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Paper 7D

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ABSTRACT

The authors will present the design of a commercial benzene-toluene splitter. This column takes the overhead product stream from a benzene-toluene column and separates it into benzene and toluene product. The column was designed as a packed column. The packing used in the design of the column was structured packing. An overview of the column's design will be detailed in the article. The authors will also present operational data from the benzene-toluene splitter. The data that will be presented includes number of the theoretical stages, temperature and pressure profiles and, product purity. The authors will do a complete comparison between the baseline design of the column and the commercial operating data from the column.

INTRODUCTION

In the styrene dehydrogenation reaction benzene and toluene by-products are produced. These by-products are recovered as a single stream in the benzene-toluene column (C-2). The benzene-toluene column recovers benzene-toluene and lighter material overhead, while the bottom product stream recovers styrene and heavier material. A project was instituted to recover the benzene from the overhead stream. A benzene-toluene splitter (C-4) was designed and installed. This splitter takes the overhead stream from the benzene-toluene column and splits it into two products. The overhead product stream from the benzene-toluene splitter is mostly benzene, which is recycled to an ethylbenzene plant for reprocessing while the bottom product stream is essentially pure toluene. This project not only decreases feedstock consumption, by reducing the fresh benzene demand, but it also produces a high purity toluene product that can be sold (**Figure 1)**.

A benzene-toluene (B/T) splitter packed with structured packing was designed to process 140% of the current Benzene-Toluene Column flow. The column was designed to recover 99.8% benzene by weight overhead and 99.2% toluene by weight in the bottom product. (**Figure 2)** represents a process flow sketch of the B/T Splitter system. The combined benzene/toluene stream from the existing benzene-toluene column flows from the existing bottom section of B/T column condenser/accumulator to exchange T-17. This exchanger raises the feed temperature from 100°F to 320°F. The feed then passes through exchanger T-18 that raises the feed temperature from 320°F to 360°F. Exchanger T-18 has the ability to be by-passed if not needed. The overhead product is sent to the overhead condenser and the condensed product is recycled to the ethylbenzene plant for reprocessing. The toluene product stream is cross-exchanged with exchanger T-17 and is then exchanged with exchanger T-16, which cools the stream down to 110°F. The product stream is sent from exchanger T-16 to a product storage tank.

TOWER DESIGN

When a distillation column is being designed it is necessary to set the values for a complete set of independent variables. The feed variables are normally already known, therefore, it is typically necessary to pick near-optimum values for the reflux ratio, column pressure, column diameter, and the product purity. From this set of independent variables, it is possible to determine the number of theoretical stages needed to achieve the desired separation.

Column Pressure

For this design the controlling factor in choosing the column pressure was the ability to use low-pressure boiler feed water as a condensation medium to produce low-pressure steam. This choice would mean that the column was going to operate under pressure. Performing this separation under pressure had a couple of advantages.

1. Increasing the column pressure would increase the vapor density and therefore the vapor handling capacity. This would lead to a reduction in the diameter of the column, which would reduce the overall cost of the project.

2. It would allow the possibility of having the benzene-toluene splitter share a condenser with a tower used to remove benzene from vent gas. Both columns' overhead products would go to

the same location. The cost of installing a complete condenser system for this column would be considerably reduced.

However, in raising the column's operating pressure there are some unfavorable effects.

1. Raising the pressure lowers the relative volatility and increases the separation difficulty.

2. Raising the pressure also raises the reboiler temperature, thereby requiring a more expensive heating medium. A reboiler with a larger heat transfer area would be required.

3. Above 100-psig pressure, the columns shell thickness increases in order to handle the higher pressures. This will constitute an increase in capital costs.

In examining the system and all the associated costs, a decision was made to design the new column with a shared condenser. Using a shared condenser would significantly reduce the cost of the project. The cost of having to purchase a new condenser, receiving drum, and associated pumps was much higher than the cost of retrofitting the existing condenser system. Therefore, the column's overhead pressure was set at 86 psig, which matches the vent gas column's overhead pressure.

