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COMPARISON OF ADVANCED DISTILLATION CONTROL METHODS

Third Annual Report

By
James B. Riggs

July 1997

Work Performed Under Contract No. DE-FC04-94AL98747

Prepared:
U.S. Department of Energy
Office of Industrial Technologies
Washington, D.C.

Prepared:
Texas Tech University
Lubbock, Texas

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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 1996 through March 1997, and is the second annual progress report for the project. This report and the two 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) and DOE/AL/98747-2 (in publication).

ABSTRACT

Detailed dynamic simulations of three industrial distillation columns (a propylene/propane splitter, a xylene/toluene column, and a depropanizer) have been used to study the issue of configuration selection for diagonal PI dual composition controls, feedforward from a feed composition analyzer, and decouplers. Auto Tune Variation (ATV) identification (Astrom and Hagglund, 1988) with on-line detuning for setpoint changes was used for tuning the diagonal proportional integral (PI) composition controls. In addition, robustness tests were conducted by inducting reboiler duty upsets.

For single composition control, the (L, V) configuration was found to be best. For dual composition control, the optimum configuration changes from one column to another. Moreover, the use of analysis tools, such as RGA, appears to be of little value in identifying the optimum configuration for dual composition control. Using feedforward from a feed composition analyzer and using decouplers are shown to offer significant advantages for certain specific cases.

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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 to apply, and how to apply it are usually determined based upon internal company politics and hearsay. 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.

RESEARCH APPROACH

The objective of the research is to develop the necessary information for the refinery and chemical industries to be able to make economic-based advanced control decisions.

The challenge to meeting these objectives is that as the particulars of a column change, the relative performance among proportional integral (PI) controls, and the advanced control options are likely to change. For example, it is likely that for some columns that are relatively easy to control there are not likely to be significant performance improvements over PI-controls. And for more difficult problems, the differences are expected to be substantial. Therefore, we must be able identify what makes a distillation column more or less difficult to control. In addition, there are general classifications of distillation cases (highly non-ideal separations, columns with multiple sidestreams, etc.) that must be considered as well.

As a result, the approach used to meet the objective is to use detailed dynamic simulators of a range of distillation columns with varying degrees of control difficulty to assess the control performance of the various control options. The nine options selected for the nine representative

control configurations are listed in Table 1.

It is well known that dynamic models of distillation columns contain the same control difficulties as real distillation columns (nonlinearity, coupling, and nonstationary behavior) when properly applied; therefore, column simulations offer an efficient means to make controlled comparisons of the various control approaches over a range of columns design and operating condition. In fact, the simulators offer a degree of control in comparing distillation control methods that a real column cannot provide.

Distillation columns encompass a wide range of applications. In order to develop advanced control comparisons that will be useful, a range of columns must be considered.

The following factors should be considered in selecting representative columns:

- binary vs. multicomponent separations
 - hydrocarbon vs. highly non-ideal separations
 - high pressure vs. vacuum columns
 - component separations vs. product cuts (e.g., crude columns)
- Based upon these factors the following types of columns would be studied:
- a high reflux ratio binary column (e.g. a propylene/propane column)
 - a multicomponent refinery column (e.g a depropanizer)
 - a non-ideal binary column (e.g. a methanol/water column)
 - a binary vacuum column (e.g a toluene/xylene column)
 - a column with product cuts (e.g. a main fractionator)

Finally, there is the question of credibility. If the results of this study are to have an impact upon the practice of process control of distillation, industry must believe the results of the study and be willing to use them.

PREVIOUS WORK

McCune and Gallier (1973) report using a tray-to-tray dynamic distillation column simulation to study configuration selection for single-ended control. Shinskey (1984) presented a thorough coverage of steady-state relative gain array Relative Gain Array (RGA) analysis applied to eight sample columns considering ten different configurations. Shinskey concluded that the (L/D, V/B) [Note that L/D is used to control the overhead product composition and V/B is used to control the bottom product composition] is preferred for all cases except two of the high reflux ratio cases. His analysis was based upon a steady-state RGA analysis and a limited dynamic analysis, but did not consider the sensitivity of each configuration to feed composition changes.

Takamatsu et al (1987) use a geometric approach to evaluate the interaction between loops for dual composition control of columns. They concluded that the (L/D, V/B) configuration provides

the minimum interaction and used simulations to confirm this finding although the controllers were not tested for feed composition upsets.

