the column. Several investigators, Imafuku, et al.,[12] working with a batch system with particles in the range of 60 to 180μ; Kato, et al.,[13] with a continuous flow system with particles in the same general range; Viswanathan, et al.,[14] Ostergaard and Michelsen,[15] among others, concluded that, in general, gas holdup in a fluid particle system is smaller than in a solid-free system under otherwise identical flow conditions. Their results, particularly for bed of particles under 1 mm diameter, are in substantial agreement that the presence of solid particles tend to decrease the gas holdup. This effect of solids on gas holdup in three-phase fluidized beds is considered to be the result of significant bubble coalescence that takes place in the presence of small particles and the consequent larger rising velocities of the coalesced bubbles. As the rising velocity of the gas bubbles increases, less time is spent by the gas in the fluidized bed. The effect of solids larger than 1 mm on gas holdup was found to be less significant, since these particles tend to cause breakup of gas bubbles. The effect of concentration of solids of a given size on gas holdup was investigated by Kato, et al.[13] This work showed that the increasing solids concentration generally decreases gas holdup but the effect becomes insignificant at high gas velocities in the range of 10 to 20 cm/sec.

The effect of gas and liquid velocities on gas holdup was extensively studied. However, in most cases, the solids considered were large (over 1 mm) and the terminal settling velocities were many times larger than the upward velocity of the fluidizing medium which, in most cases, was water. In general, increasing gas velocities increased gas holdup as shown by the investigations of Michelsen and Ostergaard,[16] Viswanathan, et al.,[14] Dakshinamurty, et al.,[17] Kim, et al.,[18,19] Kato, et al.,[13] and a number of other investigators.

The effect of liquid velocity on the gas holdup appears to depend upon the free settling velocities of the particulate phase. Sherard[20] observed that in the beds of large and heavy particles (large terminal settling velocities), the holdup decreased with liquid velocity whereas in the beds of small and/or light particles, gas holdup was essentially independent of liquid velocity. The work of Ostergaard and Michelsen,[15] using particles in 6 mm diameter range, showed that increasing liquid velocities decreased gas holdup, supporting the findings of Sherard.[20] The results of Michelsen and Ostergaard[16] and Kim, et al.,[18,19] however, differed in that they showed that the gas holdup increased with liquid velocity. Ostergaard and Gilliland[21] measured the gas holdup in beds of sand particles of 40 to 80 mesh. The fluid media were nitrogen and water. The gas holdup was largely independent of particle size or liquid velocity confirming the findings of Sherard[20] about the lack of correlation between the gas holdup and liquid velocity for small micron size particles.

Kim, et al.[18,20] extensively studied the effect of liquid properties - density, viscosity, and surface tension - on phase holdups in three phase fluidized systems. Solutions of water, sugar, and CMC were used to simulate different liquid phase properties. Their data on the effect of liquid properties on gas holdup can be approximated as:

$$\varepsilon_g \propto \left(\frac{1}{\sigma}\right)^{0.1} \qquad (3)$$

The effect of liquid density and viscosity are not considered, since the magnitude of their effects on gas holdup is negligibly small. Earlier work by Akita and Yoshida[22] for two-phase flow indicated a very similar dependence of gas holdup on liquid surface tension.

Kim, et al.[18,19] correlated their data on bed porosity and liquid holdup in terms of gas and liquid Froude numbers and liquid Reynolds number through a least squares analysis. Bed porosity is defined as the combined liquid and gas holdup. Gas holdup is obtained by subtracting liquid holdup from the combined liquid and gas holdup. There are several other correlations available for estimating bed porosity. Some of these are as offered by Dakshinamurty, et al.,[17] Ostergaard,[21] Darton and Harrison,[23] Ostergaard and Theissen,[24] and Soung.[25] However, all these correlations contain a term, U_{OL}/U_t, the ratio of superficial liquid velocity to particle settling velocity, and unless this term is much less than unity, none of these correlations give meaningful estimates for bed porosity.

The best gas holdup data for three-phase flow systems were reported by Kato, et al.[13] In their study, they measured the longitudinal concentration distribution of solid particles and gas holdup in 6.6, 12.2, and 21.4 cm diameter columns. The particles ranged in size from 68 to 177μ, the liquid velocity from 0.5 to 1.5 cm/sec, and gas velocity from 2 to 25 cm/sec.

