MASTER

OAK RIDGE NATIONAL LABORATORY

OPERATED BY
UNION CARBIDE CORPORATION
NUCLEAR DIVISION

POST OFFICE BOX 8
OAK RIDGE, TENNESSEE 37830

ORNL/MIT-272

DATE: April 26, 1978

SUBJECT: Mass Transfer in Three-Phase Fluidized Beds

Authors: B.W. Wu, Y.L. Cheng, J.R. Perini, and J.L. Roux-Buisson

Consultants: J.M. Begovich, S.D. Clinton, and J.S. Watson

NOTICE

This report was prepared as an account of work sponsored by the United States Government. Neither the United States nor the United States Department of Energy, nor any of their employees, nor any of their contractors, subcontractors, or their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights.

ABSTRACT

The effects of superficial liquid and gas velocity, particle diameter, liquid viscosity, and column diameter on liquid dispersion (E_L) and mass transfer (K_La) in three-phase fluidized beds were investigated using a water-glycerol/oxygen-nitrogen (or oxygen-argon)/glass-bead system. Overall mass transfer coefficients were calculated based on plug flow, dispersed plug flow, and continuously stirred tank models. K_La was found to increase with gas velocity and particle diameter, but no correlation of K_La with liquid velocity was observed. At low liquid velocities, K_La was lower for the more viscous liquid; the reverse was true at high liquid flow rates. E_L increased rapidly for liquid flow rates at two to three times the minimum fluidization velocity.

DISTRIBUTION STATEMENT

GROUP 1 UNLIMITED

Oak Ridge Station
School of Chemical Engineering Practice
Massachusetts Institute of Technology
S.M. Senkan, Director

Contents

	<u>Page</u>
1. Summary	4
2. Introduction	4
2.1 Background	4
2.2 Objectives	6
3. Apparatus and Procedure	6
4. Results and Discussion	10
4.1 Overall Mass Transfer Coefficient	10
4.2 Axial Dispersion Coefficient	17
5. Conclusions	17
6. Recommendations	17
6.1 Further Studies	17
6.2 Equipment Modifications	20
7. Acknowledgments	20
8. Appendix	21
8.1 Determination of Minimum Liquid Fluidization Velocity, $U_{L,MF}$	21
8.2 Summary of Experimental Data and Calculated Results	21
8.3 Location of Original Data	21
8.4 Nomenclature	21
8.5 Literature References	26

1. SUMMARY

Mass transfer experiments were conducted with two cocurrent three-phase fluidized beds. The liquid phase (aqueous glycerol) was saturated with oxygen prior to desorption by the inert gas phase (nitrogen or argon) inside the fluidized beds. Axial oxygen concentration profiles of the liquid were measured with oxygen analyzers. These profiles were used to calculate dispersion (E_L) and overall mass transfer coefficients (K_La).

Three- and six-inch diameter plexiglass columns were studied, using three glass-bead diameters (3.2, 4.6, and 6.0 mm) and two liquid viscosities (2.4 and 0.9 cp) at various liquid and gas velocities.

The overall mass transfer coefficients increased with both increasing bead diameter and superficial gas velocity. No relationship between liquid velocity and K_La was observed. K_La for the more viscous (2.4 cp) solution, however, was lower than K_La for the 0.9-cp liquid at low liquid velocities, with the reverse being observed at high liquid velocities. The liquid-phase axial-dispersion coefficient also increased with increasing column diameter and was found to increase sharply at the liquid velocity which was 2.5 times the minimum fluidization velocity.

2. INTRODUCTION

2.1 Background

Three-phase fluidized beds consisting of solid particles fluidized by co-current or countercurrent flow of gas and liquid phases are finding increased utility in industrial processes. Their principal applications exist in systems where gaseous and liquid components react on solid catalytic surfaces, especially where temperature control is important. Typical processes include the hydrogenation of liquid petroleum fractions, hydrogenation of unsaturated fats, Fischer-Tropsch processes, coal liquefaction, and slurry methanation of CO. Because of the versatility of the process, the Chemical Technology Division of ORNL began a study of the hydrodynamic and mass transfer characteristics of three-phase fluidized beds in 1974.

