COMPARISON OF ADVANCED DISTILLATION CONTROL METHODS

Fourth Annual Report

by

James B. Riggs

September 1998

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PREFACE

Work supported by the U.S. Department of Energy, Assistant Secretary for Energy Efficiency and Renewable Energy, Office of Industrial Technologies, under DOE Albuquerque Operations Office Cooperative Agreement DE-FC04-94AL98747.

This report documents the technical progress from April 1997 through March 1998, and is the fourth annual progress report for the project. This report and the three previous reports can be obtained as indicated by the notice inside the front inside cover. The descriptions of the previous reports are similar to this report except, the report numbers are DOE/AL/98747-1(DE98000103) DOE/AL/98747-2, and DOE/AL/98747-3.

ABSTRACT

Detailed dynamic simulations of three industrial columns (a propylene/propane splitter, a xylene/toluene column, and a depropanizer) have been used to evaluate configuration selection for single-ended and dual-composition control as well as compare conventional and advanced control approaches. For each case considered, the controllers were tuned by using setpoint changes and tested using feed composition upsets.

Proportional Integral (PI) control performance was used to evaluate the configuration selection problem. For single ended control, the energy balance configuration was found to yield the best performance. For dual composition control, nine configurations were considered. It was determined that in order to identify the optimum configuration, detailed testing using dynamic simulation is required.

The optimum configurations were used to evaluate the control performance of conventional PI controllers, DMC (Dynamic Matrix Control), PMBC (Process Model Based Control), and ANN (Artificial Neural Networks) control. It was determined that DMC works best when one product is much more important than the other while PI was superior when both products were equally important. PMBC and ANN were not found to offer significant advantages over PI and DMC.

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INTRODUCTION

Distillation in the refining and chemical industries consumes 3% of the total U.S. energy usage (Humphrey et al, 1991) which amounts to approximately 2.4 quad of energy annually. In addition, distillation columns usually determine the quality of final products and many times determine the maximum production rates.

Unfortunately, many times industry over-refluxes their columns in order to insure that the product purity specifications are met. That is, they use more energy than necessary to meet the product specifications. As a result, industry many times uses 30 to 50% more energy than necessary to produce their products. It has been estimated that an overall average 15% reduction of distillation energy consumption could be attained if better column controls were applied (Humphrey et al, 1991).

While there are many options for applying conventional and advanced distillation controls, industry does not know how to compare the various options. As a result, whether or not to apply advanced distillation control, what type of advanced control to apply, and how to apply it are usually determined based upon internal company politics and hear say. In fact, when industry discusses advanced control, they refer to taking a "leap-of-faith." Because it is not understood, it may be applied where it is not needed, or not applied where it should be applied. When improvements in distillation control performance are obtained, there is a tendency for industry to be satisfied not realizing that further improvements in control may be even more economically important. The bottom line is that industry does not have a consistent basis with which to compare the various options for distillation control.

CASE STUDIES AND DYNAMIC MODELS

Three different columns are studied: a propylene/propane column (a C_3 splitter), a xylene/toluene column and a depropanizer. The specifications for each column are listed in Table 1. These cases represent a wide range of distillation applications. The C_3 splitter is a low relative volatility, high reflux ratio binary column that is so sluggish that typical analyzer delays do not significantly affect feedback control. While the C_3 splitter is a high-pressure column, the xylene/toluene column is a vacuum column for which detailed dynamic pressure modeling is required (Choe and Luyben, 1987). Moreover, these columns should be similar to a large number of industrial columns. For each case, a detailed dynamic simulator was developed. Table 2 lists a summary of the assumptions used and the factors considered for each dynamic column model.

The vapor/liquid equilibrium (VLE) description for each case study was different. For the C_3 splitter, the VLE was described using a relative volatility which was an explicit function of pressure and composition (Hill, 1959). As a result, each tray has its own relative volatility which varied from 1.10 at the top to 1.19 at the bottom for the base case. For the xylene/toluene column, Raoult's law (Prausnitz, 1969) was used for the VLE calculations where the pure component vapor pressures were empirically modeled using the Antoine equation. The resulting relative volatilities

were observed to vary from 2.4 to 3.0 from the bottom to the top of the column. The VLE for the depropanizer was modeled using the Soave-Redlick-Kwong model (SRK; Soave, 1972) for the component K-values. Because the SRK method requires an iterative solution procedure, an empirical correlation for the K-values (Boston and Sullivan, 1974) was used in order to reduce the computational overhead. The empirical correlation for the K-values was reparameterized using the SRK model every 10 seconds of simulation time, or if a tray temperature changed by more than 1.0°C since the last time it was reparameterized. The relative volatility was observed to range from 1.5 at the top to 1.9 for the bottom.

