

## **Liquid Back-Mixing on Distillation Trays**

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### **ABSTRACT**

Experimental results are presented on the liquid back-mixing of distillation trays. The tests were conducted in the 1.22m diameter column of the F.R.I distillation unit using different systems at various pressures. Both bubble cap and sieve trays with different designs were tested. For each test, a pulse of non-volatile dye was injected into the reflux stream, and the liquid on several trays was continuously monitored to determine the effects of the trays on this pulse. The degree of liquid back-mixing on the tray was then deduced as a function of the change in the shape of the pulse. The results show that liquid back-mixing is strongly dependent on tray designs and system properties. Certain trays and systems had a high degree of plug flow, others were highly mixed. The test results indicate that the liquid back-mixing occurring in the tray deck is much larger than that in the downcomer. Mixing pool numbers and eddy diffusivity were computed from the experimental data. The results are compared to predictions from several literature models. The predictions from the Bennett et al. and Zuiderweg correlations agree reasonably well with the experimental results.

Key Words: Distillation column, trays, liquid back-mixing, tray efficiency

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## INTRODUCTION

The separation efficiency of any vapor/liquid contacting internal device such as sieve tray, valve tray, etc., can be interpreted in two ways: as a point/local efficiency or an overall tray efficiency. The overall tray efficiency can be determined as a “Murphree Tray Efficiency”,  $E_{MV}$ , or as an “Overall Tower Efficiency”,  $E_O$ . The theoretical interpretations of mass transfer and the prediction of tray efficiency are made in terms of point or local efficiency,  $E_{OG}$ . In order to determine the tray efficiency from point efficiency, it is necessary to know the relationship between those two efficiencies. This relationship depends on the concentration gradient which is determined by the degree of liquid mixing on the tray. On the other hand, a knowledge of the liquid mixing on a tray is also important in interpreting the observed tray efficiency. Without it, it is impossible to determine the relationship between the observed and predicted efficiencies or to compare different results in fundamental terms of point efficiency.

If the liquid on the tray is completely mixed, the point efficiency and tray efficiency are equal, i.e.,

$$E_{MV} = E_{OG} \quad (1)$$

If the vapor is completely mixed and the tray liquid is completely unmixed, or flows from the inlet to the outlet edge of the tray without back-mixing, the liquid flow is called “plug flow” and the relationship between the point efficiency and tray efficiency is related by the following Equation:

$$E_{MV-plug} = \frac{e^{\lambda E_{OG}} - 1}{\lambda} \quad (2)$$

Generally, an actual tray will have a degree of liquid back-mixing which is somewhere between these two extreme cases above. The liquid back-mixing can be described in terms of the mixing pool model. The tray in terms of the mixing pool model can be divided into a number of completely mixed pools. The concentration gradient across the tray is assumed to be made of a series of pools with each mixing pool having a uniform composition. For a very large number of very small pools, the concentration gradient becomes continuous and plug flow exists. Conversely, for a tray that is completely mixed, there is no concentration gradient and the entire tray has a uniform composition. These are two limiting conditions that may not be realized in practice. The effect of

the degree of mixing indicated by the number of mixing poolson the relationship between  $E_{OG}$  and  $E_{MV}$  as a function of the  $\lambda$ , or  $mV/L$ , is given in **Equation (3)** and shown in **Figure 1**.

$$\frac{E_{MV}}{E_{OG}} = \frac{1}{\lambda E_{OG}} \left[ \left( 1 + \frac{\lambda E_{OG}}{n} \right)^n - 1 \right] \quad (3)$$

It can be seen from this figure that liquid back-mixing, or lack of it, can have a profound effect on the ratio of  $E_{MV}$  to  $E_{OG}$ .

The liquid back-mixing on distillation trays has been studied by numerous researchers. Stone<sup>1</sup>, Brown<sup>2</sup> and Wharton<sup>3</sup>, have measured liquid back-mixing using a continuous salt tracer to measure the concentration at several points downstream from the injection point. The air-water system was studied at low liquid flow rates on rectangular trays 0.305 m wide by 1.52 m long. Both bubble cap and sieve trays were investigated. Crozier<sup>4</sup> measured liquid back-mixing on a single tray by monitoring, with a colorimeter, the decay of a steady tracer of dye that was suddenly stopped. Miller<sup>5</sup> determined the liquid back-mixing by measuring the concentration gradients on the tray. Johnson and Marangozis<sup>6</sup> assumed splashing to be the major cause of mixing and measured splashing on a sieve tray by actually catching the spray in a trough at several distances from the center of the tray. Foss, Gerster and Pigford<sup>7</sup> measured mixing by determining the increase in variance of an interrupted salt tracer on several different sieve trays. Robinson<sup>8</sup> measured mixing with a salt tracer at fairly low liquid flow rates on a sieve tray and a bubble cap tray. Gilbert<sup>9</sup> measured mixing with a sieve tray and a bubble cap tray. This was done by injecting a sodium thiosulphate tracer in a sine wave of different frequencies and measuring the attenuation of the sine wave at the outlet weir with conductivity cells. All of these studies were conducted in an air-water system at atmospheric pressure.

The degree of liquid back-mixing can be described by a number of models. Each model in turn may use a different parameter to predict the amount of back-mixing. Two commonly used models are a mixing pool model and an eddy diffusivity model. Kirschbaum<sup>10</sup> first came to a concept of mixing pools to describe the liquid mixing process on trays. It is assumed that a tray can be divided into a number of completely mixed pools. The concentration gradient across the tray is

assumed to be made up of steps with each mixing pool having a uniform composition. If there is a large number of very small pools, the concentration gradient becomes continuous and a plug flow exists. If the entire tray is one perfectly mixed pool, there is no concentration gradient and the tray is considered completely mixed. A mixing pool model is easy to visualize, but hard to integrate with the diffusion theories. Another way to describe the degree of mixing is eddy diffusivity,  $D_e$ . The first and best known eddy diffusivity model is the one published by AIChE<sup>11</sup>. This model is simply a correlation of the data without any theory or explanation to support it. Zuiderweg<sup>12</sup> tried to use different correlations for different flow regimes and account for the effect of liquid and vapor densities on liquid back-mixing. Bennett and Grimm<sup>13</sup> developed a liquid-phase eddy diffusivity correlation on sieve trays by modeling back-mixing as the trajectory motion of liquid droplet elements through the two-phase layer height on the tray. In the paper published by Bennett, Watson and Wiecek<sup>14</sup> in 1997, it is recommended that for a tray with curvature, for example a cross-flow tray, the value of eddy diffusivity derived from the Bennett and Grimm correlation be multiplied by a value of four to represent the overall mixing occurred on a tray.