Skogestad et al (1990a) analyzed four different control configurations [(L,V), (D,V), (L/D, V/B), and (D,B)] for seven different binary columns, using dynamic RGA's and optimal values. They point out that each configuration has its own sensitivity to disturbances, and, as a result, certain configurations require less need for feedback action during disturbance rejection. There are several questionable dynamic distillation modeling assumptions that were made (Riggs, 1990) and limited closed loop results were presented.

Waller (1992) presented an experimental comparison of (four configurations (L,V), (D,V), (L/D, V), and (L/D, V/B)] for a 15 tray ethanol/water column. The four schemes were compared for two-point control of two tray temperatures inside the column. The (L/D, V) and the (L,V) configurations had the best performance for a step change in feed composition while the (D,V) configuration was clearly the worst. On-line trial-and-error tuning was performed and results for only one change in feed composition were reported.

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 2. 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 3 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. Since 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 C_3 splitter and the xylene/toluene column.

C_3 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 C_3 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 C_3 splitter was bench-marked against dynamic industrial data for a C_3 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).

DIAGONAL PI TUNING

Tuning of diagonal PI controllers has a major impact upon the overall control performance. For dual composition control of a distillation column, there are four unknown tuning parameters (two gains and two reset times). On-line trial-and-error tuning (i.e., a four dimensional search for an optimum) is an extremely challenging problem and usually leads to inferior control performance. This is because columns are relatively slow responding systems, and typically are continuously subjected to a variety of disturbances. Moreover, it is essential that the resulting controller settings

are able to handle the full range of upsets in order to obtain an adequate on-line service factor (i.e., robustness).

There are a number of methods for tuning PI single loop controllers, most notably the Cohen and Coon method (Cohen and Coon, 1953) and the Ziegler-Nichols (ZN) ultimate method (Ziegler and Nichols, 1942). The Cohen and Coon method is based upon open loop responses which are difficult to obtain due to relatively slow response times of distillation columns, and are susceptible to significant error due to disturbances that occur during the identification periods. The ZN method is based upon measuring the ultimate controller gain and ultimate period but can result in excessive variations in the controlled variables.

Astrom and Hagglund (1988) used a relay feedback approach [Auto Tune Variation (ATV)] to measure the ultimate controller gain and the ultimate period. Then, the classical ZN settings can be used directly. The relay height (Δu), i.e., the chosen size of the change in the manipulated variable, can be selected to be small enough that the process is not unduly upset and large enough that the changes in the controlled variable are large accurately measured above the measurement noise. The results of an ATV test are the amplitude, a , and the period, P_u . The ultimate gain is given by

$$K_u = \frac{4\Delta u}{\pi a}$$

We have found that for distillation columns, which are typically sluggish, the Tyreus-Luyben (TL) settings (Tyreus and Luyben, 1993) are generally superior to ZN setting; therefore, all the results presented here are based upon Tyreus-Luyben (TL) settings.

The TL settings are given by

$$K_c^{TL} = K_u / 3.22$$

$$\tau_I^{TL} = 2.2 P_u$$

If TL controller settings are applied to a multivariable problem, such as a distillation column, sluggish or ringing controller performance can result. Luyben (1986) proposed the Biggest Log Modulus (BLT) method for tuning diagonal PI controllers. The BLT method is based upon decreasing each controller gain and increasing the reset times by a single detuning factor, F_{DT} (Toijala and Fagervik, 1972).

$$K_c = K_c^{TL} / F_{DT}$$

The BLT method uses a multivariable Nyquist plot to determine the value of F_{DT} that meets preset stability criteria. The procedure used here is to adjust F_{DT} to minimize the Integral Absolute Error [IAE] for a setpoint change in the product impurity levels in order to optimize tuning

$$\tau_I = \tau_I^{TL} \times F_{DT}$$

performance.