Details of relevant studies by several authors are summarized in Table 1. The gas-liquid mass transfer was found to be liquid-phase-controlled in many cases for the three-phase fluidization process (10). Volumetric mass transfer coefficients were found to increase with increasing particle size for a CO₂ absorption system (11). Dispersion coefficients - based on an axially dispersed plug flow model - were reported to increase with increasing gas velocities, but were not significantly affected by liquid-phase velocities. In similar studies in fluidized beds with particle sizes ranging from 0.22 to 0.80 mm, absorption rates increased with increasing liquid velocity and decreased with increasing particle size (9). Begovich (2) reported that the mass transfer coefficient increased with

Table 1. Summary of Previous Work

Reference No.	Gas Phase	Liquid Phase	Solid Phase	Column ID (in.)	Liquid-Phase Model	Results/Remarks
9	CO ₂ -N ₂	H ₂ O	0.22 mm silica 0.8 to 1.1 mm glass 0.26 mm sand	3.4 x 2.4 (rectangular)	-	Presence of solid particles decreased interfacial area.
11	CO ₂ -N ₂	H ₂ O	1 and 6 mm glass ballotini	4	PF	1) K_L increased with U_G and was unaffected by U_L . 2) K_{La} (6 mm glass ballotini) > K_{La} (bubble column) > K_{La} (1 mm glass ballotini).
5	CO ₂	glycerol-H ₂ O	-	1.5	DPF	1) K_{La} increased with U_G . 2) K_{La} decreased with U_L .
14	CO ₂	glycerol-H ₂ O	-	1.5 and 3	PF, DPF, and CST	1) K_{La} increased with U_G . 2) K_{La} decreased with U_L .
15	CO ₂ -Air	H ₂ O	3-4 mm alumina beads	3 and 6	PF, DPF, and CST	1) $U_{L,MF}$ decreased with U_G . 2) ϵ_s decreased with U_G and U_L and showed slope change at incipient fluidization.
3	CO ₂ -Air	H ₂ O	6.4 mm plexiglass and 3.4 mm alumina beads	3 and 6	PF, DPF, and CST	1) Alumina beads gave close to plug flow concentration profile. 2) Plexiglass beads gave close to CST. 3) K_{La} (alumina beads) > K_{La} (bubble column) > K_{La} (plexiglass beads)
7	Air	H ₂ O	3-4 mm and 1-2 mm alumina beads and 6.4 mm plexiglass beads	3 and 6	-	Correlation for ϵ_s was presented.
1	Air	glycerol-H ₂ O	4.6 mm glass beads	3	-	Correlation for ϵ_s was presented.
2	Air	H ₂ O	3, 4.6, and 6 mm glass beads and 6 mm plexiglass beads	3	PF, DPF, and CST	1) K_{La} increased with D_p and ρ_s . 2) K_{La} increased with U_G and U_L .

increasing particle diameter and density, as well as with increasing liquid- and gas-phase velocities. Axial dispersion of the gas phase was reported to be minimal in all studies.

2.2 Objectives

The objectives of the present study were to determine the effects of liquid-phase velocity and viscosity, gas velocity, particle size, and column diameter on mass transfer and hydrodynamic characteristics of three-phase fluidized beds. The method involved the determination of E_L and K_La by matching experimentally measured axial oxygen profiles in the liquid phase with model predictions based on plug-flow and axially dispersed plug-flow models.

3. APPARATUS AND PROCEDURE

The experimental apparatus is shown in Fig. 1. Inert gas (N_2 or Ar) and aqueous glycerol (or water) passed through fluid distributors at the bottom of either the 3- or 6-in.-ID column and flowed cocurrently upward to fluidize the glass beads. Oxygen concentrations and pressures were recorded at various heights in the bed. The liquid leaving the column was saturated with oxygen before it was returned to the feed tanks; however, a steady-state level of oxygen in the liquid-phase feed was difficult to obtain when the 6-in. column was used.

Oxygen profiles for each run were measured by four Instrumentation Laboratory oxygen analyzers connected to sample ports along the column, as shown in Fig. 2. These probes generated an electrical potential directly proportional to the oxygen concentration inside the sample cell. The probes were calibrated by circulating only the liquid phase through the column. Bubble entrainment through the sampling cells was minimized with a venting system.