Each column model assumed that the product composition analyzer had an analyzer delay of five minutes. Tray temperatures for the depropanizer (the 11th and 36th tray from the bottom for the stripping and rectifying sections, respectively) were found to correlate well with product compositions. As a result, tray temperatures were used to estimate the product composition for the depropanizer used the following functional form

$$\ln x = A + B/T$$

where x is the product impurity level for each product, T is the tray temperature, and A and B are empirical constants. The value of A was filtered based upon the previous (x , T) values which come from the product composition analyzer while the value of B was empirically set and remained fixed for all simulations. Tray temperatures were found not to correlate well with product impurity levels for the C₃ splitter and the xylene/toluene column.

C₃ splitter used an Euler integrator (Riggs, 1994) with a step size of 0.3 seconds and 0.2 seconds, respectively. The ratio of simulated time to CPU for a 66 MHz 486 PC using Microscroft FORTRAN 5.1 was 50 for the C₃ splitter and 15 for the depropanizer. As a result of the dynamic modeling of pressure, the xylene/toluene simulator required an implicit integrator, LSODES (Hindmarsh, 1983), and resulted in a simulated-time to CPU-time of 7.

The C₃ splitter was bench-marked against dynamic industrial data for a C₃ splitter using the (L,B) configuration. First, open loop responses from the simulator were used to qualitatively check the model against the industrial data. Next, estimated industrial response times (an eight hour response time for the overhead composition for a step 0.5% change in the reflux rate and a 25 hours response time for the bottom composition for a 1% step change in the bottoms flow rate) were used to set the hydraulic time constants for all the trays. A hydraulic time constraint of three seconds provided the best overall fit. Finally, the xylene/toluene model was found to match the results presented by Choe and Luyben (1987).

IMPLEMENTATION APPROACH FOR EACH CONTROLLER

Conventional Proportional Integral (PI) controls, Dynamic Matrix Control (DMC), nonlinear Process Model Based Control (PMBC), and Artificial Neural Networks (ANN) were applied to the simulators of each column for dual composition control. The PI, nonlinear PMBC controllers, DMC, and ANN controllers were applied using the same configuration and each controller was tuned for setpoint changes. Setpoint changes, using 50% changes in impurity, were chosen for controller tuning in order to provide a consistent tuning procedure that is likely to be robust for a wide range of upsets. All controllers used a 5 min control interval, because new analyzer readings were available every 5 min. All controllers were treated as unconstrained, although the column model did not allow for negative flow rates.

The diagonal PI composition controllers were tuned using Auto Tune Variation tests (ATV; Astrom and Hagglund, 1988) with on-line determination of the overall detuning factor. ATV tests were used to identify the ultimate gain and ultimate period for the overhead and bottoms. The Ziegler-Nichols (Ziegler and Nichols, 1942) PI settings were then calculated. Both controllers were detuned to provide minimum Integral Absolute Error (IAE) for setpoint changes in the overhead product using 50% impurity changes. Detuning was accomplished by dividing both controller gains and multiplying both reset times by the same detuning factor. The diagonal PI controllers were also tuned using pulse tests for the identification of transfer function models followed by the application of the Biggest Log Modulus (BLT) tuning procedure (Luyben, 1986) as a comparison to the ATV tuning procedure. The control performances of the controllers tuned by each procedure were found to be essentially equivalent. Because the ATV test with on-line detuning was easier to implement and is more realistically applied in an industrial setting, it was chosen as our PI tuning procedure.

The DMC controller was provided to us by AspenTech. The step response models for the DMC controllers were developed for each input (z, mole fraction light component in the feed) (f, feed flow rate to the column) (L, reflux flow rate) (B, bottom by-product flow rate)/output (x,y) pair. The output for the high purity products were log transformed in an effort to linearize the overall process behavior; e.g.,

$$y' = \log(1 - y)$$

Where y' is the purity of the overhead product. At least 12 independent step tests were conducted for each input variable. Identification software (DMI provided by AspenTech) was applied to all the step test data in order to develop the step response models for each input/output pair used by the DMC controller. The step response models were supplied to the DMC controller, and the final controller tuning was performed for setpoint changes. Standard tuning of the DMC controller was used, i.e., a control horizon of 30 control intervals and a model prediction horizon of 120 control intervals (the maximum available).