This paper presents experimental studies of liquid back-mixing on distillation trays by Fractionation Research, Inc. (F.R.I.). The studies were conducted in a 1.22m diameter column on various trays by injecting a pulse of non-volatile dye into the liquid stream of a tray in the F.R.I. distillation unit. The degree of liquid mixing on the tray is determined as a function of the change in the shape of the pulse as it passes from tray to tray. The test results are compared to predictions from several literature models.

## **EXPERIMENTAL SETUP AND DATA REDUCTION**

**Figure 2** shows the F.R.I. experimental unit that consists of two commercial size distillation columns and their support systems. For most operation modes only one column is used. The 1.22m inside diameter high pressure column is 8.4m tall from the bottom headseam to the top flange and rated for pressures up to 37 bar. The low pressure column is rated from deep vacuum to 11.4 bar and consists of two sections. The lower section is essentially identical to the high pressure column but topped with a 3.66m tall transition zone. The upper section has a 2.44m inside diameter and is 6.7m

tall. Each column has a flanged head and clean inner wall design which allows installation of hardware at any location in the column. Sight windows are strategically located to provide visual observation points inside the column. Couplings are available every 152mm along the shell, which permits temperatures and pressures to be measured and samples to be withdrawn. The high pressure column was used for this study.

F.R.I. studied liquid back-mixing on both bubble cap trays and sieve trays. **Figure 3** shows a schematic drawing of the dye injection equipment and measuring apparatus. A pulse of non-volatile dye was injected into the reflux line of the experimental column in the F.R.I. distillation unit. Three trays, 3, 4 and 8, were sampled by continuously withdrawing a stream from the tray through a 6.35mm copper sample line to a continuously recording colorimeter. The continuous-flow explosion-proof colorimeters monitored the optical density of a sample stream taken from the three trays. The signal from these colorimeters was amplified and recorded by a multi-channel oscillograph. **Figure 4** is an actual dye injection recording taken during a typical run. The source light for one of the colorimeter cells is turned off automatically as a charge of dye is injected into the column and this is indicated by a pulse. The recorded parameter is the percent transmittance of light through the colorimeter cell. An oil soluble dye was used for the hydrocarbon systems. 70 grams of Du-Pont Oil Red powder was mixed with 5 gallons of hydrocarbon. The saturated dye solution was allowed to settle and was filtered before use. The dye was tested for color stability at temperatures above the operating conditions for several hours. The dyes were found to have a substantial color stability safety factor for the tests conducted.

The degree of liquid back-mixing on the tray was determined as a function of the change in the shape of the pulse as it passed from tray to tray. The curves obtained from the oscillograph can be characterized by a statistical mean and a variance as shown in **Figure 5**. The difference between the statistical means of the curves is approximately equal to the residence time on different trays. The dye concentration curve spread out as the dye moved down the column. This effect is due to liquid back-mixing. A measure of the "spread" of the curve is the variance. During the tests, dye

concentration was measured at one half second intervals. The mean,  $\theta$ , and variance,  $S^2$ , are determined by numerical integration as follows:

$$\theta = \frac{\int Ctdt}{\int Cdt}$$

$$S^2 = \frac{\int Ct^2dt}{\int Cdt} - \theta^2 \quad (4)$$

The mixing and delay in the sampling system were subtracted and corrected from the colorimeter cell measurement data. After the corrected  $\theta$  and  $S^2$  are obtained, the number of mixing pools,  $N$ , on a tray can be determined as a function of the time taken for the mean of the dye concentration-time curve to cross one tray and the change in variance of the curve caused by the tray action by the following equation (Foss et al., 1958):

$$N = \frac{(\Delta\theta)^2}{\Delta S^2} \quad (5)$$

- where  $N$  = the number of mixing pools on one tray
- $\Delta\theta$  = the time taken for the mean of the dye concentration-time curve to cross one tray
- $\Delta S^2$  = the change in variance of the dye concentration-time curve caused by the tray

For illustration only, the following tables show how to obtain the number of mixing pools on one tray from the measurement of the oscillograph for an individual test run. **Table I** shows how to determine the mean and variance from the dye concentration-time curve.

**Table I**

	Obtained from numerical integration			Calculated	
	$\int Ctdt$	$\int Ct^2dt$	$\int Cdt$	$\theta$	$S^2$
<b>Tray 3</b>	1240.09	29,150.7	54.75	22.65	19.41
<b>Tray 4</b>	1164.05	25,007.5	56.75	20.60	16.30
<b>Tray 8</b>	806.37	12,153.9	55.01	14.66	6.05

The values obtained in **Table I** only indicate the mean and variance of the dye concentration-time curve in the colorimeter cells. The parameters of interest are the mean and variance values from the curve on the tray only. The difference is the increase in these values due to the mixing and delay in the sampling system that is decided by tests. The effect of the sampling system must be subtracted as shown in **Table II**.