The experimental conditions investigated are summarized in Table 2. Liquid and gas flow rates were calibrated according to procedures outlined by Reber *et al.* (14). Superficial liquid velocities in Table 2 were expressed as multiples of the minimum fluidization liquid velocity ($U_{L,MF}$) for each gas velocity, particle size, liquid viscosity, and column diameter. See Appendix 8.1 for the determination of $U_{L,MF}$. Experimental results are tabulated in Appendix 8.2.

The mathematical model was developed assuming a steady-state operation, gas phase in the plug-flow regime, liquid phase in dispersed-plug-flow mode, characterized by an axial dispersion coefficient (E_L) and an overall mass transfer coefficient (K_La), uniform fluidization throughout the bed, and constant K_La and E_L throughout the vessel.

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

THREE-PHASE FLUIDIZED BED APPARATUS

DATE	DRAWN BY	FILE NO.	FIG.
2-24-78	BMW	CEPS-X-272	1

8

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

THREE-PHASE FLUIDIZED BED -
OXYGEN CONCENTRATION MEASUREMENTS

DATE 4-24-78	DRAWN BY BWW	FILE NO. CEPS-X-272	FIG. 2
-----------------	-----------------	------------------------	-----------

Table 2. Experimental Program*

D_p (mm)	Column ID (in.)	Viscosity (cp)	$U_G \sim 5$ cm/sec						$U_G \sim 10$ cm/sec						$U_G \sim 15$ cm/sec					
			$U_L/U_{L,MF}$						$U_L/U_{L,MF}$						$U_L/U_{L,MF}$					
			1.25	1.5	2	3	5	7	1.25	1.5	2	3	5	7	1.25	1.5	2	3	5	7
3.2	3	0.9	x	x	x	x			x	x	x	x	x	x	x	x	x	x		
		2.4								x		x	x	x						
	6**	0.9	x	x	x	x			x	x	x	x			x	x	x	x		
4.6	3	0.9	x	x	x	x			x	x	x	x	x	x		x	x	x		
		2.4		x		x	x	x		x	x	x	x	x		x		x	x	x
5	3	0.9	x	x	x	x			x		x	x	x	x	x	x	x	x		
		2.4								x		x	x	x						

* All numbers are intended to give the range.

** U_G 's in this case are 1.2, 2.4, and 3.9 cm/sec.

Derivation of the transport equations describing oxygen concentration profiles along the bed and their solutions were presented by Saad *et al.* (15). In this work a computer program was written to determine the values of $K_L a$ and E_L by matching experimental and theoretical axial oxygen concentration profiles using previous results.

4. RESULTS AND DISCUSSION

4.1 Overall Mass Transfer Coefficient

Typical concentration profiles for three gas velocities are shown in Fig. 3. Mass transfer coefficient, $K_L a$, increased with increasing gas velocity (see Fig. 4), which was in agreement with previous investigations (5, 10, 11, 12, 14). This was due to the increase in gas holdup which resulted in greater interfacial areas and thus in larger $K_L a$.

No consistent relation was found between liquid velocity and $K_L a$. Several other investigators (10, 11, 12) reported $K_L a$ to be independent of U_L , although one reported $K_L a$ to increase with increasing U_L (2). Liquid velocity affected the concentration profile in a consistent manner, as shown in Fig. 5. Increasing U_L caused the lower portion of the bed to become more fluidized, thereby increasing mass transfer resulting in lower concentrations in this section of the bed.

When gas bubbles and solid particles collide, the inertial forces of the particles induce penetration of the gas bubbles while viscous and surface tension forces resist this penetration. Therefore increasing particle size or density increases bubble disintegration, resulting in increased $K_L a$. This phenomenon, observed by others (5, 10, 11, 12), is shown in Fig. 6. Figure 7 shows the effect of particle size on the concentration profile.

It was also shown that surface tension rather than viscosity was the other controlling condition of bubble size. Since the surface tensions of the two liquid phases examined were nearly equal, the bubble sizes were about the same. However, bubble rise velocity was lower in the more viscous liquid (13), resulting in larger gas holdups and thus larger interfacial area in the more viscous water-glycerol solution.