The nonlinear PMBC controller using tray-to-tray models was applied using the approach

presented by Riggs et al., 1993. The control law calculates target setpoints (x_{ss} , y_{ss}) based upon proportional and integral feedback.

$$x_{ss} = x + K_{11}(x_{SP} - x) + K_{21} \int (x_{SP} - x) dt \quad (1)$$

$$y_{ss} = y + K_{12}(y_{SP} - y) + K_{22} \int (y_{SP} - y) dt \quad (2)$$

Then, the values of x_{ss} and y_{ss} are used by the tray-to-tray model to calculate the energy input to the column. Because eqs 1 and 2 can result in values of x_{ss} that are negative and values of y_{ss} that are greater than 1.0, limits are used to restrict the maximum and minimum values of x_{ss} and y_{ss} . See page 9, Nomenclature, for definitions.

An overall material balance was used to calculate the value of the bottoms flow rate. Since the column responds much faster to energy changes (e.g., reflux flow) than to material balance changes (e.g., bottoms flow), the target product compositions (x_{ss} and y_{ss}) are applied by K_{xMB} according to the following equations in an effort to improve the response of the bottom product composition.

$$x_{MB} = x_{SP} + K_{xMB}(x_{ss} - x_{SP}) \quad (3)$$

$$y_{MB} = y_{SP} + K_{yMB}(y_{ss} - y_{SP}) \quad (4)$$

Then

$$\frac{B}{F} = \frac{y_{MB} - z}{y_{MB} - x_{MB}} \quad (5)$$

The feed rate and the feed composition used by the model were each dynamically compensated using a first order lag and a dead time. When a feed compensation analyzer is not available, the product purities and product flow rate are used to calculate an estimate of the feed composition. This back-calculated feed composition is filtered, and the filtered value is used as the feed composition by the tray-to-tray steady-state model. When eqs 1 and 2 are used for setpoint changes, the resulting changes in the manipulated variables are much too sharp; therefore, a filter on the setpoint changes was used to stabilize the controller for setpoint changes. The nonlinear PMBC controller was tuned for setpoint changes based upon minimizing the IAE for the overhead product.

The tray-to-tray steady-state controller model used by the nonlinear PMBC controller used the relative volatility modeled as a function of liquid composition and pressure but used a stage wise

tray efficiency while the dynamic simulator used a Murphree tray efficiency.

An ANN steady-state model was used to replace the tray-to-tray steady-state binary model used by the nonlinear PMBC controller. The feedforward ANN Model consisted of three input nodes, three hidden nodes (one hidden layer), and two output nodes. The transfer functions used were sigmoidal, and the learning algorithm applied was the Levenberg-Marquardt method (Marquardt, 1963). The ANN model considers x_{ss} , y_{ss} , and z as input and calculates the manipulated variables as its output. Because the ANN model did not always match the simulator at steady state, a filtered bias was used to keep the ANN model in agreement with the process (dynamic column simulator). That is, the difference between the measured manipulated variables and its calculated value was filtered on-line. When control calculations were required, the values of x_{ss} , y_{ss} , and z were fed to the ANN model and the resulting manipulated variables were added to the current value of their respective filtered bias. A similar procedure was used for calculating an on-line bias for the bottom flow rate. The ANN model was trained over the expected range of inputs using 700 steady-state data sets from a tray-to-tray steady-state simulator. The ANN model based controller was tuned for setpoint changes.

CONFIGURATION SELECTION

A major degree of freedom in designing a distillation control system is the choice of the manipulated/controlled (u,y) variable pairings. For a two product column, there are, in general, five choices of controlled variables (x , y , L_{AC} , L_{RB} , P) and five manipulated variables (D , L , V , B , Q_{COND}). In addition, there are a variety of ratios that can also be used (e.g., L/D , V/B , L/B , etc). As a result, there are an enormous number of possible (u,y) pairings.