These data reported by the investigators were smoothed and correlated by the following equation.

$$E_g = 0.092 \ln U_{OG} - 0.024 \qquad (4)$$

Equation (4) was arrived at by correlating data developed with air and water as fluidizing media. The effect of liquid properties on gas holdup was inferred from the work of Kim and Akita and Yoshida. By incorporating these property effects, the following equation results:

$$E_g = (\sigma_L)^{-0.1} \left(0.14 \ln U_{OG} \; -0.04\right) \tag{5}$$

Regarding the effect of the diameter of vessel on gas holdup, it has been shown that under identical conditions, larger columns result in lower gas holdup. This has been explained on the basis that in the larger column, random circulation patterns (eddies) exist and the gas is actually rising in regions where liquid is also rising. Smaller columns exhibit relatively higher holdups because of the absence of these eddies. Based on the work of Kato, et al.[13] and Reith, Renken, and Israel,[26] this effect of column diameter on gas holdup has been determined to exhibit this form:

$$E_g \propto D^{-0.08} \tag{6}$$

The effect of column diameter on gas holdup is derived by extrapolating the available data in the 12 to 29 cm range. Whether this holds true to very large sizes is questionable. Akita and Yoshida[22] suggest that the diameter effect may vanish above a diameter of 61 cm. Recent work of Kataoka, et al.[27] on a large 5.5 m diameter column appear to confirm this. Ying, et al.[28] recently reported gas holdup data from a comparatively large scale column (30 cm X 600 cm). They claimed that the two-phase correlation of Akita and Yoshida adequately represented their data. The same conclusion was reached by our own work at Gulf Science & Technology Company on a 30 cm X 762 cm column. For two-phase flow, one of the best known correlations was that of Akita and Yoshida.[22] They used dimensional analysis approach to correlate their data taken on a 61 cm column. They found that the liquid superficial velocity, gas density, and the orifice diameter for gas distribution at the inlet to the column have an insignificant effect on gas holdup. The effect of gas distribution at the inlet is significant only for small sized columns.

Interfacial area available for mass transfer in a bubble column is an important design variable and is related to gas holdup and mean bubble diameter. Akita and Yoshida,[22] using a dimensional analysis, developed a correlation for the average bubble size in the column. Other correlations for the interfacial area and bubble diameter can be found in the literature in the works of Reith,[29] Kastenek,[30] Schugerl,[31] and others.

Several correlations are available to predict slurry phase mass transfer coefficients; i.e., Calderbank and Moo-Young,[32] Akita and Yoshida,[22] Kataoka and Miyauchi,[33] Tadaki and Maeda,[34] etc. Kataoka, et al.,[35] measured values of liquid phase mass transfer coefficients in a large 5.5 m diameter column. After comparing their data with the different existing correlations, they concluded that the mass transfer coefficients are not affected by the scale of the column, and the correlation of Akita and Yoshida represented their data with reasonable accuracy.

4. Backmixing

The extent of backmixing in two- and three-phase flow, particularly in two-phase, gas-liquid upflow has been the subject of numerous investigations in the past decade. Shah, et al.[36] provided an extensive review of these investigations. The degree of axial mixing present in the column plays a very important role in the design and scaleup of SRC-II reactors. It is therefore important to be able to properly and adequately characterize this aspect of bubble column behavior, i.e., axial dispersion or mixing in the column as a function of column geometry, flow rates, and fluid properties. Any fluid dynamic model used for this purpose has to take into account the flow complexity that exists in the large bubble column reactors. Generally, however, the approach has been to assume that either a simple axial dispersion model or tanks-in-series model adequately describes the mixing patterns in the slurry phase of the vessel. This approach is unlikely to give good estimates of backmixing in the large sized column and therefore for scaleup purposes, higher order models with flow bypass, recirculation or backflow or stagnant regions or network models where several flow units are connected in parallel or in series may have to be used. The multiple cell model of Joshi and Sharma[5] is a step in the right direction. In their model, a number of mixing cells in the axial direction are visualized. The overall mixing is then characterized by the number of cells and the extent of circulation. The existing theory predicts the number of cells, but not the extent of interaction between the cells.

In the one-dimensional axial dispersion model, dispersion coefficient is the parameter which describes the extent of backmixing. Estimates for these coefficients are obtained from tracer tests. The following correlation proposed by Deckwer, et al.[37] is one of the most commonly used correlations for two-phase gas-liquid flow.

$$E_Z = 2.7 \quad D_t^{1.4} \quad U_{OG}^{0.3} \tag{7}$$

where

E_Z = dispersion coefficient, cm^2/sec

D_t = column diameter, cm

U_{OG} = superficial gas velocity, cm/sec

As can be seen, this correlation does not include physical property effects, if any, on the dispersion coefficient. Cova[38] investigated the effect of fluid properties of both the gas and liquid and concluded that liquid viscosity and surface tension do not have a significant effect on the dispersion coefficient and even the effect of liquid density is only minor, and that it is important only for flow in small tubes.