Two PI controllers were applied to the xylene/toluene simulator using the (L,V) configuration. One PI controller was tuned using the BLT method using ATV tests to identify integrator plus dead-time models (Friman and Waller, 1994) resulting in a value of F_{DT} of 4.5. The other controller had the same ZN settings but was detuned by trial and error resulting in a value of F_{DT} of 1.0. The IAE of the controller tuned using the BLT method was four times as large as the controller tuned using online detuning for the bottom composition controller. The BLT tuning procedure uses a fixed but conservative approach to stability analysis, and does not directly consider the magnitude of disturbances. Therefore, a linear, stationary process, subjected to low disturbance levels, could be under-tuned using the BLT method while a highly nonlinear, nonstationary process, subject to large disturbance levels, could be tuned too aggressively using the BLT procedure. Because the BLT procedure uses linear, fixed gain models and does not consider the magnitude of disturbances, there can be a substantial difference between the value of F_{DT} it calculates and the optimal value of F_{DT} . In fact, the one-dimensional search required to adjust F_{DT} should be relatively straight forward in an industrial setting using an analysis of setpoint changes and/or disturbance rejection performance since the user only has to decide if the controller is too aggressive or too sluggish. Moreover, the value of F_{DT} typically ranges between 0.5 and 3.0.

In order to investigate the sensitivity of ATV tuning with on-line detuning to normal process variations, ATV tests were conducted on the xylene/toluene simulator with a five percent decrease in tray efficiency, a five mole percent decrease in feed composition, and a ± 5 percent noise on the analyzer readings. Using the corresponding TL settings (i.e., without the changes in the efficiency, change in feed composition, and measurement noise), the diagonal PI controllers were applied to the base case (i.e., without the changes in tray efficiency, change in feed composition, and measurement noise) and detuned on-line. The resulting controller performance was essentially equivalent to the control performance obtained at the base case without the process variations. Thus, these results indicate the lack of sensitivity of this tuning procedure to typical process variations. The trial-and-error adjustment of F_{DT} tends to absorb a certain amount of error itself. In addition, because the ATV procedure is much faster than the response time of the process, the ATV procedure provides more of a "snap shot" of the process because disturbances do not have as much time to corrupt the results. Figure 1 shows four cycles of an ATV test as well as an open loop response for the bottom composition of the C_3 splitter. Note that the four cycles for the ATV test required approximately one-eighth the time required for the open loop step response.

In summary, the ATV procedure provides relatively fast identification for slow processes (almost an order of magnitude faster than open loop tests) and can usually be adjusted so that the process is not unduly upset. The ATV procedure was not significantly sensitive to changes in disturbances and nonstationary behavior. Further, ATV with on-line detuning provides near optimal tuning of diagonal PI multivariable controllers in an industrially relevant fashion.

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 1.

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. Skogested 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 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. The Relative Gain Array (RGA; Bristol, 1966) provides a steady-state measure of coupling in multivariable systems and is used to evaluate the steady-state coupling of each configuration.

In addition, each configuration will have its own sensitivity to disturbances. Feed composition changes usually represent the most challenging disturbance for distillation column control on a regular basis. The (L, V) configuration has been found to offer significant advantages for rejecting feed composition disturbances (e.g., Haggblom and Waller, 1990). This results because the energy required to meet product specifications does not usually change a great deal for feed composition changes while the D/B ratio does. When feed composition changes occur using the (L, V) configuration, the majority of the required adjustment to D and B are made by the reboiler and accumulator level controllers. Each configuration, for each test column, is evaluated with regard to its sensitivity to feed composition changes.

Finally, dynamic factors are also important for configuration selection. The initial response of a controlled variable to a change in its manipulated variable is important. A low deadtime, high dynamic gain response is desirable while a sluggish, low gain response is not. Also, dynamic coupling can be important. For example, the steady-state coupling may be low while the dynamic coupling can be large due to high order dynamics (e.g., inverse action, overshoot). One measure of dynamic effects is the dynamic RGA (McAvoy, 1983). The dynamic RGA is based upon linearized response models that assume fixed gains and time constants. But distillation columns are known to exhibit nonlinear gains and nonlinear time constants (Stathaki et. al., 1985). As a result, the dynamic analysis used here will be based upon positive and negative open loop responses obtained from the nonlinear column simulators for each manipulated variable in each configuration. In this manner,

the degree of overdamped response, inverse action, and nonlinearity in the time constants for each configuration can be directly assessed.

Following is the analysis of configuration selection for the three case studies: a C_3 splitter, and a xylene/toluene vacuum column, and a depropanizer. Analysis with respect to steady-state coupling, disturbance rejection, and dynamic factors is considered. In addition, control comparisons for feed composition disturbances are made among the various configurations using ATV-based controller tuning for single composition control, dual composition control, feedforward control and decoupler performance. Feed composition upsets were chosen because they represent the most difficult day-to-day operational upset. In each case, the tuning was performed for setpoint changes in the impurity levels and was based upon minimum IAE.