Mass transfer coefficient, K_L , decreases with increasing viscosity due to higher diffusional resistance and increased hydrodynamic boundary layer thickness (4, 5). Therefore, the overall coefficient, $K_L a$, changes accordingly on increasing U_L , as shown in Fig. 8. In the lower U_L region, K_L is the controlling factor and does not change significantly with liquid velocity. In the more turbulent, higher U_L region, differences in K_L due to boundary layer thickness were minimized, and interfacial area became the dominant factor.

3-in.-ID Column

$D_p = 4.6$ mm

$U_L/U_{L,MF} = 3$

$\mu = 0.9$ cp

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

EFFECT OF SUPERFICIAL GAS VELOCITY
ON CONCENTRATION PROFILES

DATE 4-24-78	DRAWN BY BWW	FILE NO. CEPS-X-272	FIG. 3
-----------------	-----------------	------------------------	-----------

3-in.-ID Column

$D_p = 3.2 \text{ mm}$

$\mu_L = 0.9 \text{ cp}$

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

EFFECT OF SUPERFICIAL GAS VELOCITY
ON THE OVERALL MASS TRANSFER COEFFICIENT

DATE
4-24-78

DRAWN BY
BWW

FILE NO.
CEPS-X-272

FIG.
4

3-in.-ID Column
 $D_p = 1.6 \text{ mm}$
 $\mu = 0.9 \text{ cp}$
 $U_G = 9.2 \text{ cm/sec}$

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
 SCHOOL OF CHEMICAL ENGINEERING PRACTICE
 OAK RIDGE NATIONAL LABORATORY

EFFECT OF SUPERFICIAL LIQUID VELOCITY
 ON CONCENTRATION PROFILES

DATE 4-24-78	DRAWN BY JLRB	FILE NO. EPS-X-272	FIG. 5
-----------------	------------------	-----------------------	-----------

3-in.-ID Column

$\mu_L = 7.9 \text{ cP}$

$U_G \sim 9.5 \text{ cm/sec}$

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

EFFECT OF PARTICLE DIAMETER ON
OVERALL MASS TRANSFER COEFFICIENT

DATE
4-24-78

DRAWN BY
JLRB

FILE NO.
CEPS-X-272

FIG.
6

3-in.-ID Column
 $\mu_L = 0.9 \text{ cp}$
 $U_L/U_{L,if} = 5.0$
 $U_G \sim 9.4 \text{ cm/sec}$

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
 SCHOOL OF CHEMICAL ENGINEERING PRACTICE
 AT
 OAK RIDGE NATIONAL LABORATORY

EFFECT OF PARTICLE SIZE ON
 CONCENTRATION PROFILE

DATE 4-24-78 DRAWN BY JLRB FILE NO. CEPS-X-272 FIG. 7

3-in.-10 Column

$D_p = 3.2 \text{ mm}$

$U_G = 9.1 \text{ cm/sec}$

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

EFFECT OF LIQUID VISCOSITY ON
OVERALL MASS TRANSFER COEFFICIENT

DATE
4-24-78

DRAWN BY
JLRB

FILE NO.
CEPS-X-272

FIG.
8

4.2 Axial Dispersion Coefficient

Axial dispersion coefficients (E_L) in the 6-in.-ID column were higher than those in the 3-in.-ID column since the larger column permitted more solids motion and liquid backmixing (see Fig. 9). This is in qualitative agreement with the results of Østergaard (12), who found E_L in a 9-in.-ID column to be higher than that in a 6-in. column.

E_L increased markedly in the 3-in.-ID column at $U_L/U_{L,MF} \sim 2-3$, which was attributed to the change in flow pattern, as shown in Figs. 9 and 10. It was observed that the bottom section of the bed became more fluidized with increasing U_L and was well-fluidized for $U_L/U_{L,MF} > 2-3$, which was also reflected in the concentration profiles shown in Fig. 5. For $U_L/U_{L,MF} = 1.25$, the profiles were adequately described by the plug-flow model. However, for $U_L/U_{L,MF}$ of 3 to 7, the profiles intersected with those of the lower liquid velocities. The decrease in oxygen concentration in the lower portion of the bed was due to increased axial dispersion.