**Table II**

	Within colorimeter cells		Correction		Values on tray	
	$\theta$	$S^2$	$\theta$	$S^2$	$\theta$	$S^2$
<b>Tray 3</b>	22.65	19.41	2.05	1.10	20.60	18.31
<b>Tray 4</b>	20.60	16.30	2.05	1.10	18.55	15.20
<b>Tray 8</b>	14.66	6.05	2.70	1.40	11.96	4.65

The number of mixing pools,  $N$ , on a tray can be determined by **Equation (5)** as shown in

**Table III:**

**Table III**

	$\theta$	$\Delta\theta_{8-3}$	$\Delta\theta_{8-4}$	$\frac{\Delta\theta}{tray}$	$\left(\frac{\Delta\theta}{tray}\right)^2$	$S^2$	$\Delta S_{8-3}^2$	$\Delta S_{8-4}^2$	$\frac{\Delta S^2}{tray}$	$N$
<b>Tray 3</b>	20.60	8.64		1.72	2.95	18.31	13.66		2.73	1.08
<b>Tray 4</b>	18.55		6.59	1.65	2.72	15.20		10.55	2.64	1.03
<b>Tray 8</b>	11.96					4.65				

From the measured number of mixing pools,  $N$ , the eddy diffusivity,  $D_e$ , can be obtained from the following equation by Ashley and Haselden<sup>15</sup>:

$$D_E = \frac{Q_L \cdot l_{fp}}{2h_1 \cdot (N - 1)} \quad (6)$$

where  $l_{fp}$  = length liquid flow path, m  
 $Q_L$  = liquid flow rate per unit weir length, m<sup>3</sup>/s-m  
 $h_l$  = clear liquid hold-up, m

Liquid hold up,  $h_l$ , can be obtained from the model developed by Colwell<sup>16</sup>. The model is based on the Francis weir formula as follows:

$$h_l = \alpha h_w + 0.73 \left( \frac{\alpha^{0.5} Q}{C_d l_w} \right)^{2/3} \quad (7)$$

where  $h_w$  = height of the outlet weir, m  
 $\alpha$  = liquid volume fraction of the two-phase dispersion  
 $Q$  = liquid flow rate, m<sup>3</sup>/s  
 $l_w$  = length of the outlet weir, m

If  $\frac{h_f - h_w}{h_w} \leq 8.135$ , then

$$C_d = 0.61 + 0.08 \frac{h_f - h_w}{h_w}$$

otherwise,

$$C_d = 1.06 \left( 1 + \frac{h_f - h_w}{h_w} \right)^{1.5}$$

And liquid volume fraction,  $\alpha$ , is calculated by the following equations:

$$\alpha = \frac{1}{1 + 12.6 Fr_b^{0.4} \left( \frac{A_h}{A_b} \right)^{-0.25}}$$

where  $Fr_b = \left( \frac{\rho_v}{\rho_L - \rho_v} \right) \left( \frac{u_b^2}{gh_l} \right)$

$u_b$  = vapor velocity based on bubbling area, m/s

$A_h$  = hole area, m<sup>2</sup>

$$A_b = \text{bubbling area, m}^2$$

## RESULTS

Liquid back-mixing has been determined on bubble cap trays and sieve trays using cyclohexane/n-heptane (C6/C7) and i-butane/n-butane (iC4/nC4) systems. Only a selection of the results obtained is presented in the following figures. The measured liquid back-mixing results are also compared to the predictions of the models published in the literature.

**Figure 6** compares the liquid back-mixing occurring on a bubble cap tray and a sieve tray with the C6/C7 system at 1.65 bar and 0.013 m<sup>3</sup>/s-m outlet weir liquid load. Except for the difference of the tray decks, the other dimensions of those two trays are very similar if not the same. As shown in the figure, tray design has a very significant effect on the amount of liquid mixing occurring on a tray. The eddy diffusivity on the sieve tray is consistently less than that on the bubble cap tray. This means that the sieve tray exhibited more plug-flow than the bubble cap tray. The absence of hardware on sieve trays appears to have less liquid back-mixing or more plug-flow as compared to the bubble cap tray. The test results from trays with different hardware configurations in the bubbling area show that the degree of the back-mixing varies with tray deck designs.

**Figure 7** shows the effect of system properties on the liquid back-mixing. The results in this figure were obtained using an 8 percent sieve tray with same weir liquid load but with three different test systems, C6/C7 at 0.34 and 1.65 bar and iC4/nC4 at 11.4 bar. It is indicated in the figure that system properties have a profound effect on the back-mixing occurring on the tray. The C4 system has less back-mixing or more plug-flow than the C6/C7 system. For the C6/C7 system, the experimental results from the 1.65 bar test exhibits lower eddy diffusivity or less back-mixing than the 0.34 bar test. Similar results were obtained with bubble cap trays.

The effect of the flow path length (size of bubbling area) on liquid back-mixing is shown in **Figure 8**. It compares the liquid back-mixing of a sieve tray with a liquid flow path length of 0.64m with a sieve tray having a flow path length of 0.97m. The test system used is the C6/C7 system at 1.65 bar. This figure compares the mixing of a sieve tray with a bubbling area of 90 percent of the

tower area and 60 percent of the tower area. It can be seen from the figure that the tray with the longer and larger bubbling area (smaller downcomer size) has similar liquid back-mixing with the tray with the shorter and smaller bubbling area (larger downcomer area). This figure also indirectly indicates that the liquid back-mixing occurring in the tray deck is much larger than that in the downcomer. Similar effects were also found in tests with bubble cap trays.