For each case, robustness of one of the better configurations was tested by implementing a reboiler duty upset. Robustness or reliability has to do with the ability of controller to remain in service for the full range of process upsets. The most severe upset that a distillation column control system will likely encounter is a reboiler duty upset; therefore, the ability of a control system to effectively handle a reboiler duty upset was used as an indicator of controller robustness. The reboiler upset was implemented on each simulation by externally decreasing the nominal boilup rate by a certain fraction and then allowing the controller to deal with the impending upset (Shinskey, 1995). The robustness of the controller was measured by the maximum reduction in reboiler duty that the controller could handle and remain stable.

C_3 SPLITTER

Following is an analysis of the configuration selection process for the C_3 splitter test case for single and dual composition control as well as an analysis of feed forward control, decoupling control, and robustness.

RGA Analysis. Table 4 shows the RGA analysis (λ_{11} values) for eight of the nine possible configurations [Note that the (D,B) configuration has an infinite steady-state RGA (Skogestad et al, 1990b)]. The RGA results were developed based upon reducing the size of the change used to develop the RGA values in the steady-state simulator until the RGA values converged; therefore, these RGA values are locally valid at the base case conditions. Note that the (L/D, B), (L, B) and the (L/D, V/B) configurations appear to be the most promising based upon this RGA analysis.

Sensitivity to Feed Composition Changes. Table 5 shows the relative changes in manipulated variables for a feed composition change while maintaining the product purity at its base case value. It should be pointed out that because the equimolal overflow assumption was used, the open loop response of each configuration would produce the exact same changes in the product composition. Note that reflux and reflux ratio have the least sensitivity to feed composition changes of the manipulated variables in the overhead while boilup rate has the least sensitivity for the bottoms (Table 5). As a result, configurations involving L, L/D, and V should have the smallest upsets to

absorb by feedback action. Overall, it is the sensitivity to feed composition changes combined with the decoupling performance of the configuration that determines the ability of a configuration to handle feed composition upsets.

Dynamic Analysis. For the dynamic analysis, open loop responses for positive and negative changes in each manipulated variable were generated using the column simulator for each of the nine configuration considered. First, consider the effect of each manipulated variable upon its controlled variable, i.e., the diagonal responses, e.g., V on x, L/D on y, D on y, etc.. In all cases, except the effect of V and V/B on x in the (D,V) and (D, V/B) structures, respectively, the open loop responses were well-behaved, i.e., the controlled variable has very nearly first order responses to a step change in its manipulated variable. In fact, when either L or V is kept constant during the step test, consistent trends in the responses are observed. For example, for the (L,V), (L, V/B), and (L,B) configurations for step changes in V, V/B, and B, respectively, the response time for the bottom composition became longer in the order V to B. Likewise, for the (L,V), (L/D, V), and (D,V) configurations for step changes in L, L/D, and D, respectively, the response times for the overhead compositions became longer in the order L to D. This results because L and V have the most direct effect on the column while D and B depend upon level control in the accumulator and reboiler to affect the product compositions.

Overall, the (L,B) configuration had the fastest response time. For step changes in the overhead manipulated variable, the (L,B) configuration had a response time that was about four times faster than the next fastest configuration [(L/D, V/B)]. For the step tests in the bottoms manipulated variable, the (L,B) configuration was nearly as fast as the fastest configuration [(L,V)]. In general, the time constants for positive and negative changes were significantly different for the overhead. Time constant ratios for positive and negative changes of five to one were observed for most cases. For the bottom, the time constants for positive and negative changes were relatively linear for most configurations because it is a lower purity product..

Consider the open loop responses for the (D,V) and (D, V/B) configurations for the step changes in V and V/B, respectively. In both cases, they exhibit a relatively small steady-state gain with an overshoot that is at least five times larger than the gain. This result, which was pointed out by Shinskey (1984), results because of the dynamic differences between liquid flow and vapor flow through the column.

It is interesting that the (L,B) configuration did not exhibit strong overshoot in the overhead composition for changes in the reflux rate because the same type of dynamic mismatch between the liquid and vapor exists for the (L,B) configuration as the (D,V) structure. The difference appears to be due to the order that the various lags are applied. For the (L,B) configuration, the step change in reflux gradually works its way down the column instead of a very quick change in vapor rate for the (D,V) structure. Furthermore, before all the liquid rates down the column can change completely for the (L,B) structure, the vapor rate up the column begins to change damping the initial effect on the top composition even for the loose level control considered in the C_3 splitter. The (L,B) configuration did show an overshoot in the overhead composition for a step change in L, but it was

small compared to the steady-state change.

