1.0 OBJECTIVE

The overall objective of this project was to study solids accumulation and suspension of various gas/liquid/solid systems in cold-flow tubular columns to provide data to design the coal dissolver for the SRC-I Demonstration Plant.

The specific objectives of this program were:

1. to study the effects of slurry velocity, gas velocity, solid particle size, solids concentration, liquid viscosity, and surface tension on the performance of a cold-flow tubular column;
2. to develop an effective technique to withdraw slurry from the bottom of a tubular column as a means to control solid concentration;
3. to study the performance of cold-flow tubular columns both with an improved distributor and without a distributor.

2.0 INTRODUCTION

A major part of the coal-dissolution section of any liquefaction plant is the dissolver. Although considerable liquefaction occurs in the preheater, most of the necessary chemical changes occur in the dissolver, including sulfur removal, oil and distillate formation, and solvent rehydrogenation.

Vertical tubular reactors are used in all of the major processes currently under consideration for commercial coal liquefaction. In the Solvent Refined Coal (SRC), Exxon Donor Solvent (EDS), and H-Coal processes, slurry and gas are concurrently fed upward through these vessels. In the EDS and SRC processes, the reactors are basically empty vessels, whereas in the H-Coal process a bed of ebullating catalyst is maintained in the reactor. Between the EDS, SRC-I, and SRC-II processes, the major differences in dissolver operation are the compositions of the feed streams and reactants within the dissolver. Different hardware such as the distributor plates, draft tubes, or recycle loops can also affect the behavior of slurries in these vessels.

In order to design a technically feasible and cost effective dissolver, the physical behavior of three-phase systems in tubular columns must be clarified. All of the major processes under development require an understanding of backmixed three-phase systems. In each process, part of the dissolver volume is backmixed. As the design of the 6,000-ton-per-day (tpd) SRC-I Demonstration Plant progresses, the increased vessel size (and other considerations) may dictate the use of reactors in series, which would decrease the overall backmixed characteristic of the commercial plant.

The SRC-I Demonstration Plant dissolver will be scaled up considerablly from the Wilsonville and Ft. Lewis Pilot Plant dissolvers. The relative sizes of the dissolvers for the pilot plants and the SRC-I Demonstration Plant are compared below (at a residence time of 0.56 hr, a gas-feed rate of 20,000 mscf per ton of dried coal, and a 38% coal slurry):

Plant	Size (tpd)	Volume (ft^3)	Dissolver diameter	Height (ft)	Superficial Velocities (ft/sec)	
					Liquid	Gas
Wilsonville	6-10	18.1	12 in.	23	0.012	0.074
Ft. Lewis	50-61	106.8	24 in.	34	0.017	0.10
SRC-I Demo Plant	6,000	10,454.0	11 ft	110	0.06	0.36

The reactor volumes and diameters differ dramatically. Thus, to make intelligently good decisions about the designs, we need more information on the flow properties of three phase systems in large vessels.

Also important is the effect of differences in gas and liquid superficial velocities on slurry behavior. A fivefold difference in velocity exists between Wilsonville and the commercial SRC-I design. This difference can have a considerable impact on the process because the gas and liquid superficial velocities have a strong effect on (a) gas void volume, (b) actual solids concentration in the dissolver, and (c) the relative degree of backmixing. As velocity through the dissolver increases, the tendency for solids to remain behind diminishes, causing a decrease in the actual concentration of ash particles in the reactor. Those particles that do remain will tend to be larger. Considerable evidence shows that reactor solids have a definite catalytic effect. Larger particles will have fewer surface areas exposed, and thus will probably have diminished catalytic activity. Knowing the particle sizes that can accumulate under commercial flow conditions will give us some indication of the size of dissolver solids that should be examined for catalytic activity.

3.0 BACKGROUND

3.1 Gas/Liquid Flow

Numerous investigators (1-28) have studied the behavior of the gas and liquid components in two-phase flow. With the fluid as the continuous phase, the gas-flow pattern can be categorized as bubbly, slug, or churn-turbulent. The pattern is determined primarily by the competing rates of bubble coalescence and breakup, which, in turn, depend on liquid-phase properties such as viscosity, surface tension and density. During bubble coalescence, as the bubbles draw together, the thin film of liquid separating the bubbles ruptures. Apparently, gas bubbles coalesce by the capture of one bubble in the wake of another rising bubble; this has been extensively studied by Crabtree and Bridgwater (18). Calderbank et al. (24) reported that increased liquid viscosity enhances the rate of bubble coalescence. Bubble breakup is due to disturbance at the interface. In the early fifties, Davies and Taylor (17) performed a theoretical study on the instability of liquid surfaces. Clift and Grace (23) proposed in 1972 that bubble breakup is due to this instability.