Product Purity

The column was designed to achieve a toluene purity of 200 ppm in the column's overhead product. The benzene in the bottom product stream was set at 5000 ppm. Market considerations dictated the purity specification.

Feed Stream

The benzene-toluene splitter design phase began by creating a simulation model of the tower and the associated equipment. Overhead product from the benzene-toluene column (C-2) was collected and analyzed. After reviewing six months of data and looking at all of the possible operating scenarios a design basis was developed. **Table 1** gives the design basis that was used to design the benzene-toluene splitter.

TABLE 1

Design Parameter	Parameter Range
Feed Rate(LB/Hr)	Normal Case
Feed Temperature(°F)	100
Feed Inlet Pressure(Psia)	135
Inlet Toluene Mass Fraction	.69
Inlet Benzene Mass Fraction	.31

Simulation Model

In order to estimate what the required theoretical stages would be for benzene-toluene separation, calculations were performed using the Gilliland plot to determine the minimum number of stages at minimum reflux. From these calculations, it was determined that 34 theoretical stages were needed to achieve the desired separation. The Fenske equation was used to determine the best-feed stage location. Based on 34 theoretical stages, the best-feed location was determined to be stage 12.

After performing a rigorous computer simulation, it was determined that 40 theoretical stages were required to achieve the desired separation. The rigorous simulation was conducted with a reflux ratio that was 1.3 times the calculated minimum reflux ratio. The column was designed with 40 theoretical stages, which include the condenser and the reboiler. The feed to the column was located on theoretical stage 19.

The design simulation was performed using a commercial simulation package. The design simulation was run using the feed information given in **Table 1**. The thermodynamic package used in the simulation was NRTL. This package was found to give the best results. Cases were run at the three different feed rates. However, for the purposes of this paper we will concentrate on the normal feed rate case. **Table 2** shows the result of the final simulation case for the normal flow rate case.

TABLE 2

Simulation Parameters	Simulation Results
Overhead Product Stream Temperature(°F)	315.7
Overhead Product Stream Pressure(Psig)	89.3
Overhead Product Stream Purity(Toluene)	200 ppm
Reflux Ratio(L/D)	11.98
Bottom Product Stream Temperature(°F)	388
Bottom Product Stream Pressure(Psig)	90.3
Bottom Product Stream Purity(Benzene)	5000 ppm
Bottom Product Stream Purity(Toluene)	99.2%
Reboiler Duty(mm Btu/Hr)	1.545

In order to verify the accuracy of the simulation an x-y diagram was prepared **Figure 3**. This is a most useful graphical technique for analyzing computer simulation results. From the construction of an x-y diagram we can find.

1. Pinched regions - Pinching is readily seen on an x-y diagram.

2. Mislocated feed points - The feed point should be where the q-line intersects the equilibrium curve. This is generally the rule in binary distillation. However, it is not always true in multicomponent distillation. This is why in the design phase a key ratio plot was developed, **Figure 4**. This type of plot is far superior to an x-y diagram for identifying mislocated feeds.
3. Determining if the column is being over refluxed or reboiled - This can recognized by too wide of a gap between the component balance line and the equilibrium curve throughout the column.

4. Identify cases where feed or intermediate heat exchangers are needed.

As **Figure 3** shows there are no pinch regions or any indication of a mislocated feed. **Figure 4** was developed a check to see if the feed was located correctly. As **Figure 4** shows the feed was located at the proper stage.

Column Design

The internal liquid and vapor traffic was obtained from the simulation model of the column. A preliminary packed tower sizing was performed. From these calculations, it was determined that the column needed to be 3 feet in diameter.

The next step of the process was to work with the commercial tray and packing vendors to come up with a solution to minimize the diameter of the column and to get the desired efficiency. Because the column was going to be 3 feet in diameter and most probably smaller, using trays in this column was ruled out.