5. CONCLUSIONS

1. Overall mass transfer coefficient $K_L a$ increased on increasing U_G and D_p .
2. At low U_L , $K_L a$ for high-viscosity liquid was lower than that for low-viscosity liquid. The reverse was true at high U_L .
3. Axial dispersion of the liquid, E_L , increased as column diameter increased.
4. At $U_L/U_{L,MF} \sim 2-3$, a sharp increase in E_L was observed due to changes in the flow regime.

6. RECOMMENDATIONS

6.1 Further Studies

1. The dispersed-plug-flow model should be modified to include variations of E_L and $K_L a$ with column height.
2. More hydrodynamic studies should be conducted to determine the factors influencing gas holdup and bubble-size distribution.
3. Effects of column diameter and height should be examined.

$U_B \sim 4$ cm/sec

$D_p = 3.2$ mm

$\mu_L = 0.9$ cp

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

EFFECT OF COLUMN DIAMETER
ON AXIAL DISPERSION

DATE
4-24-78

DRAWN BY
JLRB

FILE NO.
CEPS-X-274

FIG.
9

3-in.-ID Column

$U_G = 9.2$ cm/sec

$D_p = 4.6$ mm

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

EFFECT OF SUPERFICIAL LIQUID VELOCITY
ON AXIAL DISPERSION

DATE	DRAWN BY	FILE NO.	FIG.
4-24-78	JLRB	CEPS-X-272	10

6.2 Equipment Modifications

1. Gas entrainment in the probes should be eliminated by drawing samples through tubes pointing downstream.
2. The 6-in.-ID column should be supported with larger pumps and larger aeration vessels to permit a wider range of fluidization and to help ensure sufficient oxygen reabsorption.
3. The gas-distributor plates should be redesigned to generate a more uniform distribution of gas bubbles.

7. ACKNOWLEDGMENTS

The authors thank their consultants, J.M. Begovich, S.D. Clinton, and J.S. Watson, for their help throughout the project.

8. APPENDIX

8.1 Determination of Minimum Liquid Fluidization Velocity, $U_{L,MF}$

Minimum liquid fluidization velocities, $U_{L,MF}$, at fixed U_G were determined by the method of Bloxom *et al.* (1). $U_{L,MF}$ is defined as the velocity at which the pressure drop across the bed becomes independent of further increases in U_L . As an example, for $U_G = 4.84$ cm/sec, $D_p = 4.6$ mm, and $\mu_L = 0.9$ cp in a 3-in.-ID column, the plot in Fig. 11 was produced, which yields $U_{L,MF} = 2.2$ cm/sec.

8.2 Summary of Experimental Data and Calculated Results

Fit of experimental concentrations to model predictions were measured by the variance:

$$\sigma^2 = \frac{\sum_{i=1}^N (C_{exp} - C_{calc})_i^2}{N}$$

where N is the sample port number. $K_L a$ and E_L were chosen on the basis of minimum variance for each experimental run. Table 3 summarizes the experimental conditions and the results of theoretical calculations.

8.3 Location of Original Data

The original data are summarized in ORNL Databooks A-8143-G, pp. 48-100, and A-8256-G, pp. 31-100, on file at the MIT School of Chemical Engineering Practice, Bldg. 300i, ORNL.

8.4 Nomenclature

a	interfacial area/volume of equipment, cm^{-1}
C	concentration, gmole/cm^3
CST	continuously stirred tank
D_p	particle diameter, mm
DPF	dispersed plug flow