One of the most important aspects of gas flowing through a tubular reactor is its availability at the reaction site. The transfer of gas to that site, be it in a homogeneous liquid-phase transition state or a heterogeneous surface-activated state, is ultimately related to the interfacial area between the liquid and gaseous phases. In a two-phase gas/liquid system, the interfacial area is related to the volume fraction actually occupied by gas. This volume fraction, which is typically referred to as the gas holdup or void fraction, is a value that is extremely important to optimum reactor design.

In a two-phase flow reactor, gas residence time is governed by the gas holdup at the operating conditions and mass transfer is governed by gas contact at the interfacial phase barrier. Numerous investigators (8,10,11,13,14) have studied gas holdup with many different systems, covering a wide range of liquid viscosity (0.58-152.0 cp), surface tension (22.3-76.0 dynes/cm), and density (0.79-1.70 gm/cm^3). Several correlations have been developed to predict the gas-void fraction in two-phase flow systems. In general, the gas-void fraction is expressed in terms of gas superficial velocity, liquid surface tension, viscosity, and density. A partial list of the existing correlations is provided in Table 1.

Generally, gas-void fraction does not depend on liquid input rate. Gas holdup is usually higher for a smaller column diameter because of the wall proximity effect. Ellis and Jones (25) studied the effect of column diameter on gas holdup and concluded that, for diameters greater than 3 in., gas holdup is independent of column diameter. Directionally, work at Air Products agrees with their results.

Mass transfer in bubble columns has been studied extensively (1,7,9,12, 13,16,19,21,22,26). Because the mass-transfer resistance in the gas phase is negligible compared with that in the liquid phase, the overall mass-transfer coefficient is dominated by that of the liquid phase. In general, the volumetric liquid-phase mass-transfer coefficient (k_La) depends on gas-input velocity, column diameter, bubble diameter, and liquid properties (diffusivity, surface tension, viscosity, and density). Several correlations have been developed (7,12,16,19-22,24). The dependence of mass transfer on liquid velocity and gas properties (density and viscosity) has also been proposed (7,19,20) but this is not unanimously accepted. A higher gas-input rate increases mass transfer because it enhances interfacial area and liquid turbulence.

Another significant aspect of gas/liquid flow is the dispersion (back-mixing) of the fluid phase. In studies at Air Products, we found that in a gas/liquid flow column an extremely small amount of gas flow can trigger a high degree of dispersion in the liquid phase. Shah, et al., (15) have extensively reviewed backmixing in gas/liquid reactors. In general, backmixing in the liquid phase depends on column diameter, gas velocity, and the nature of the gas distributor plate. Most investigators agree that the axial dispersion coefficient is independent of liquid-flow rate and liquid-phase properties (surface

Table 1

Correlations for Predicting Gas Holdup

- Hughmark (16)

$$\varepsilon_g = \frac{V_{sG}}{U_s} \left[\left(\frac{62.4}{\rho_L}\right)\left(\frac{72}{\sigma}\right)\right]^{1/3}$$

- Akita and Yoshida (26)

$$\frac{\varepsilon_g}{(1-\varepsilon_g)^4} = C_1 \, (N_{Bo})^{1/8} \, (N_{Ga})^{1/12} \, (N_{Fr})$$

- Hikita and Kikukawa (3)

$$\varepsilon_g = 0.505 \, V_{sG}^{\,0.47} \, (72/\sigma)^{2/3} \, (1/\mu)^{0.05}$$

where ε_g = gas-void fraction
V_{sG} = gas superficial velocity (ft/sec)
U_s = slip velocity (ft/sec)
ρ_L = liquid density (lb/ft^3)
σ = liquid surface tension (dynes/cm)
N_{Bo} = $gD^2\rho_L/\sigma$ (Bond Number)
N_{Ga} = $gD^3/\upsilon_L^{\,2}$ (Galileo Number)
N_{Fr} = $V_{sG}/\sqrt{gD}$ (Froude Number)
C_1 = 0.20 for nonpolar solutions, 0.25 for polar solutions
D = column diameter (ft)
υ_L = liquid kinematic viscosity (ft^2/sec)
g = gravitational acceleration (32.2 ft/sec^2)
μ = liquid viscosity (Cp)

tension, viscosity, and density). Many correlations have been developed to predict the axial dispersion coefficient in gas/liquid flow reactors. Table 2 lists some correlations for gas as the dispersion phase and liquid as the continuous phase.