In practice, the choice of (u,y) pairings for distillation control is much more limited. First, condenser duty is usually set and not directly manipulated (e.g., refinery columns typically operate at maximum condenser duty in order to operate at minimum pressure and maximum relative volatility while a significant portion of other columns use a vent and/or inject inerts in the overhead system for pressure control at a generally fixed condenser duty). Second, it is generally not desirable to choose a manipulated variable from one end of the column to control a product composition at the other end (There are exceptions to this rule, e.g., Shinskey (1984) recommends such an arrangement [L/D , D] for a very special class of columns). Note that a configuration designated by (A, B) means that manipulated variable A is adjusted to control the bottoms product composition). As a result, there are three possible manipulated variables for the overhead composition (L , D , L/D) and three possible for the bottoms composition (V , B , V/B); therefore, the total number of configurations to be considered can be reduced to nine: (L , V); (L , B); (L , V/B); (D , V); (D , B); (D , V/B); (L/D , V); (L/D , B); and (L/D , V/B), as indicated in Table 3.

At this point, it should be emphasized that the four manipulated variables (L , D , V , B) should be implemented as ratios to feed flow rate (i.e., L/F , D/F , V/F , B/F). This is because, for a column operating at a constant overall tray efficiency, L , D , V , and B will scale exactly with feed flow rate. Note that in each case, the feed rate used is dynamically compensated. This approach will

greatly reduce the size of the upsets caused by feed flow rate changes. Skogestad et al (1990b) failed to use this approach when testing the (L,V) configuration for feed flow rate upsets. As a result, they observed unrealistically poor control performance for the (L,V) configuration for feed flow rate changes.

There are several ways to implement L/D or V/B control. For example, for L/D control, the distillate flow rate, D, could be set by the accumulator level controller and the reflux flow rate set as a reflux ratio times D. This approach suffers from coupling between the composition controller and the level controller. The two controllers can be decoupled by having the accumulator level controller set the sum of L and D [i.e., $(L+D)_{sum}$] then D and L can be calculated as

$$D = \frac{1}{L/D + 1} [L+D]_{sum}$$

$$L = \frac{L/D}{L/D+1} (L+D)_{sum}$$

Note that as L/D is changed by the overhead composition controller, the proportions of L and D change but their sum (i.e., the draw rate from the accumulator) remains relatively constant. We have observed superior composition control performance for this approach to reflux ratio control compared with the previous approach. V/B control can be implemented in a similar manner in order to decouple the composition control problem from the reboiler level control problem. As a result, this approach has been used to implement all controllers which use L/D or V/B as manipulated variables.

In selecting a configuration from among the nine choices, there are three factors that should be considered: steady-state coupling, sensitivity to disturbances, and dynamic behavior. Each configuration will have its own coupling characteristics which represents a major factor in the configuration selection process for distillation control.

Here the configuration selection was evaluated based upon the performance of each configuration using a conventional PI controller. In this manner, each configuration was tuned for setpoint changes and tested for feed composition upsets.

Single-ended control The configuration selection problem for single composition control (i.e., controlling the composition of only one product) was evaluated by comparing L, D, and L/D for overhead composition control and by comparing V, B, and V/B for bottoms composition control. For example, when L is used to control y, V is fixed and when B is used to control x, L is fixed. In each case, the controllers were tuned for setpoint changes in the product impurity level and tested

for feed composition upsets. Table 4 lists the IAE's for the overhead and bottoms product composition for each of the configuration and for each of the three columns considered. Note that the reflux consistently provided the best control performance for the overhead product while the boilup rate provided the best overall performance for the bottom product. That is, the (L, V) configuration performed best for single-ended control.

Dual Composition Control. Table 5 shows the control performance for each of the nine configurations considered for each column for a step change in the feed composition. For the C₃ splitter, the (L,B) and (L, V/B) configurations provided the best overall performance. For the xylene/toluene column, the (L/D, V) configuration provided the best performance. For the depropanizer, the (L, V/B) configuration provided the best performance. While general guidelines can be given for general selection procedures resulting in reasonable configuration selection, there is currently no reliable procedure to determine the optimum configuration without resorting to control comparisons using detailed dynamic simulators as was done in the study. This is disturbing, because there is considerable difference between the optimum configuration and a reasonable one.