## DISCUSSION

To predict the liquid back-mixing (eddy diffusivity) on distillation trays, several models were published in the literature. The first and best-known correlation for eddy diffusivity was published by AIChE in 1958 in the Bubble Tray Design Manual<sup>11</sup>. This model is simply a correlation based on experimental data. Zuiderweg<sup>12</sup> used different eddy diffusivity correlations for the spray/mixed flow regime and the emulsion flow regime. The criteria for separating these two regimes are also presented in his paper. Unlike the AIChE and Zuiderweg models, Bennett et al<sup>13</sup> assumed that the liquid back-mixing is caused primarily by the droplet motion mechanism. Based on that assumption and steady-state dispersion data from rectangular trays, they developed a so-called Free Trajectory Model (FTM) to predict the eddy diffusivity. In the subsequent study published by Bennett, Watson and Wicinski<sup>14</sup> in 1997, it is recommended that for a tray with curvature, for example a cross-flow tray, the value of eddy diffusivity derived from the Bennett and Grim correlation be multiplied by a value of four to represent the overall mixing occurring on a tray. The details of those models can be found in their papers. **Figure 9** shows the comparison between F.R.I. experimental results and the model predictions. As indicated in the figure, the AIChE model consistently predicts the liquid flow on the trays as plug-flow, which is substantially different from the measured results and observations. As expected, the predicted eddy diffusivity from Bennett et al., published in 1991; is substantially less than the measured overall mixing results, which means the overall mixing on a cross-flow tray is substantially larger than that due to the simple eddy diffusivity predicted by the Bennett et al. correlation. **Figure 10** compares the measured overall mixing results with the predictions from Zuiderweg model and Bennett et al model published in 1997. As shown in the figure, the model predictions agree reasonably well with the experimental results except the predictions are scattered. The discrepancy between the predictions of the literature models and the F.R.I. experimental results is

possibly attributed to several reasons. As shown in the experiment set-up, the measured liquid back-mixing results presented in this paper are the overall mixing occurring in both the tray deck and tray downcomer; however, the model predictions do not include the mixing in the downcomer. On the other hand, since the measured data and the model prediction do not show obvious or distinguishable bias, it indicates that the mixing in the downcomer is less than that occurring on the tray deck for the trays tested. That is consistent with the results shown in **Figure 8** that indicates that the liquid back-mixing occurring in the tray with large downcomer is similar to that in the tray with much smaller downcomer. The other sources of the discrepancies are that literature models are based on test data from small or rectangular columns; and most of the literature data were collected from the tests with air water system. Finally, there are limitations on the Foss's method to derive the liquid back-mixing from the dye injection since the **Equation (5)** was based on the assumption of a uniform velocity pattern and constant eddy diffusivity throughout the tray deck. Some of the discrepancies may be caused by this one-dimensional data analysis model.

## **CONCLUSIONS**

Liquid back-mixing for both bubble cap and sieve trays was determined using different distillation systems at various pressures. The results show that the liquid back-mixing varies with tray types and tray designs. Less back-mixing is found for high pressure, high liquid rate systems. The liquid back-mixing occurring in the tray deck is much larger than that in the downcomer. The measured data agree reasonably well with the predictions from Bennett et al. and Zuideweg correlations.

## ACKNOWLEDGEMENT

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## NOMENCLATURE

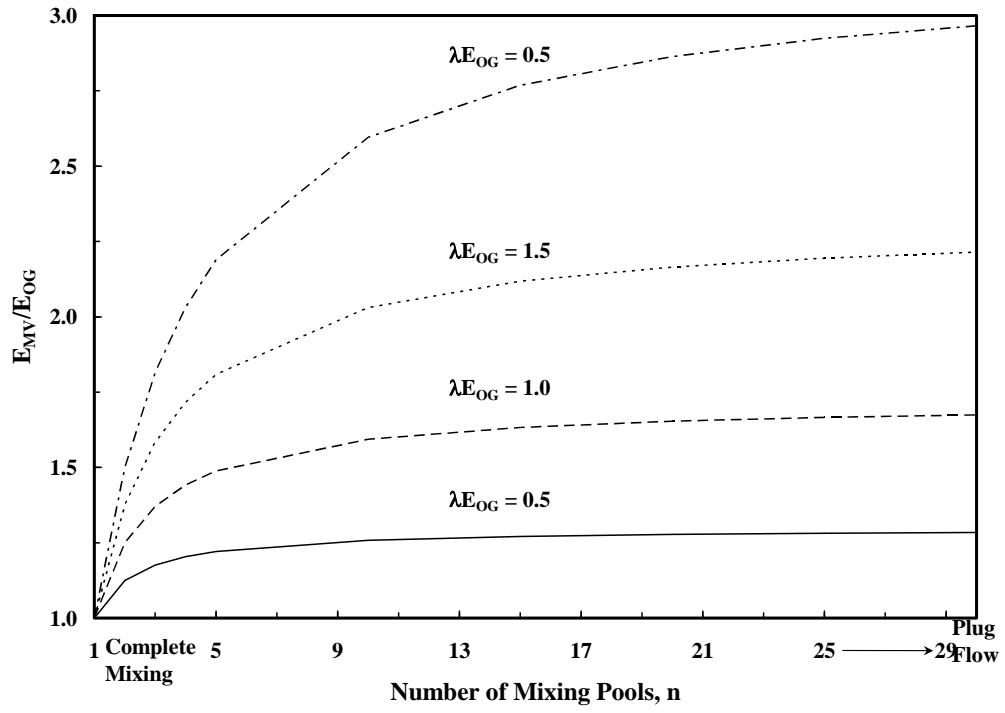
$A_b$	=	Bubbling area, $m^2$
$A_h$	=	Hole area, $m^2$
$AT$	=	Column cross section area, $m^2$
$C$	=	Dye concentration
$D_E$	=	Eddy diffusivity, $m^2/s$
$E_{MV}$	=	Murphree tray efficiency
$E_O$	=	Overall tray efficiency
$E_{OG}$	=	Point efficiency
$G$	=	Acceleration due to gravity, $m/s^2$
$h_l$	=	two-phase layer height defined in <b>Equations (7)</b>
$h_f$	=	Effective froth height, m
$h_w$	=	Height of the outlet weir, m
$l_{fp}$	=	Length of liquid flow path, m
$l_w$	=	length of outlet weir, m
$N$	=	Number of mixing pools on one tray
$Q$	=	Liquid flow rate, $m^3/s$
$Q_L$	=	Liquid flow rate per unit weir length, $m^3/s-m$
$S$	=	Variance of the dye concentration curve
$U_b$	=	Vapor velocity based on bubbling area, m/s