In order to get the required capacity with the desired efficiency it was decided that using structured packing was the best option. After, the vendor performed hydraulic calculations with several different kinds of packings the decision was made to use surface enhanced structured packing in the column. The structured packing used has a surface area of 95.2 ft^2/ft^3 and a .32" crimp height. This packing would meet our capacity criteria while providing us with high efficiency. The design HETP value was based on literature provided by the tray vendor. After reviewing the vendor data it was decided to use a HETP of 15 inches to set the packed bed heights. Using an HETP value of 15 inches would make the column slightly taller but there would be enough packing height to ensure that the column would make the required separation.

Based on the hydraulic data given in **Table 3** the column diameter was set at 2 feet 4 inches.

Table 3 details the packing hydraulic calculations.

TABLE 3

Packing Hydraulic Parameter	Hydraulic Value
Liquid Loading(gpm/ft^2) Rectifying Section	9.0
Liquid Loading(gpm/ft^2) Stripping Section	14.3
% Capacity Rectifying Section	56%
% Capacity Stripping Section	74%
Pressure Drop Rectifying Section("H_2O/ft)	.117
Pressure Drop Stripping Section("H_2O/ft)	.210
C-Factor Rectifying Section(Ft/Sec)	.142
C-Factor Stripping Section(Ft/Sec)	.173
Calculated Tower Diameter(Inches)	2'-4"

*** All Hydraulic parameters were based on the Maximum column flow rate ***

The physical layout of the column was a three packed bed design, **Figure 5**. The packed bed in the rectifying section of the column was set at a height of 255.6 inches (21'-.3"). The two beds in the stripping are 170.4 inches high (14'-.2"). The packed bed heights were held between 14 and 20 feet high in order to minimize efficiency loss due to excessive bed height. The effect of bed depth on packing HETP is attributed to liquid maldistribution. Uneven liquid flow generates an uneven concentration profile and localized pinching at the bottom of the bed. The recommended published criteria suggest that it is best to redistribute liquid every 20 feet and to have no more than 10 theoretical stages per bed.

Pan distributors were used because of the diameter of the column. With a column diameter of 2 feet 4 inches using a trough distributor is not practical. In addition, the liquid loading above and below the feed is too high to use trough distributors. The distributors were designed with a drip point density of 10.6-drip points/ft^2. From published information it was determined that a drip point density of 10.6 drip points/ft^2 is sufficient to provide good liquid distribution to each packed bed. Smaller crimp packings require larger drip point densities. 10.6-drip points/ft^2 was the largest drip point density that could be manufactured without making the orifices too small. As a practical matter, 2-3 mm is normally considered the minimum liquid orifice diameter that can be manufactured. The liquid orifice diameters in this column ranged from 5-9 mm.

Operational Performance

A test run on the benzene-toluene splitter was performed. Operational and laboratory data from the column was collected and processed. The results of the test run are given in **Table 4**. A comparison to the simulated values is also given in **Table 4.**

The feed composition going to the column was very similar to the composition of the feed stream used in the simulation model. There was a 50 Lb/Hr difference between the test run flow rates and the design flow rates.

TABLE 4

Simulation Parameters	Operational Data	Simulation Results
Overhead Product Stream Temperature(°F)	318	315.7
Overhead Product Stream Pressure(Psig)	89	89.3
Overhead Product Stream Purity(Toluene)	36 ppm	200 ppm
Reflux Ratio(L/D)	11.99	11.98
Bottom Product Stream Temperature(°F)	390	388.2
Bottom Product Stream Pressure(Psig)	89.77	90.3
Bottom Product Stream Purity(Benzene)	ND	5000 ppm
Bottom Product Stream Purity(Toluene)	99.60%	99.2%
Reboiler Duty(mm Btu/Hr)	1.4899	1.545
HETP	(12.93 Actual)	(15 Design)

***Design and Operational Flow Rates were Similar**

From the information contained in **Table 4** the column is performing better than predicted. The overhead product flow is less than predicted. The column was operated at the design reflux ratio. The column was designed for a bottom benzene concentration of 5000 ppm. Actual plant data indicated there was no benzene in the bottom product. This was extremely important as benzene is considered an impurity when high purity toluene is marketed. Also, all of the recovered benzene goes back into the process, reducing the overall cost of raw materials.