$$U_G = 4.8 \text{ cm/sec}$$

$$D_p = 4.6 \text{ mm}$$

$$\mu_L = 0.9 \text{ cp}$$

MASSACHUSETTS INSTITUTE OF TECHNOLOGY
SCHOOL OF CHEMICAL ENGINEERING PRACTICE
AT
OAK RIDGE NATIONAL LABORATORY

SAMPLE DETERMINATION OF $U_{L,MF}$

DATE
4-24-78

DRAWN BY
JLRB

FILE NO.
CEPS-X-272

FIG.
11

Table 3. Summary of Experimental Data and Calculated Results

3-in.-10 Column, $\mu_L = 0.9$ cp, $D_p = 6$ mm									
Run No.	U_L (cm/sec)	U_L/μ_L (cm/sec)	U_G (cm/sec)	$K_a(\text{CST})$ (sec ⁻¹)	$K_a(\text{PF})$ (sec ⁻¹)	$K_a(\text{DPF})$ (sec ⁻¹)	E_L (cm ² /sec)	r^2	
1	4.71	1.25	4.39	0.176	0.100	0.102	2.814	0.0977	
2	5.59	1.5	4.39	0.285	0.145	0.147	3.32	0.0713	
3	7.46	2.0	4.39	0.271	0.157	0.160	4.414	0.0661	
4	11.3	3.0	4.39	0.325	0.186	0.207	86.33	0.0556	
5	2.47	1.25	9.74	0.670	0.152	0.157	1.06	0.209	
6	4.55	2.25	9.74	0.279	0.129	0.132	2.72	0.065	
7	5.48	2.75	9.74	0.303	0.147	0.154	8.55	0.0537	
8	6.54	3.25	9.74	0.334	0.168	0.175	11.05	0.0540	
9	2.5	1.25	14.05	1.358	0.195	0.201	1.0	0.935	
10	3.02	1.5	14.05	0.809	0.165	0.170	1.4	0.205	
100	4.0	2.0	14.05	0.461	0.156	0.159	2.4	0.218	
11	5.0	2.5	14.05	0.435	0.170	0.174	3.0	0.127	
12	5.96	3.0	14.05	0.429	0.185	0.198	14.2	0.131	
91	11.15	5	9.15	0.261	0.148	0.179	161.6	0.214	
92	15.6	7	9.15	0.262	0.157	~0.2	~700	~0.4	
3-in.-10 Column, $\mu_L = 0.9$ cp, $D_p = 4.6$ mm									
37	3.62	1.25	4.84	0.248	0.109	0.109	1.538	1.391	
38	3.59	1.5	4.84	0.194	0.103	0.103	1.803	0.738	
39	4.78	2.0	4.84	0.128	0.082	0.0823	7.598	0.0961	
40	7.17	3.0	4.84	0.199	0.120	0.120	4.964	0.153	
41	2.14	1.25	9.24	0.571	0.117	0.121	1.3	0.2207	
42	2.64	1.5	9.24	0.342	0.108	0.111	1.6	0.1214	
43	3.27	2.0	9.24	0.236	0.101	0.103	2.0	0.1197	
44	4.91	3.0	9.24	0.192	0.108	0.108	19.027	0.120	
45	2.26	1.5	13.8	0.574	0.121	0.125	1.4	1.172	
46	3.02	2.0	13.8	0.435	0.129	0.133	1.8	0.1615	
47	4.53	3.0	13.8	0.284	0.130	0.130	2.71	0.0708	
93	8.50	5.0	9.15	0.1545	0.0987	0.1108	86.49	0.3266	
94	11.90	7.0	9.15	0.1561	0.1039	0.1226	289.7	0.0831	
3-in.-10 Column, $\mu_L = 0.9$ cp, $D_p = 3.2$ mm									
13	2.04	1.25	4.63	0.135	0.0657	0.067	0.807	3.11	
14	2.44	1.5	4.63	0.0736	0.048	0.048	1.197	0.580	
15	3.25	2.0	4.63	0.0362	0.0289	0.0289	1.952	0.041	
16	4.88	3.0	4.63	0.0311	0.0270	0.0270	32.97	0.0958	
17	1.57	1.25	9.27	0.320	0.077	0.0797	1.0	0.597	
18	1.83	1.5	9.27	0.205	0.0712	0.0729	1.1	0.637	
19	2.51	2.0	9.27	0.0959	0.0541	0.0541	1.488	0.157	
20	3.76	3.0	9.27	0.0752	0.0525	0.0525	18.605	0.0537	
21	1.4	1.25	13.88	0.354	0.0745	0.0770	0.9	0.320	
22	1.7	1.5	13.88	0.259	0.0744	0.0765	1.0	0.260	
23	2.26	2.0	13.88	0.170	0.0733	0.0749	1.4	0.153	
24	3.39	3.0	13.88	0.129	0.0730	0.0915	39.938	0.0175	
95	6.30	5.0	9.1	0.0757	0.0560	0.0625	100.1	0.252	
96	8.82	7.0	9.1	0.0945	0.0582	0.0765	161.8	0.1529	