The (D,B) structure has integrating responses and is open loop unstable. When either B or D is increased, the impurities in the overhead and bottoms increase continuously.

Now consider the dynamics of the coupling interactions, e.g, V on y, L/D on x, D on x, etc. Most of the interactions approximately followed first order dynamics. The exceptions exhibited an inverse response as a result of the aforementioned dynamic difference between liquid and vapor flows. Those cases for which an inverse response was observed were (1) the bottom composition for a change in reflux for the (L,B) configuration, (2) the bottom composition for a change in the reflux ratio for the (L/D, B) configuration, (3) the overhead composition for a change in vapor flow rate for the (D,V) configuration, and (4) the overhead composition for a change in the boilup ratio for the (D, V/B) configuration. Note that these cases are all the possible material balance configurations in which L, L/D, V or V/B (whichever is present) is changed. Inverse action was also observed for the bottoms composition for changes in the distillate flow for the (D, B) configuration.

Single-Composition Control Results. The configuration selection problem for single composition control was evaluated by comparing L, D and L/D for over head 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 controller was tuned for a 0 impurity setpoint change and was tested for a step change in feed composition from 70 to 65 mole % propylene. Table 6 lists the IAE's (given in units of mole fraction x seconds) for control of the overhead and bottom product compositions for each of the configurations considered for a step change in feed composition. Note that reflux flow rate (L) provided the best control the best control performance for single composition control of the overhead product composition, and the boilup rate (V) provided the best control performance for the bottom product composition. Figure 2 shows the control performance for the three manipulated variables (L, D, and L/D) considered for single composition control of the overhead product.

Dual-Composition Control Results. Each of the nine configuration considered was tuned (i.e., F_{DT} selected) using setpoint changes in the overhead product impurity level. At $t = 100$ minutes, the setpoint for the overhead product was reduced from 0.3 to 0.15 mole % propane, and at $t = 2000$ minutes, the setpoint was changed to 0.45 mole % propane. The detuning factor F_{DT} for each configuration was selected to minimize the IAE for the overhead product alone because the overhead product is used as a feed stock for a polymer reactor while the bottoms product is fuel grade propane.

Each configurations was tested for a step change in feed composition from 70 to 65 mole % propylene. The IAE's (given in units of mole fraction x seconds) for each configuration and for both products are listed in Table 7. Note that the best performance was provided by the (L, V/B) and (L, B) configurations. Figure 3 shows the control performance for this test for the (L, V/B), (L,B) and (D,B) configurations and Figure 4 shows the corresponding manipulated variable action for the (L, V/B) and (L,B), and (D,B) configuration for sinusoidal feed composition upsets. The closed loop Bode plots generally agree with the IAE results for a step feed composition change and the

results shown in Figure 5.

Based upon steady-state RGA analysis, the (L/D,B), (L,B) and (L/D, V/B) configurations, were identified as likely candidates. By consideration of the sensitivity to feed composition changes, the (L, B) configuration would be preferred from among these three choices. Finally, the (L, B) configuration was found to be the fastest responding configuration. Therefore, the (L, B) configuration would be selected as the configuration of choice which proved to be generally consistent with the performance indices for the disturbance rejection cases.

Feedforward Results. For the previous results, it was assumed that the feed composition was unmeasured. When a feed composition analyzer is available, feedforward compensations to the feedback controller can be made. Feedforward controllers for feed composition changes were developed using step test results and were implemented as lead/lag elements (Marlin, 1995).

Feedforward controllers were added to the (L,B) feedback controller and tested for a step change in feed composition from 70 to 65 mole % propylene. The resulting IAE's (given in units of mole fraction x seconds) for both products for feedforward control to the overhead control loop only, feedforward control to the bottoms control loop only, feedforward control to both loops, and feedback only control are listed in Table 8. Note that the feedforward controller significantly worsened the control performance for the overhead product for feedforward to the overhead and feedforward to both loops. Feedforward to the bottom loop and feedforward to both loops did reduce the IAE for the bottom product composition by a factor of about 2. Based upon results for the sensitivity to feed composition changes shown in Table 5, one would expect that the greatest benefit of feedforward control would occur for the bottom composition control loop for the (L, B) configuration which is consistent with the results shown here.