The measured pressure drop in the column is higher than predicted. The column pressure drop as measured is .77 psi or .435 inches H_2O/ft. The calculated pressure drop for the column is .327 inches H_2O/ft. This is a 32% difference in measured pressure drop to calculated pressure drop. The reason for such a discrepancy is the limitations of packed-tower pressure drop correlations. There have been many published accounts of pressure drop correlations that give an excellent statistical fit to experimental data. However, the same correlation can give a poor prediction for applications in industry. Therefore, the difference in predicted pressure drop versus the measured pressure drop of the column was not unexpected.

Conclusions

A simulation model can be used to do process design work on a benzene-toluene column provided that the proper graphical checks are performed to verify the validity of the model. Structured packing is a good option to use for this type of separation. Structured packing gives the option of low-pressure drop and good efficiency. A typical rule of thumb says do not use structure packing in services that operate at 100 psia or better. The data given in **Table 4** has shown that for the benzene-toluene system at 89 psig a high degree of separation can be achieved using structured packing.

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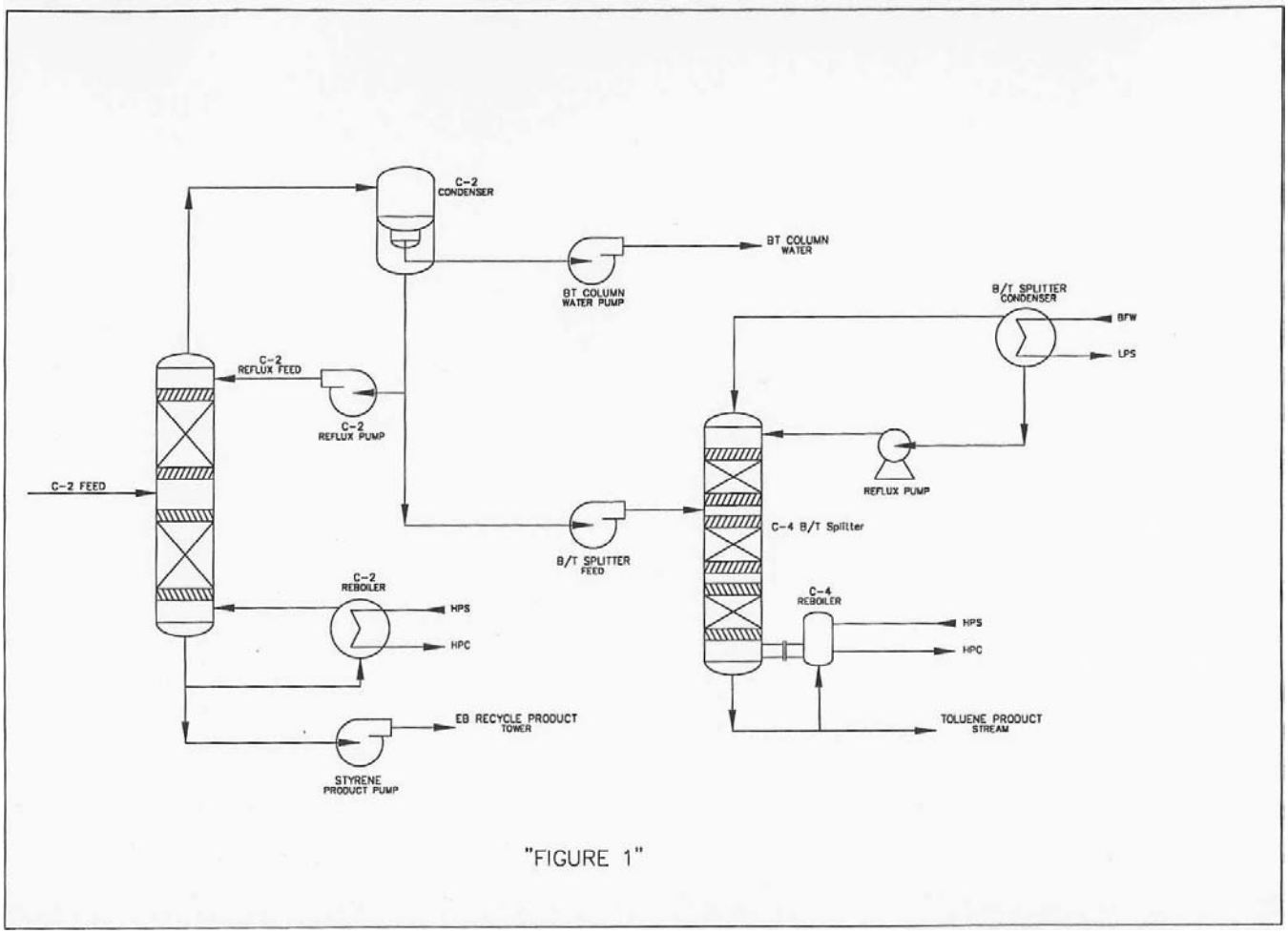
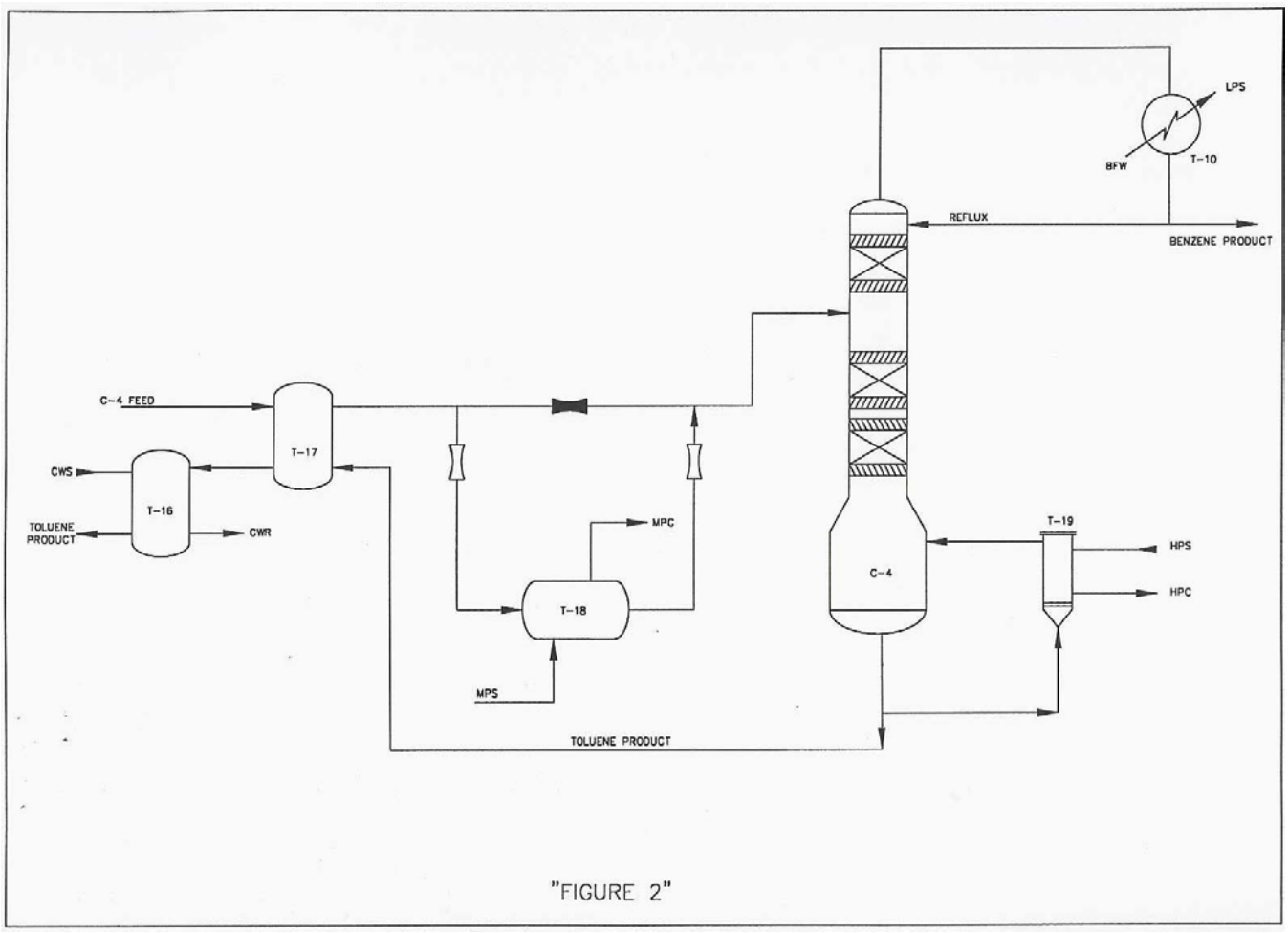
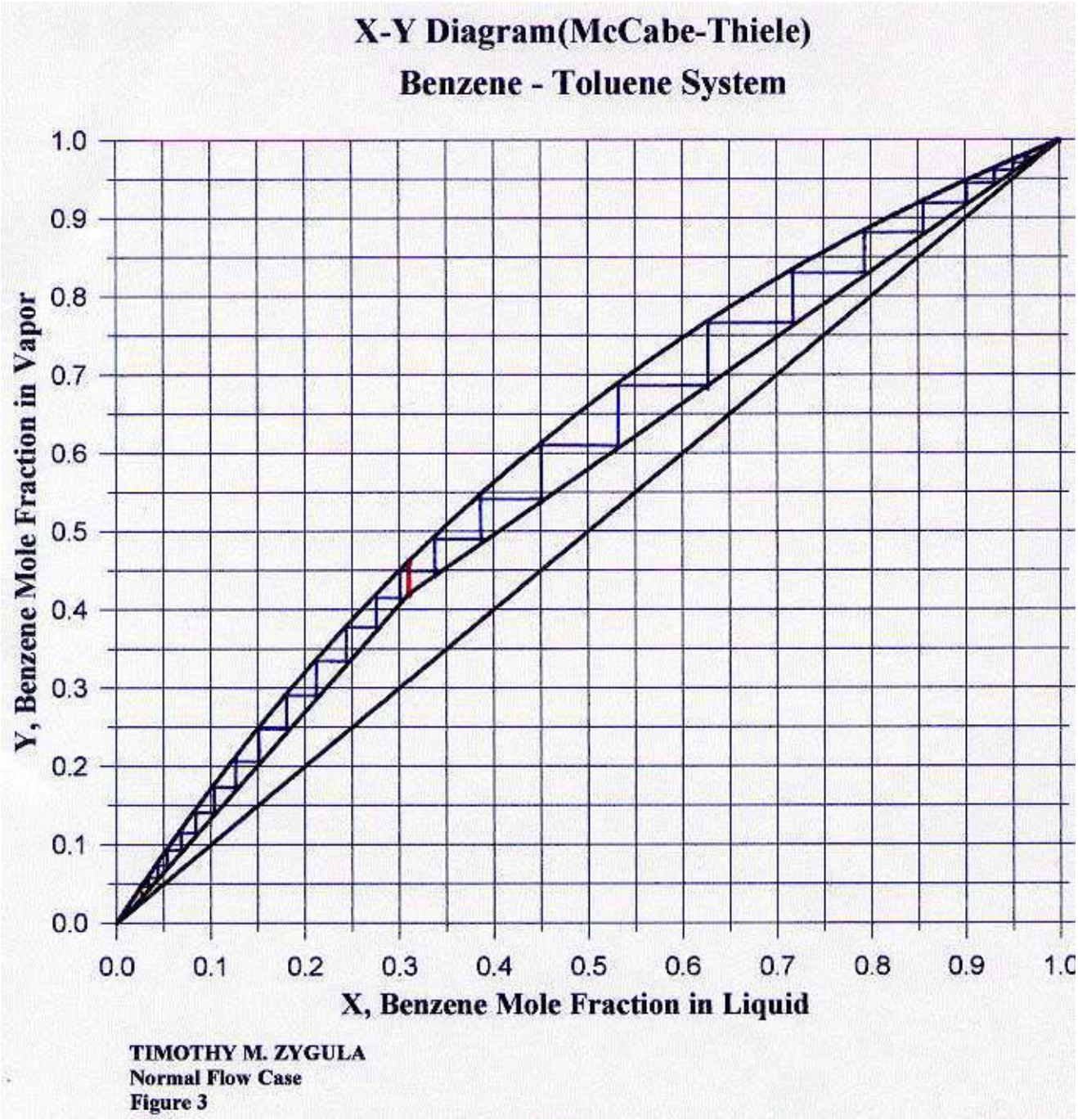
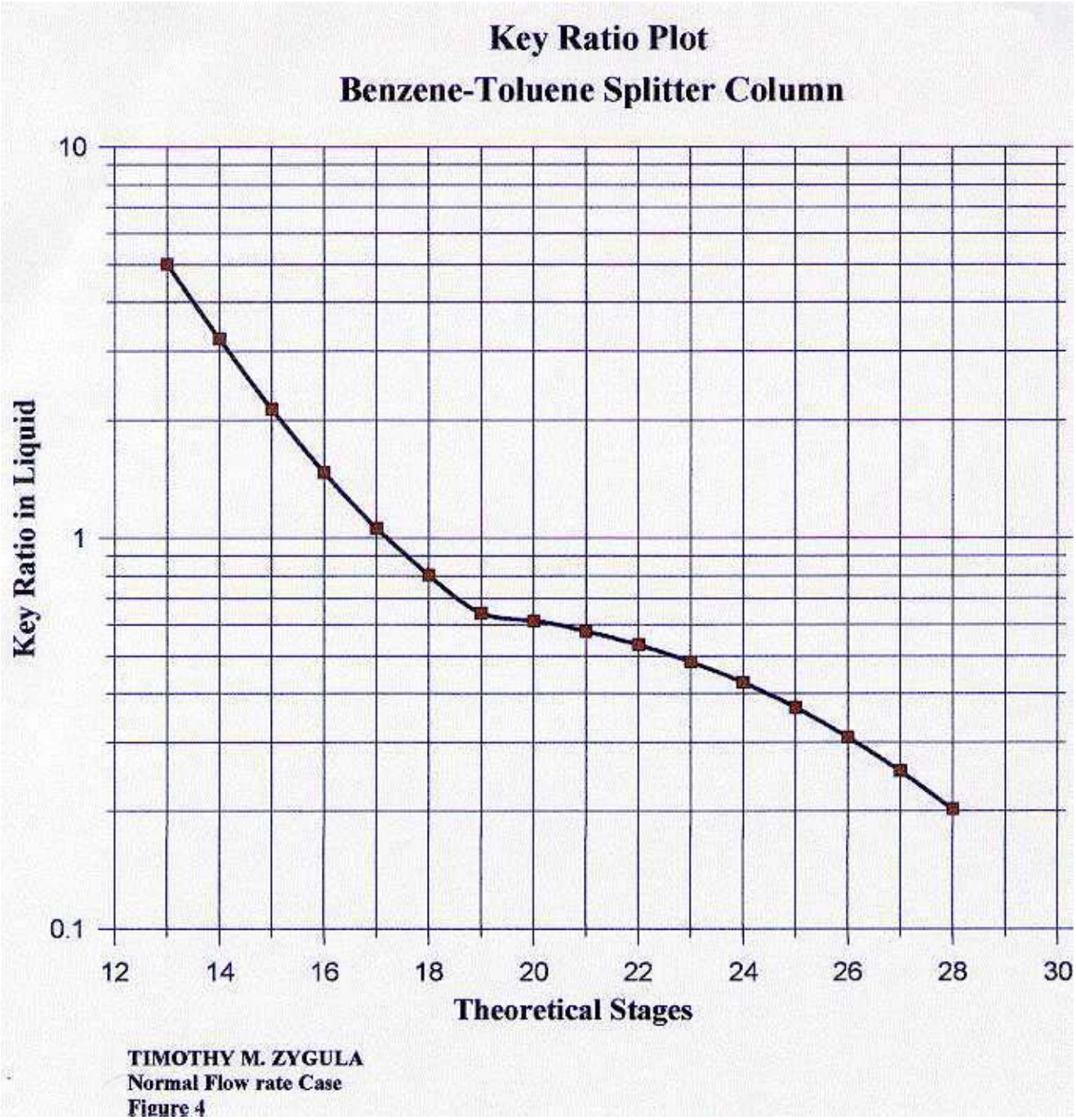