Table 3. (Continued)

3-in.-ID Column, $\mu_L = 2.4$ cp, $D_p = 4.6$ mm								
Run No.	U_L (cm/sec)	$U_L/U_{L,PF}$ (cm/sec)	U_G (cm/sec)	$K_L a$ (CST) (sec ⁻¹)	$K_L a$ (PF) (sec ⁻¹)	$K_L a$ (DPF) (sec ⁻¹)	E_L (cm ² /sec)	σ^2
51	2.3	1.5	4.84	0.148	0.0674	0.0674	1.37	0.521
52	4.65	3.0	4.84	0.148	0.0851	0.0851	3.26	0.0809
53	7.68	5.0	4.84	0.166	0.109	0.109	5.31	0.138
54	9.85	7.0	4.84	0.131	0.0872	0.107	337.24	0.187
60	1.9	1.5	9.2	0.165	0.0646	0.0702	4.7	0.1318
61	2.6	2.0	9.2	0.158	0.0735	0.0735	2.56	0.551
62	3.9	3.0	9.2	0.191	0.0980	0.110	20.3	0.498
63	6.5	5.0	9.2	0.144	0.0897	0.1025	64.14	0.549
64	9.0	7.0	9.2	0.202	0.116	0.1316	76.38	0.192
65	1.6	1.5	13.9	0.214	0.0663	0.07556	5.6	0.768
66	3.1	3.0	13.8	0.314	0.114	0.130	12.9	0.8528
67	5.1	5.0	13.8	0.258	0.115	0.137	45.4	0.616
68	7.2	7.0	13.8	0.329	0.154	0.154	4.72	1.71
3-in.-ID Column, $\mu_L = 2.4$ cp, $D_p = 3.2$ mm								
70	1.7	1.5	9.1	0.0706	0.0386	0.0479	16.58	0.529
71	3.4	3.0	9.1	0.0964	0.061	0.0703	31.2	0.146
72	5.7	5.0	9.1	0.0541	0.064	0.0731	78.56	0.0552
73	7.9	7.0	9.1	0.130	0.083	0.0933	89.9	0.0873
6-in.-ID Column, $\mu_L = 0.9$ cp, $D_p = 3.2$ mm								
25	3.29	1.25	1.17	0.0333	0.0297	0.0333	2.958	0.115
26	3.90	1.50	1.17	0.00460	0.00453	0.00459	1.041	0.609
27	5.26	2.0	1.17	0.023	0.0219	0.0231	19.89	0.00056
28	8.0	3.0	1.17	0.0771	0.0652	0.0770	91.64	0.251
29	3.07	1.25	3.91	0.0440	0.0377	0.0440	154.36	0.00037
30	3.70	1.50	3.91	0.0523	0.0448	0.0523	18.95	0.0238
31	5.04	2.0	3.91	0.0751	0.0639	0.0751	33.84	0.240
32	7.44	3.0	3.91	0.0713	0.0640	0.0713	14.6	0.201
33	3.32	1.25	2.42	-0.00225	-0.002269	-0.00225	55.75	0.04098
34	3.99	1.5	2.42	0.0163	0.0153	0.0163	44.88	0.220
35	5.32	2.0	2.42	0.0166	0.0158	0.0166	2.08	0.3388
36	8.0	3.0	2.42	0.110	0.0873	0.0873	64.31	0.237

E_L	liquid-phase axial-dispersion coefficient, cm^2/sec
H	height of column, bed, or highest sample port, cm
h	height above bottom of column or bed, cm
K_L	mass transfer coefficient based on liquid side, cm/sec
m	Henry's Law constant, $C_{L,\text{eq}}/C_G$
ΔP	pressure drop, cm H_2O
PF	plug flow
U	superficial fluid velocity, cm/sec
X	dimensionless approach to equilibrium, $\frac{mC_{G,\text{in}} - C_L}{mC_{G,\text{in}} - C_{L,\text{in}}}$