## Greek Symbols

$\alpha$	=	Froth density
$\beta$	=	Trajectory Angle, degree
$\lambda$	=	Stripping factor
$\rho_L$	=	Liquid density, $kg/m^3$
$\rho_V$	=	Vapor density, $kg/m^3$
$\phi_e$	=	Effective froth density defined in <b>Equation (10)</b>
$\theta$	=	Mean of dye concentration curve

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**Figure 1.** The Effect of Liquid Back-mixing on Tray Efficiency

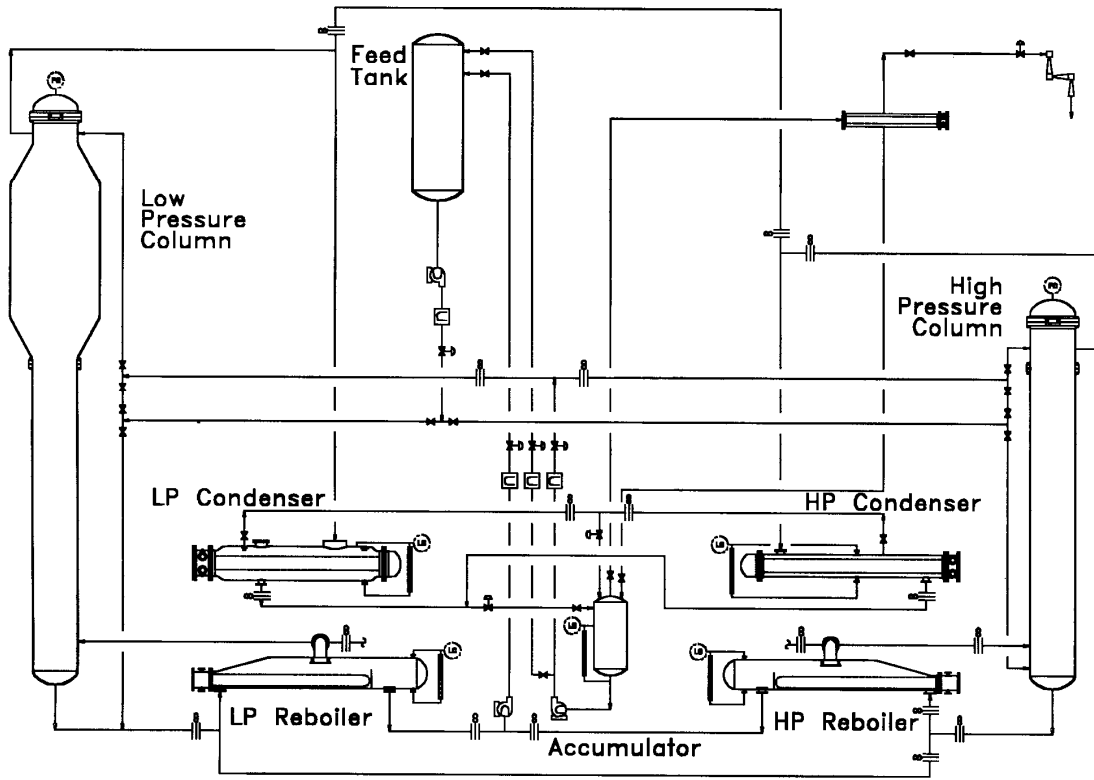
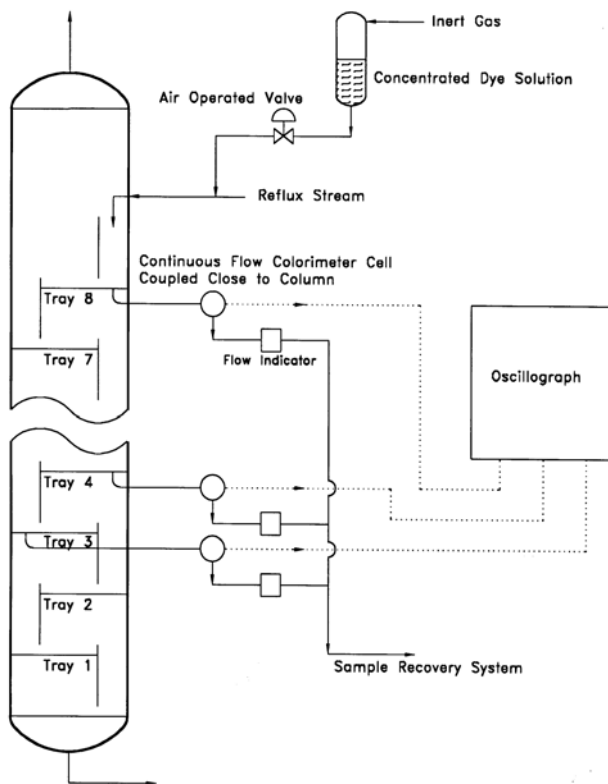