C-2
CONDENSER
BT COLUMN
WATER
BT COLUMN
WATER PUMP
B/T SPLITTER
CONDENSER
BFW
LPS
C-2
REFLUX FEED
C-2
REFLUX PUMP
C-2 FEED
REFLUX PUMP
B/T SPLITTER
FEED
C-4 B/T Splitter
C-2
REBOILER
HPS
HPC
C-4
REBOILER
HPS
HPC
EB RECYCLE PRODUCT
TOWER
TOLUENE PRODUCT
STREAM
STYRENE
PRODUCT PUMP

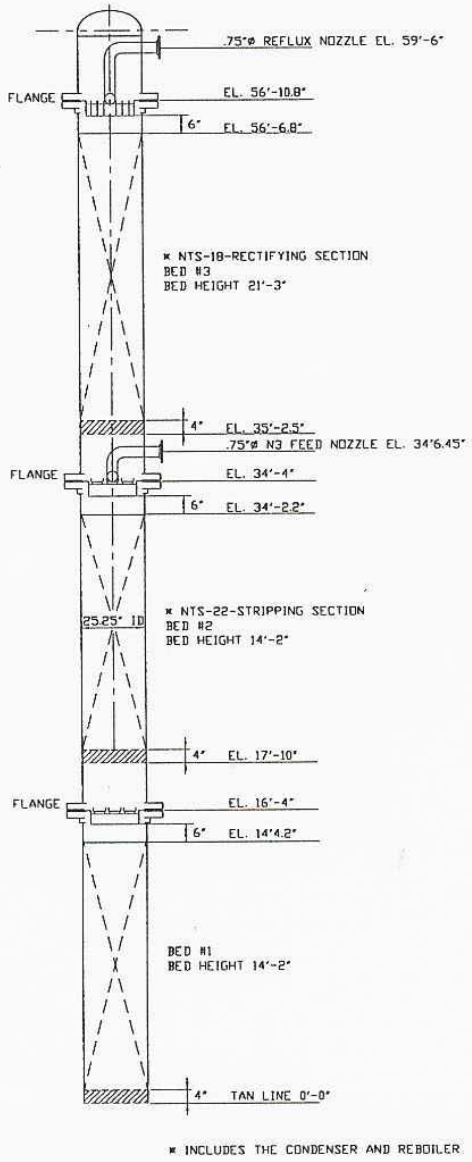
"FIGURE 1"


LPS
BFW
T-10
REFLUX
BENZENE PRODUCT
C-4 FEED
T-17
CWS
T-16
TOLUENE PRODUCT
CWR
MPC
T-18
MPS
C-4
T-19
HPS
HPC
TOLUENE PRODUCT

"FIGURE 2"


X-Y Diagram(McCabe-Thiele)
Benzene - Toluene System
Y, Benzene Mole Fraction in Vapor
X, Benzene Mole Fraction in Liquid
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
0.0
0.0
0.1
0.2
0.3
0.4
0.5
0.6
0.7
0.8
0.9
1.0
TIMOTHY M. ZYGULA
Normal Flow Case
Figure 3


Key Ratio Plot
Benzene-Toluene Splitter Column
10
1
0.1
Key Ratio in Liquid
12
14
16
18
20
22
24
26
28
30
Theoretical Stages
TIMOTHY M. ZYGULA
Normal Flow rate Case
Figure 4



C-4 LAYOUT SKETCH

"FIGURE 5"