At low gas velocities, the agreement among these published correlations is poor. For example, for a 5-in.-diameter column, we estimated a sixfold difference in the predicted dispersion coefficient at a 0.02 ft/sec superficial gas velocity:

Axial Dispersion Coefficients (ft^2/sec) from Published Correlations

Reference	Low Gas Velocity (0.02 ft/sec)	High Gas Velocity (.33 ft/sec)
28	0.047	0.190
1	0.088	0.204
2	0.022	0.053
3	0.134	0.219
4	0.129	0.229
6	0.091	0.239

However, at high gas velocities these correlations give better agreement. Because of the inconsistencies in these published correlations, guidelines are unavailable for predicting the dispersion coefficient, which is an essential value for designing a reactor.

To anyone initiating studies in this area, experiments in two-phase flow should be performed to make certain that experimental techniques are adequate to reproduce existing data. Air Products has performed such studies and has developed standard techniques that do reproduce existing data. From these studies, a keen understanding of the value of the correlative models has been gained, which will be applied in the proposed program.

Table 2

Correlations for Predicting Axial Dispersion Coefficients

Kato and Nishiwaki(4)

$$E_{ZL} = \frac{V_G D \,(1 + 6.5\, Fr_G^{0.8})}{13\, Fr_G}$$

Towel and Ackerman(28)

$$E_{ZL} = 1.23\, D^{1.5}\, V_G^{0.5}$$

Cova(2)

$$E_{ZL} = 0.0759\, V_G^{0.32}\, \rho_L^{0.07}$$

Deckwer, et al(1)

$$E_{ZL} = 0.678\, D^{1.4}\, V_G^{0.3}$$

Hikita and Kikukawa(3)

$$E_{ZL} = (0.366 + 0.674\, V_G^{0.77})\, D^{1.25}\, (1/\mu_L)^{0.12}$$

Baird and Rice(6)

$$E_{ZL} = 0.35\, D^{4/3}\, (V_G g)^{1/3}$$

Ying(49)

$$E_{ZL} = 0.27\, D\, V_G \left(\frac{Dg}{V_G^2}\right)^{0.32}$$

where Fr_G = Froude Number = $\sqrt{V_G^2/gD}$
E_{ZL} = axial dispersion coefficient (ft^2/sec)
V_G = gas superficial velocity (ft/sec)
D = column diameter (ft)
ρ_L = liquid density (g/cm^3)
μ_L = viscosity of liquid (cP)
g = acceleration of gravity (ft/sec^2)

3.2 Gas/Liquid/Solid Flow

Adding solid particles to a gas/liquid flow reactor greatly increases the complexity of the system. The presence of the solid itself, as well as the particle size, size distribution, and density, can affect the behavior significantly. The rate of bubble coalescence increases in the presence of solid particles. Kato et al. (29, 30) observed larger coalesced bubbles in a gas/liquid/solid system. However, this enhancement in the rate of bubble coalescence decreases to an insignificant level in a turbulence regime at high gas velocities.

Studies of gas holdup in a three-phase flow reactor (Air Products work) showed that the gas-void fraction decreases in the presence of solid particles at low gas velocities. At high gas velocities, the dependence of gas holdup on solid particles disappears. These effects of solid particles on gas holdup are directly in line with the observation of the bubble coalescence rate in the presence of solid particles.

A unique feature of three-phase flow is the contraction of the fluidized bed (34, 35, 37, 39-42) upon addition of gas to the system, as first reported by Turner (34, 35). Empirical expressions have been developed and an iterative method has been employed to determine the volume fraction of the phases (37, 39, 41).

Several investigators have also studied the effects of solid particles on liquid backmixing in three-phase flow systems (31, 38). Some data of (31) indicate that the presence of solid particles reduces liquid backmixing.