Figure 2. F.R.I. Experimental Unit



**Figure 3.** Schematic of Test Set-Up

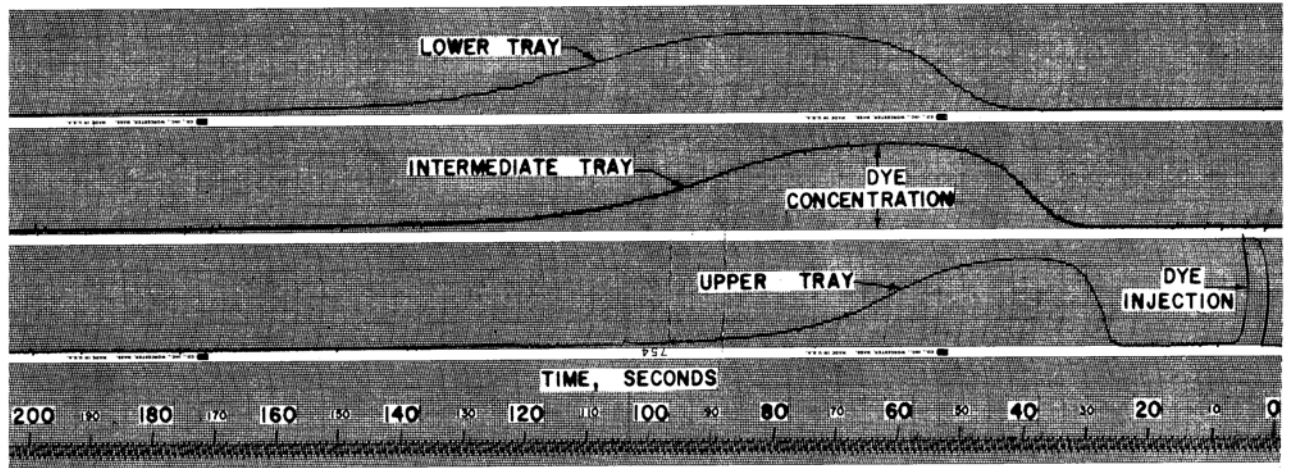
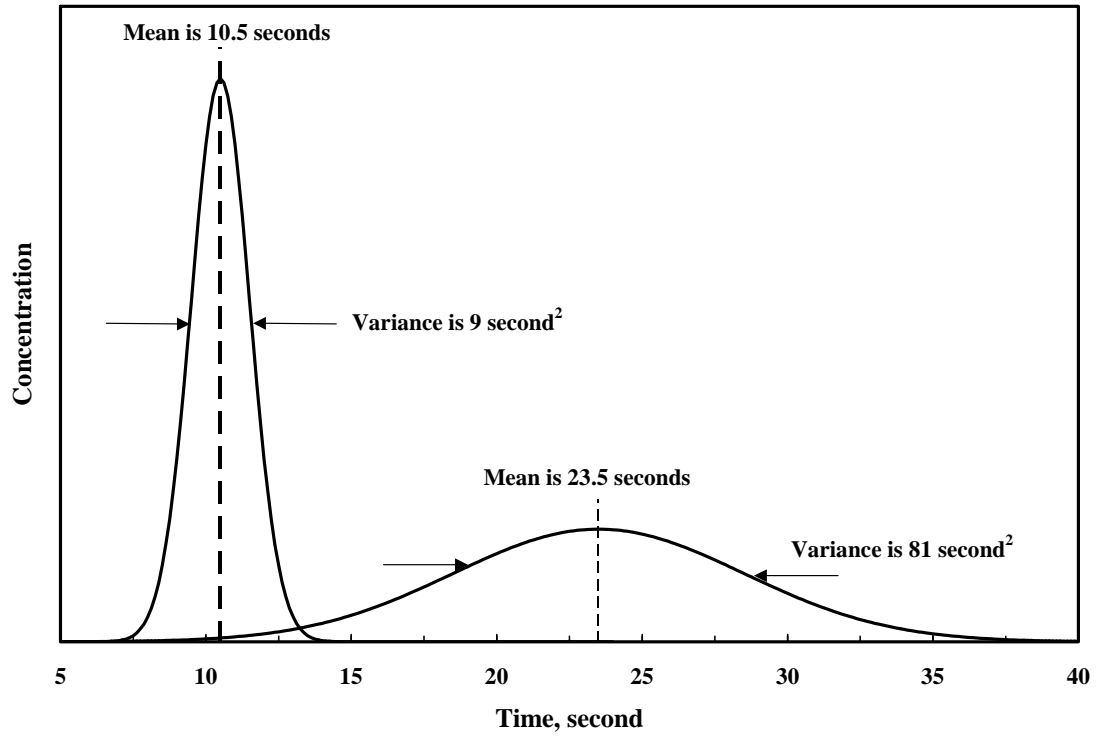
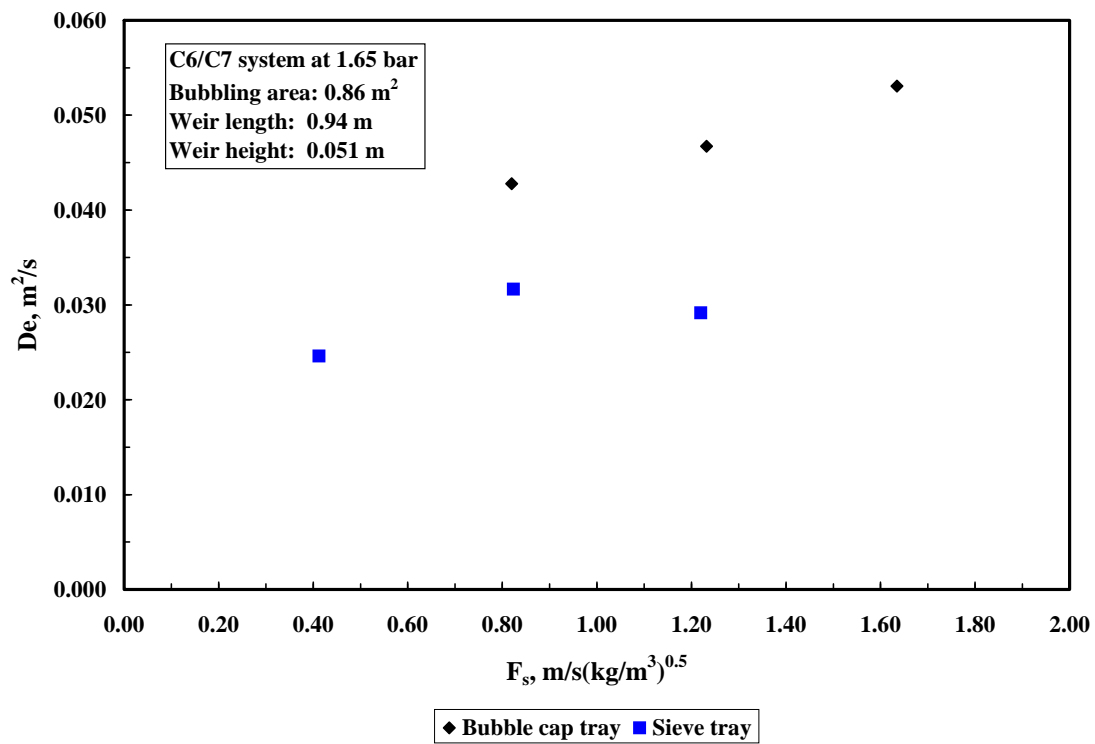