COMPARISON OF CONTROLLERS

Table 6 lists the comparison among the control performance for the conventional PI control, the DMC controller, and the PMBC controller. For the C₃ splitter, the DMC controller provides the best control performance for the overhead product, which is the most important control objective, but generally poorer control performance for the less important bottom product. The PMBC controller is not quite as effective as the DMC controller for the overhead product but does a much better job on the bottom product.

For the xylene/toluene and depropanizer cases, the PI controller clearly out performs the DMC and PMBC controllers. For the xylene/toluene case, the impurity level is 0.1% in both products and it was assumed that both products are equally important. For the depropanizer, the impurity level is 0.5% in both products and once again it was assumed that both products are equally important.

Another design of the C₃ splitter was studied in which both impurity levels were set at 2.0% and both products were equally weighted. The comparison between PI and DMC are shown in Table 7. Note for this case, PI performs better than DMC.

Table 8 shows the results for a different xylene/toluene column with the overhead impurity set at 0.1% and the bottom impurity at 2% with control of the overhead selected as the highest priority. Note that in this case, the DMC controller provides the best control of the overhead product which is the primary control objective in this case.

A number of similar comparison studies were conducted and in each case it was observed that when equal weighting of both products occurred, PI out performed DMC. But when one product is given a high priority for control, DMC provided to be superior.

The PMBC controller did not appear to offer advantages over the PI and DMC controllers. Even the high purity case for which one would have expected the PMBC controller to perform best but, it was generally inferior to PI.

Because the ANN controller is simply a PMBC controller with a neural network steady-state model, ANN also did not offer any advantages over PI or DMC.

CONCLUSIONS

By studying a wide range of column types for a number of designs the following conclusions are made

- For single-ended control apply the (L ,V) configuration
- For dual ended control, the optimum configuration will generally require simulated control studies.
- PMBC and ANN did not offer any advantages over PI and DMC
- DMC out performed PI where the control performance of one product was valued more than the other which is usually the case industrially.

NOMENCLATURE

B	-	bottoms flowrate
D	-	distillate flow rate
F	-	column feed rate
L	-	reflux flow rate
L_{AC}	-	accumulator level
L_{RB}	-	reboiler level
P	-	column pressure
Q_{COND}	-	condenser duty
T	-	tray temperature
u	-	manipulated variable
V	-	vapor boilup rate
x	-	impurity in bottoms product
y	-	controlled variable, purity of overhead product
z	-	mole factor of light component in the feed

Greek Symbols

τ_I	-	reset time
τ_I^{ZN}	-	the ZN settings for reset time

Subscripts

AC	-	accumulator
COND	-	condenser
MB	-	material balance
RB	-	reboiler
sp	-	setpoint
ss	-	steady-state target
sum	-	total

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Table 1

Design Specifications for Each Case Study

	C₃ Splitter	Xylene/Toluene	Depropanizer
Number of Trays	232	49	50
Feed Tray Location (From Bottom)	64	24	18
Feed Flow Rate	13.44 kg/sec(106,400#/HR)	29.03kg/sec (2,332,007#/HR)	16.50 kg/sec[130,422#/HR]
Feed Comp. (mole %)			
Lighter than the Light Key			C ₂ - 0.0193
Light Key	C ₃ - 70	Tol - 67	C ₃ - 0.3154
Heavy Key	C ₃ - 30	Xyl - 33	iso C ₄ - 0.0844
Heavier than Heavy Key 1			n C ₄ - 0.2097
Heavier than Heavy Key 2			C ₅ - 0.1596
Heavier than Heavy Key 3			C ₆ - 0.2116
Factor times minimum reflux	1.3	1.1	1.25
Column Diameter	3.96 m (13ft)	3.96 m (13ft)	2.93 m (9.6 ft)
Overhead Pressure	15.0 atm (221 psia)	0.12 atm (90 mm Hg)	18.0 atm (264 psia)
Overhead Product Impurity	C ₂ - 0.3 mole %	Xyl - 0.1 mole %	0.5 mol% in iC ₄
Bottoms Product Impurity	C ₃ - 2.0 mole %	Tol - 0.1 mole %	0.5 mol% in C ₃
Overhead Flow rate	9.21 kg/sec (73,100 #/HR)	18.53 kg/sec (148,091 #/HR)	4.02 kg/sec (31,800 #/HR)
Overhead Temperature	34.7°C (94.4 °F)	49.5°C (121.1°F)	45.1 °C (113.18°F)
Bottom Flow rate	4.21 kg/sec (33,400 #/HR)	10.5 kg/sec (83, 916 #/HR)	13.05 kg/sec (103,226 #/HR)
Bottom Temperature	42.3°C (108.1 °F)	118.7°C (245.7°F)	139.0°C (282.2°F)
Reboiler Vapor flow rate	131.24 kg/sec (1,041,165 #/HR)	41.87 (334,625 #/HR)	2497 kg/sec (189,936 #/HR)
Reflux Ratio	12.6	1.21	4.26
Feed Quality	Saturated	Saturated	Saturated