Decoupler Results. Decouplers can be added to diagonal PI controllers in an effort to improve control performance. In principle, a decoupler can reduce coupling allowing more aggressive tuning and consequently better disturbance rejection. In each case, the decouplers were designed using first order plus deadtime models for the coupling effect and manipulated variable effect on the control objective. Then, the results of the first order plus deadtime models were implemented as a lead/lag element. Finally, the decoupler and feed back controller were tuned using impurity setpoint changes. Decoupler results for the C₃ splitter are shown in Table 9. First of all, note that the resulting detuning factors, F_{DT} for each of the decouplers was slightly higher than the detuning factor for the feedback controllers without decoupler except for the one-way decoupler to the overhead product. As a result, the decouplers tested had a control performance that was worse than that for the feedback controller without a decoupler except for the one-way decoupler to the overhead product.

Robustness Analysis. The (L,B) configuration was studied to evaluate the robustness of the tuning procedure used here. First, the size of the setpoint change used to tune the controller was adjusted. It was found that the resulting optimum detuning factor ($F_{DT} = 0.8$) remained unaffected by the size of the setpoint change. Next, the (L,B) configuration was tuned based upon a step change in feed composition. Feed composition changes ranging from 1 mole % to 15 mole % were used, and it was

found that a detuning factor of 0.4 provided the best performance in all cases.

Finally, a procedure suggested by Shinskey (1995) was used to evaluate the controller robustness for both cases (i.e., F_{DT} equal 0.4 and 0.8). The column controller was reduced by a certain fraction at which point the bottom composition controller was turned on. As a result, the bottom composition controllers acts as if the reduced boilup rate is sufficient and the column is sharply upset. For the tuning based upon the feed composition upsets ($F_{DT}=0.4$), the maximum reduction in V that the controller could handle in a stable manner was 40%. The controller tuned based upon setpoint changes ($F_{DT}=0.8$) was able to handle a 70% reduction in V . In addition, even though the disturbance tuned controller remained stable up to a 40% reduction in V , it exhibited significant ringing even for a 1% reduction in reboiler duty.

These results support the use of setpoint changes as a basis for controller tuning, i.e., the composition controllers which were tuned using setpoint changes were clearly more robust and did not exhibit ringing, and therefore, should represent a better compromise between reliability and performance.

XYLENE/TOLUENE COLUMN

Following is an analysis of the configuration selection process for the xylene/toluene column for single and dual composition control as well as an analysis of feedforward control, decoupling control, and controller robustness.

RGA Analysis. The steady-state RGA results (λ_{11} values) for the eight of the possible configurations are listed in Table 4. Note that the (L/D, V/B), (L, V/B), (L/D, V) and (L/D, B) configurations appear to be the most promising based upon the RGA analysis.

Sensitivity to Feed Composition Changes. Table 5 lists the relative changes in each manipulated variable for a 5% increase in feed composition based upon steady-state simulation runs keeping the product compositions at their respective setpoints. Note that the reflux rate has the least sensitivity to feed composition changes of the manipulated variables available for the overhead while the boilup rate shows the least sensitivity for the bottoms.

Dynamic Analysis. For the dynamic analysis, positive and negative open loop tests were generated using the column simulator for all nine configurations. For all the open loop tests (i.e., diagonal interaction and coupling), no significant inverse action was observed. This was probably due to tight level control and the particular characteristics of the xylene/toluene column. Almost all the open loop responses were well represented as first order responses, but some highly overdamped responses were also observed. That is, for the diagonal interaction, a negative change in reflux ratio for the (L/D, V) configuration, and a positive change in distillate flow for the (D, V) configuration resulted in an "S" shaped response for the overhead composition. For coupling interactions, only the overhead composition exhibited a highly overdamped response and only for the cases in which the overhead composition increased. All configurations had an "S" shaped response, when the

overhead impurity increased except the (D, V) and (D, V/B) configurations.

The (L, V) configuration was clearly the fastest responding configuration. It was almost three times faster than the next fastest configuration [(L/D, V)] for the bottoms composition, and was as fast as the fastest configuration for the overhead composition. The (L/D, V/B) configuration was three to five times slower than the (L, V) configuration for the bottoms composition, but almost as fast for the overhead composition. The (L/D, V/B) configuration exhibited time constants that were fairly linear for positive and negative changes while the (L, V) structure had time constants that differed by a factor of about two for positive and negative changes.