The effects of particle size, gas velocity, and liquid velocity are interrelated. An empirical correlation was developed by Kato et al. (29, 30) to predict the dispersion coefficient of liquid in a slurry. The correlation does not depend on particle size or liquid velocity.

$$E_{ZSL} = \frac{V_{SG}\,D(1 + 8\,Fr_G^{0.85})}{13Fr_G}$$

where E_{ZSL} = dispersion coefficient of liquid in a slurry (ft^2/sec).

Comparison of this correlation with an earlier correlation (4) for liquid dispersion in a gas/liquid system indicates that the liquid-phase dispersion coefficient is higher in a slurry than in a clear fluid, except when the Froude number is very small. This result differes from the findings of Vail et al.

Kato et al., (29) also studied the dispersion of the solid phase in a gas/liquid system. Their results lead to two empirical correlations, depending on the particle size. For small particles the dispersion coefficients for both the liquid and solid phases are identical. For large particles, the empirical condition includes the effect of particle diameter, as shown in the following expression.

$$E_{ZS} = \frac{V_{sG}D(1 + 8Fr_G^{0.85})}{13Fr_G(1 + 0.009Re_pFr_G^{-0.8})}$$

where $Re_p = d_pV_p/\upsilon_L$;
E_{ZS} = dispersion coefficient of solid in a slurry (ft^2/sec);
d_p = particle diameter (ft);
V_p = particle terminal velocity in stagnant fluid (ft/sec); and
υ_L = liquid kinematic viscosity (ft^2/sec).

The suspension of solid particles in a batch column depends on the gas-input rate. Roy et al., (36) and Imafuku et al. (32) have studied the critical gas velocities for a complete solid suspension. Critical gas velocity is defined as the minimum gas-input rate needed to fully suspend the solid particles. Imafuku et al., (32) observed that the critical gas velocity depends on the shape of the bottom inlet. They recommend conical bottom; the position of the gas distributor plate is also important to obtain complete solid suspension.

Our experience in three phase flow behavior indicates that pertinent data to define behavior in these systems is lacking.

3.3 Solids Accumulation

High solids accumulation has been experienced in the dissolvers at Ft. Lewis and Wilsonville Pilot Plants. Hydrocarbon Research, Inc., (HRI) when running an SRC study (44), also observed solids buildup in their experimental reactor. However, the most quantitative data on solids accumulation are those available from Wilsonville. For over 2 years, they have reported extensive data on rates of accumulation, quantities, particle sizes, and chemical composition of accumulated materials.

To show the levels of solids that can accumulate, data for four Wilsonville runs, each on a different coal, are presented in Table 3. The highest level of accumulation was in Run 100 with Amax Belle Ayr coal from which 1,300 pounds of reactor solids was discharged. This corresponds to a solids concentration of 72 lb/ft^3. On a volume basis, these solids occupy about 40% of the total dissolver space. Whether or not these solids are completely suspended at this concentration in unknown. Obviously, such information is critical for designing a commercial vessel that has a diameter of about 11 ft, as proposed for the SRC-I Demonstration Plant.

In addition to the data from Wilsonville on solids composition, an extensive evaluation of reactor solids was carried out at Pennsylvania State University (45). That study led to an understanding of the mechanism of solids formation. The relative growth of particles was related to the calcium content in the feed coal: as calcium content increased, particle growth and the quantity of accumulated solids also increased.

Wilsonville also generated considerable data on the particle-size distributions of the recovered solids that accumulated when plant grind coal (95% through 200 mesh) was fed. The solids recovered from

Table 3

Wilsonville Solids Accumulation

Coal	Run length (days)	LHSV[a] in dissolver (hr^{-1})	Superficial velocities (ft/sec)		Solids in reactor (lb/ft^3)	% ash in solids
			Slurry	Gas		
Kentucky 9 & 14	68	1.7	0.011	0.09	33	83
Monterey	95	1.7	0.011	0.05	22	70
Indiana V	129	0.9	0.006	0.03	53	61
Amax Bell Ayr	100	1.1	0.007	0.06	72	71

[a] LHSV, liquid hourly space velocity

the reactor grew considerably, as demonstrated by the amounts of material that accumulated on 25- and 50-mesh screens. For example, the particle-size distribution for the solids collected after Run 68, (Table 3) was as follows:

Mesh size	wt %
25	5
25-50	7
50-100	5
100-200	50
200-325	24
325	9

Obviously, the larger particles would not have been in the feed coal. Other Wilsonville reports shown even higher concentrations of larger particles in the reactor solids.