Table 2**Assumptions for Each Column Model**

	C₃ Splitter	Xylene/Toluene	Depropanizer
Liquid Dynamic	Hydraulic time Constant	Frances Weir Formula	Hydraulic Time Constant
Negligible Vapor/Holdup	yes	yes	yes
Value dynamic on all flows	no	yes	yes
Accumulation and Reboiler level control	PI	P only	PI
Analyzer delays on product composition	5 minutes	5 minutes	5 minutes
Eqimolal overflow	yes	no	no
Residence time in reboiler	5 minutes	5 minutes	5 minutes
Residence time in accumulator	5 minutes	3 minutes	10 minutes
Heat transfer dynamics modeled	no	yes	yes
Saturated liquid feed	yes	yes	yes
Subcooled reflux	no	no	no
Pressure dynamics modeled	no	yes	no
Perfect mixing of liquid on trays	yes	yes	yes
Ideal VLE	no	yes	no

Table 3

Combinations for Nine Control Configurations

Top

L = Reflux Flow Rate

D = Distillate Flow Rate

L/D = Reflux Ratio

Bottom

V = Boil Up Rate

B = Bottom Flow Rate

V/B = Boil Up Ratio

Table 4**Single Composition Control Results (IAE) For Feed Composition Upset**

Manipulated Variable		C₃ Splitter	Xylene/Toluene Column	Depropanizer
Control of Overhead Product Composition	L	2.1	2.3	51.5
	D	3.8	2.8	148.6
	L/D	19.2	3.0	78.3
Control of Bottoms Product	V	44.4	7.2	14.7
	B	198	8.4	72.1
	V/B	114	7.8	31.1

Table 5

Control Performance (IAE) for Each Configuration for Each Column for Feed Composition Disturbance

<u>Configuration</u>	<u>C₃ Splitter</u>		<u>Toluene/Xylene Column</u>		<u>Depropanizer</u>	
	<u>Overhead</u>	<u>Bottoms</u>	<u>Overhead</u>	<u>Bottoms</u>	<u>Overhead</u>	<u>Bottoms</u>
L, V	15	798	1.6	9.5	45.27	25.31
L, B	4.0	90	2.3	9.2	200.84	99.39
L, V/B	3.4	18	1.3	6.4	29.20	14.53
D, V	5.9	150	6.4	6.4	79.72	22.46
D, B	10.5	354	*	*	85.24	71.25
D, V/B	14.2	114	17.2	9.5	59.05	11.34
L/D, V	5.2	1200	2.2	2.6	30.64	13.09
L/D, B	8.6	1600	1.6	10.5	116.46	53.09
L/D, V/B	5.7	120	0.8	5.2	27.40	10.64

*Due to poor control performance we were unable to obtain results for the configuration

TABLE 6

Comparison (IAE) Among PI, DMC, and PMBC Control Performance for
Feed Composition Step Change

Configuration	C ₃ Splitter (L, B)		Xylene/Toluene (L, V/B)		Depropanizer (L/D, V/B)	
	$\bar{x}$	$\bar{y}$	$\bar{x}$	$\bar{y}$	$\bar{x}$	$\bar{y}$
PI	1.49	0.067	0.052	0.040	10.6	27.4
DMC	5.73	0.013	0.10	0.060	23.5	59.4
PMBC	1.61	0.021	0.739	0.051	47.7	46.9

Table 7

Comparison of Control Performance (IAE) between DMC and PI for a Low Purity C_3 Splitter with Equal Priority on Both Products

	x(IAE)	y(IAE)
PI	1.28	0.578
DMC	4.02	0.530

Table 8

Comparison of Control Performance (IAE) Between DMC and PI for Vacuum Column Purity Given to the Overhead Product

	x (IAE)	y(IAE)
PI	0.21	0.39
DMC	0.024	3.71