Greek Symbols

ϵ	volume fraction
μ	viscosity, $\text{gm}/\text{cm}\cdot\text{sec}$
ρ	density, gm/cm^3
σ^2	variance

Subscripts

calc	calculated value
eq	equilibrium
exp	experimental value
G	gas phase
in	inlet
L	liquid phase
MF	minimum fluidization
s	solid phase

8.5 Literature References

1. Bloxom, S.R., J.M. Costa, J. Herranz, S.R. Roth, and G.L. MacWilliam, "Determination and Correlation of Hydrodynamic Variables in a Three-Phase Fluidized Bed," ORNL/MIT-219 (1975).
2. Begovich, J.M., "Advanced Technology Section Semiannual Report: Engineering Science Program, April 1977 to September 1977," (in preparation).
3. Burck, G.M., K. Kodama, R.G. Markeloff, and S.R. Wilson, "Cocurrent Three-Phase Fluidized Bed, Part 2," ORNL/MIT-213 (1975).
4. Danckwerts, P.V., "Gas-Liquid Reactions," McGraw-Hill, New York, 1970.
5. Dawson, S.G., E.A. Bagnall, H.J. Herzog, and V. Zakaria, "The Effects of Viscosity and Mass Transfer in Open Bubble Columns," ORNL/MIT-199 (1974).
6. Davidson, J.F., and D. Harrison, "Fluidization," pp. 751-780, Academic Press, London, 1971.
7. Khosrowshahi, S., S.R. Bloxom, C. Guzman, and R.M. Schlapfer, "Cocurrent Three-Phase Fluidized Bed, Part 3," ORNL/MIT-216 (1975).
8. Kunii, D., and O. Levenspiel, "Fluidization Engineering," Wiley, New York, 1969.
9. Massimilla, L., A. Solimando, and E. Squillace, "Gas Dispersion in Solid-Liquid Fluidised Beds," Brit. Chem. Eng., p. 232 (April 1961).
10. Østergaard, K., and P. Fosbøl, "Transfer of Oxygen Across the Gas-Liquid Interface in Gas-Liquid Fluidised Beds," Chem. Eng. J., 3, 105 (1972).
11. Østergaard, K., and W. Suchozebrski, "Gas-Liquid Mass Transfer in Gas-Liquid Fluidized Beds," Proc. 4th European Symp. Chem. Reaction Eng., Pergamon Press, Oxford, 1969.
12. Østergaard, K., "Hold-up, Mass Transfer, and Mixing in Three-Phase Fluidization," presented at AIChE 69th Ann. Meeting, Chicago (1976).
13. Perry, R.H., and C.H. Chilton, "Chemical Engineers' Handbook," 5th ed., McGraw-Hill, New York, 1973.
14. Reber, S.A., J.H. Austin, P.B. Jackson, R.Y. Shah, and R.M. Train, "The Effects of Liquid Flow Rate and Viscosity on Mass Transfer in Open Bubble Columns," ORNL/MIT-201 (1974).
15. Saad, E.T., H.D. Ayala, W.M. Burke, and D.K. Clarkson, "Cocurrent Three-Phase Fluidized Bed, Part 1," ORNL/MIT-209 (1975).

ORNL/MIT-272

INTERNAL DISTRIBUTION

- 1-2. J.M. Begovich
- 3. D.A. Canonico
- 4-5. S.D. Clinton
- 6. H.D. Cochran, Jr.
- 7. C.W. Hancher
- 8. J.R. Hightower, Jr.
- 9. R.E. Leuze
- 10. Yvonne Lovely
- 11. G.E. Moore
- 12. J.J. Perona
- 13. J.B. Ruch
- 14. C.D. Scott
- 15. J.S. Watson
- 16-17. Central Research Library
- 18. Document Reference Section
- 19-21. Laboratory Records
- 22. Laboratory Records, ORNL R.C.
- 23. ORNL Patent Office
- 24-38. MIT Practice School

EXTERNAL DISTRIBUTION

- 39. J.E. Vivian, MIT
- 40. Morgantown Energy Research Center, DOE, Morgantown, W. Va.
- 41. Director, Research & Technical Support Div., DOE, ORO
- 42-43. Technical Information Center