Since mid-1976, attempts have been made to control the concentration of solids in the Wilsonville reactor. Reactor solids were added at the start of individual runs to bring the dissolver to a rapid line-out. More recently, a solids withdrawal system was added to maintain a steady solids concentration in the reactor. With the capability to both add and withdraw solids, valuable data will undoubtedly be obtained that can be used for the commercial plant design. However, additional data at higher flows and with different geometries will be needed to effectively use such a withdrawal system on a larger scale.

At commercial design conditions, accumulation of solids will almost certainly occur. Exxon's results on 8-mesh feed coal showed the necessity for a solids removal system (46). Also, the DOE Project Review Meeting on Preheater Design held at Oak Ridge on 21 March 1979, C. Ackerman (P & M Mining Co.) reported 1/16-1/8 in. rock accumulation in the Ft. Lewis dissolver after only 3-12 days operation with 1/8 in. coal feed. Data on Ft. Lewis reactor design (see Section 2) and data from Wilsonville (Table 3) Run 68 show that relative superficial gas velocities are very similar, i.e., 0.11 and 0.09 ft/sec, respectively. At these velocities, solids that accumulate when plant grind coal is used have particle sizes less than 100 mesh. Certainly much higher levels of accumulation would be expected with the larger coal feed size planned for the commercial plants.

The relevance of the above data at superficial gas velocities of 0.1 ft/sec is apparent when the alternate option of running the 6,000 tpd dissolver at 0.18 ft/sec (parallel dissolvers) is a distinct possibility. Knowing the behavior of solids in a column at these flow velocities is good engineering practice.

Very little data on the retention of solids in dissolvers are available from other sources. Ft. Lewis data have been limited largely to the time that a considerable plug was discovered in their dissolver (48). Otherwise, the behavior of solids in their system has not been studied systematically. H-Coal researchers have studied the flow rates that allow retention of solid catalysts within their reactor. However, the catalyst particles used in the H-Coal process are larger than the ash particles anticipated to accumulate in the SRC dissolver. The H-Coal process will probably use a 1/16 in. extrudate for their commercial design.

Apparently, particle growth within the dissolver causes the large accumulation of solids having particle sizes greater than those present in the feed coal. The program at Pennsylvania State University has shown that the particles can be attributed to a shell growth effect.

At Air Products, to understand the problem of solids accumulation, critical gas velocities were determined for different particle sizes in three liquid media using a 5-in.-diameter column. Critical gas velocity is the superficial velocity beyond which the particles will be in complete suspension.

Table 4 summarizes the results of the critical gas velocity study. The effect of particle size on solids suspension or critical gas velocity is significant; the critical gas velocity increases substantially as particle size increases. When the gas velocity dropped below the critical value, sand particles settled to the bottom and the gas channeled through the settled bed. The amount of settled solids decreased with increasing gas velocity. As the critical gas velocity was reached, the momentum transferred from the gas phase was large enough to suspend all the sand particles.

Also shown in Table 4 is the sieve opening (microns) for each particle range. Assuming spherical particles have diameters equal to the sieve opening, the settling velocities of sand particles were calculated based upon particle drag in an infinite water medium (Table 5). It is reasonable to assume that the largest particle, or the maximum sieve opening in each group, would govern the critical gas velocity. It is then interesting to note that the critical gas velocity in each group (Table 5) is approximately twice the settling velocity of the largest particle of the corresponding group (with the exception of Group II). Extrapolating these results, the Group VI particles (20-30 mesh) are predicted to have a critical gas velocity of about 0.7 ft/sec. In any event, the results of this critical gas velocity study strongly suggest the problem of settled solids when large particles are used.

The physical properties of a liquid also have a significant impact on solid suspension (Table 4). When a mixture of ethanol and water was tested, the critical gas velocities for Groups II and V were reduced by approximately 30% of those measured in pure water. After the physical properties of these two solutions were examined, two possible causes for this reduction were apparent: surface tension and viscosity. However, the experiment with pure ethanol eliminates the possibility of the surface tension effect. The surface tension of the mixed solution was determined experimentally to be 37.43 dynes/cm, which is more than a factor of two lower than that of pure water. If liquid surface tension were responsible for the observed reduction of the

Table 4

Effects of Particle Size and Liquid Properties On Critical Gas Velocities

Group No.	Particle size		Critical gas velocities (ft/sec) at various liquid phases		
	Mesh	μm	Water	Ethanol/water[a]	Ethanol
I	140 minus	Less than 105	0.13	---	---
II	80-120	177-125	0.137-0.165	0.10-0.11	0.137-0.165
III	60-80	250-177	0.193-0.217	---	---
IV	45-60	345-250	0.298-0.362	---	---
V	30-45	595-354	0.482-0.503	0.344	---
VI	20-30	841-595	No full suspension detected up to 0.57-ft/sec gas velocity	---	---

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[a] 30 wt % ethanol/70 wt % water mixture.