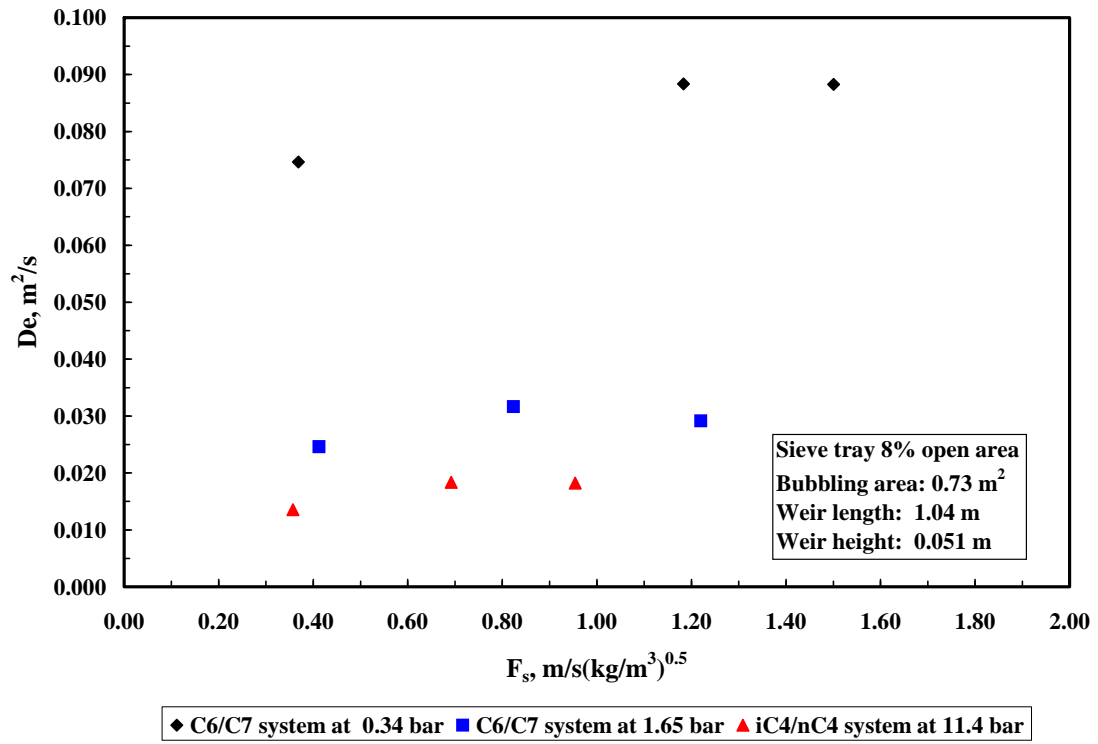
Figure 4. Typical Dye Injection Recording



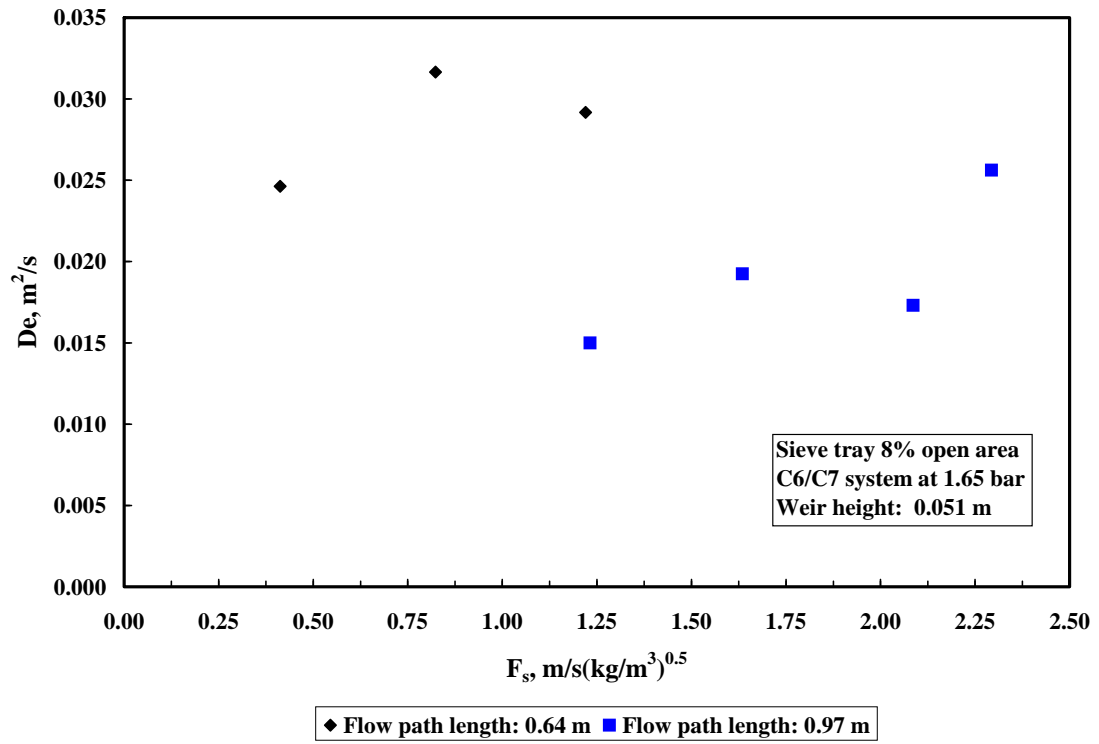
**Figure 5.** Characteristics of Mean and Variance from Smoothed Colorimeter Output