The effect of suspended particles on axial dispersion in the liquid phase was studied by several investigators: Michelsen and Ostergaard,[16] Ostergaard,[11] Kim, et al.,[18,19] Imafuku, et al.,[39] Farkas and Lebland,[40] among others. Kato, et al.[13] correlated the longitudinal dispersion coefficient for the slurry phase by the following relationship:

$$E_Z = 2.41\ D_t^{1.5} \left(1 + 0.43 \frac{U_{OG}^{0.85}}{D_t^{0.43}} \right) \tag{8}$$

Since the assumption that a simple dispersion model can adequately describe the complex mixing patterns in the vessel is implicit in the above correlations, problems arise when such an assumption is invalid as when the column is operated outside the bubble flow regime. As stated earlier, this is generally the case with the large sized industrial bubble columns. As a result, it is not reasonable to expect any of the existing two- or three-phase correlations can adequately predict backmixing in the industrial (large size) bubble columns. Recent work of Field(41) experimenting with a large 3.2 m diameter column confirmed the inadequacy of this approach to predict axial mixing in a bubble column. In their experiment, the measured dispersion coefficient was 0.22 m^2/sec compared to a predicted value of 3.6 m^2/sec using Deckwer's correlation. More recently, the tracer experiments conducted on Ft. Lewis' reactor and on the cold flow model at GS&TC substantiated this concern about the generally used approach of assuming the applicability of an axial dispersion model to describe complex fluid flow patterns in the large scale bubble column reactors and using any of the existing correlations to predict axial mixing in such large scale systems. This concern points to an urgent need for identifying models which would represent more closely physical phenomena in SRC reactors.

5. Solids Accumulation

In the SRC-II reactor design and scaleup, it is important to know how the solids accumulate or disperse relatively in pilot plant and commercial size reactors. Accumulated solids, apart from affecting the slurry residence in the reactor, also affect the reaction kinetics by the mineral matter contained in these accumulated solids. The variables influencing the solids accumulation are the gas and liquid velocities, particle size, density, and the degree of backmixing in the slurry phase. The dissolver solids sampling program at Ft. Lewis gave a representative size distribution of solids in the dissolver. The median size was found to be 3μ and the average specific gravity was measured to be 2.3. The dissolver solids sampling program revealed no significant solids holdup gradient in the dissolver. Cova's model[42] was successfully used in simulating the observed solids holdup profile. Further confidence in the method was obtained by the recent cold flow model experiment on a 30 cm X 762 cm column at Gulf Science & Technology Company.

CONCLUSIONS AND RECOMMENDATIONS

Flow regime is the most critical consideration in evaluating the bubble column characteristics. Large industrial bubble columns operate in churn-turbulent flow regime, characterized by intense churning or recirculation currents established in the liquid or slurry phase. The fluid dynamic behavior of these columns is complex and no simple or easy theoretical treatment is available at this time for design and scaleup. Empirical extrapolations of information available on small scale laboratory columns is the only recourse at the present time. The predominant flow regime in these columns is the bubble flow regime. The fluid flow behavior of columns operating in this regime can easily be represented by a one-dimensional dispersion model. Axial dispersion in these columns is adequately characterized by a dispersion coefficient, the parameter in the axial dispersion model. There have been numerous studies relating the dispersion coefficient to variables of the system like column geometry, flow rates, and fluid properties. It is common practice in bubble column design to use these correlations to predict backmixing in large scale bubble columns. In the first place, these are not tested for applicability beyond the range of experimental conditions used for their development. Secondly, the axial dispersion model is inadequate to represent the complex fluid flow patterns of large scale columns. The fluid flow in these columns is characterized by the presence of radial gas holdup, velocity variations, and the presence of radial velocity components. Further, the gas phase is much more backmixed. Also, with the presence of stagnant pockets, large residence times of slurry in the vessel are encountered. These flow complexities cannot reasonably be represented by any simple model. Higher order models involving flow bypass, backflow, or network models are required to properly represent the flow complexities of the large bubble columns.

Limited information was reported in the literature on the large bubble columns. The results from these reported studies and some of the work done at GS&TC on the 25 ft. plexiglass column indicate that the literature correlations, basically developed on small scale laboratory columns, do not accurately predict backmixing in the large columns. It appears that they overestimate backmixing by several orders of magnitude. For confident scaleup, therefore, additional work must be done. The first effort should be to identify and devise fluid dynamic models that better represent the physical flow phenomena in the large bubble column reactors.

In the large bubble column reactors, there is a strong tendency for potential gas/slurry separation or streaming at increased heights, leading to reduced bubble surface area for mass transfer. This can only be avoided by limiting reactor height or by gas redistribution along the column length. Future work in this area should be directed to evaluating this concern for reduced mass transfer at increased heights and ways to alleviate this potential problem. Specifically, bubble interactions, leading to coalescence, breakup or aggregate formation tendencies are to be examined as a function of column depth. One of the ways suggested to redistribute the gas is by introducing gas at appropriate locations along the column length. The feasibility of this approach needs to be examined.

One of the process concerns in the design of the Demonstration Plant dissolver was the potential for inadequate thermal and mass dispersion in the large dissolvers. Without adequate thermal dispersion, heat generated in the upper part of the column would not be available to raise the temperature of the incoming feed to the reaction threshold temperature - requiring a larger slurry preheater. Downcomers were offered as possible means of inducing increased slurry circulation within the dissolver. Their design and effectiveness should be further studied before adopting them in any dissolver design.

There is a reasonable level of confidence in the existing correlations for predicting phase holdups in the large scale bubble column reactors.

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