Table 5

Particle Settling Velocity in Infinite Water Medium

Mesh No.	Diameter(μm)	Settling velocity (ft/sec)
80	177	0.0567
60	250	0.0971
45	354	0.1535
30	595	0.2592
20	841	0.3707

critical gas velocity in the mixture, pure ethanol, which has a surface tension (22.75 dynes/cm) lower than that of the mixture, should show a similar reduction. Surprisingly, our results showed that the critical gas velocity for the 80-120-mesh solids in pure ethanol was indistinguishable from that observed in pure water (Table 4). Hence, liquid surface tension is not responsible for the reduction effect observed in the mixture, thereby leaving the liquid viscosity as the sole explanation.

The solid settling velocity was lower in the mixed solution than in pure water because the viscosity between these two solutions differed by a factor of three. Going through the exercise of particle drag calculation, the settling velocities of a 30-mesh (595 microns) particle in the mixed solution and the pure ethanol were found to be 0.13 and 0.24 ft/sec, respectively. The insignificant difference between pure ethanol and pure water clearly explains the indistinguishable critical gas velocities measured in those two solutions. On the other hand, the twofold difference in the settling velocities of the 30-mesh particle in mixed solution and pure water explains the reduction in critical gas velocity. This also agrees with the findings on particle-size effect, which showed that critical gas velocity increased with increasing particle settling velocity. As a rough rule of thumb, the ratio of the critical gas velocity to the particle settling velocity is about 2, in the absence of liquid flow.

The operating conditions for the SRC-I pilot and demonstration plants were presented in Section 2. Results presented in Table 4 show that under demonstration plant conditions, solids smaller than 30-45 mesh would not be expected to accumulate, whereas both Wilsonville and Ft. Lewis dissolvers accumulate solids. The actual distribution of mesh sizes cannot be determined, because larger particles displace smaller ones and any size larger than 80-100 mesh may accumulate at Wilsonville. That this is clearly so at Wilsonville has been reported.

3.4 Data Needs for Dissolver Design

The design goal for the commercial dissolvers is to ensure that solids passing through the reactors remain suspended in sufficient quantity and distribution to provide the necessary catalytic effect. Solids and ash particles can accumulate over an extended operating period. Data from Ft. Lewis on rock accumulation in the dissolver when 1/8-in. coal is fed illustrates the problems that may occur commercially. From a design standpoint the Ft. Lewis results are quite alarming, because Tarrer and co-workers have shown that coal fed to the process unit would immediately dissolve into liquid, leaving extremely small ash particles suspended in the reactor.

Information necessary for dissolver design requires the following data:

- Determine the necessary superficial gas and slurry velocities needed to suspend these particles at turndown velocities
- Determine the solids sizes necessary to ensure adequate solids accumulation within the reactor at commercial linear liquid and gas velocities
- Determine adequate slurry correlations to predict accumulations in the commercial reactors, especially for start-up mode and equilibrium operation after solids removal
- Determine dispersion (backmixing) as a function of fluid and gas velocities, solids concentration, density, and particle size
- Determine in a large system the effect of surface tension on gas void fraction and solids suspension
- Determine factors that influence adequate solids removal

- Determine the influence of distributor plates on fluid properties as a function of position relative to the plate

- Determine the efficiency of solids removal with regard to plate design and location

- Examine solids removal as a function of velocity through the draw-off line and its effect on dissolver behavior.

Upon starting the design of the 6,000-tpd SRC-I Demonstration Plant, Air Products realized the necessity for large-scale cold-flow simulator studies. A glass column 25 ft tall with a 12-in. internal diameter was selected as having the required material strength and affording maximum visual access to the behavior of the fluid within the column. This design also corresponds exactly to the dissolver in use at Wilsonville.