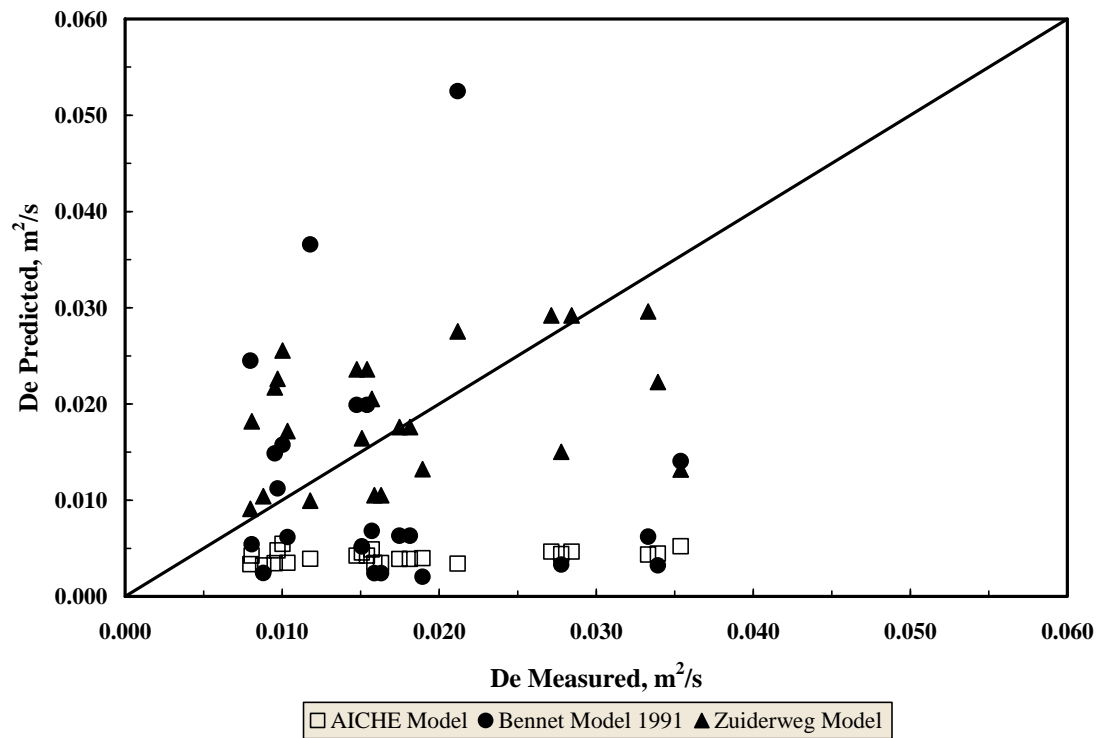
**Figure 6.** Effect of Tray Design on Liquid Back-mixing



**Figure 7.** Effect of System Properties on Liquid Back-mixing



**Figure 8.** Effect of Flow Path Length on Liquid Back-mixing



**Figure 9.** Parity Plot of Eddy Diffusivity

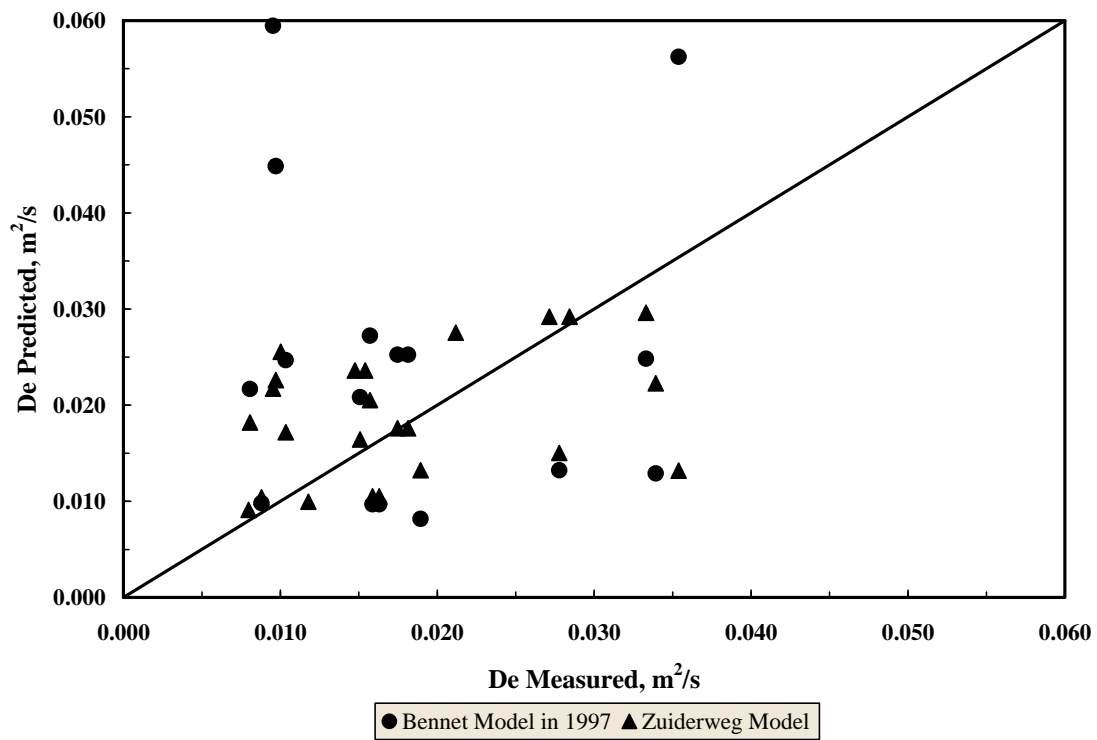


Figure 10. Parity Plot of Eddy Diffusivity