Single-Composition Control Results. In a manner similar to the approach used for the C₃ splitter, single composition control 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. Each controller was tuned for a 50% impurity setpoint change and was tested for a step change in feed composition from 67 to 62 mole% toluene. Table 6 lists the single composition control results for the xylene/toluene column. Note that L is slightly better than D and L/D for single composition control of the overhead while V is only marginally better than B and V/B for single composition control of the bottoms product.

Dual-Composition Control Results. Each of the nine configurations was tuned using setpoint changes in the overhead and bottoms product impurity levels. Once tuned, each configuration was tested using a step change in feed composition from 67 to 62 mole% toluene. Table 7 lists the control performance (IAE) for each product for each configuration. Note that the (L/D,V) configuration had the best performance followed by the (L/D, V/B) and (L, V/B) configurations. Figure 6 shows the control results for the feed composition upset for the (L/D, V), (L/D, V/B) and (L, V) configurations while Figure 7 shows the corresponding changes were found to be consistent with the IAE results in Table 7.

Based upon the RGA analysis, the (L/D, V/B), (L, V/B), (L/D, V), and (L/D, B) configurations looked the most promising, but the (L, V) configuration was found to be the least sensitive to feed composition upsets and to be the fastest responding configuration.

Feedforward Results. Feedforward controllers using on-line feed composition measurements were developed from open loop step tests and implemented as lead/lag elements with the existing feedback controllers.

Feedforward controllers were added to the (L/D, V/B) feedback controller and tested for a change in feed composition from 67 to 62 mole toluene. The resulting IAE's for both products for feedforward control to both control loops only, feedforward control to the bottoms control loop only, feedforward control to both control loops, and feedback only control are listed in Table 9. Note that feedforward provided an improvement in all cases especially for feedforward to both loops. Using feedforward to both control loops reduces the IAE for the overhead loop by a factor of 4 and the bottoms loop by a factor of 3.

Feedforward controllers were also applied to the (L,V) configuration. Feedforward to the bottom control loop reduced the IAE of the top loop by a factor of two and the bottom loop IAE by a factor of 5. Overall the (L/D, V/B) configuration with feedforward to both control loops provided the best overall performance.

Decoupler Results. The (L,V) configuration has the advantage of being the best configuration with regard to feed composition changes and dynamics response, but the worse configuration with regard to coupling. The (L/D, V/B) configuration is generally the best with regard to coupling. For a boilup rate change, ΔV , the (L/D, V/B) configuration puts part of the resulting vapor rate change at the top of the column back down the column as reflux. This action helps to decouple the overhead composition from boilup rate changes.

The same approach can be added to the (L, V) configuration in order to improve its decoupling characteristics. That is, when the bottom composition controller calls for a change in the boilup rate, $\Delta V_{\text{control}}$, a fraction, F_c , of the change can be added directly to the setpoint for the reflux flow controller.

F_c can be tuned by making changes in the boilup rate and adjusting F_c until the resulting effect upon y is minimized. F_c does not normally require dynamic compensation because boilup rate changes propagate up the column relatively quickly. In a similar manner, the effects of changes in the reflux flow rate, called for by the overhead composition controller upon the bottom composition,

$$\Delta L_{DC} = F_c \Delta V_{\text{control}}$$

can be reduced using another simple decoupler. This is,

$$\Delta V_{DC} = F_c \Delta L_{\text{control}}$$

In this case, F_c requires dynamic compensation in the form of a dead time plus a lag because of the presence of liquid dynamics.

The simple decoupler results for the xylene/toluene column are listed in Table 9. Note that the two-way decoupler and the one-way decoupler to the bottom product composition controller provided substantial improvement over the feedback controller without a decoupler. For the two-way decoupler, the IAE for the both products was reduced by a factor of about 2.5. Note that the detuning factor for the decouplers was significantly lower (i.e., more aggressive) than the feedback controllers without a decoupler.

Robustness Tests. The (L,V) configuration was used to evaluate the tuning procedures used here. It was found that the size of the setpoint change that was used to tune the (L,V) controller did not affect the tuning results.

The Shinskey robustness test was used to evaluate the robustness for the controllers tuned using setpoint changes and using feed composition upsets. The feed composition tuned controller was 67% more aggressive than the setpoint tuned controller. The feed composition tuned controller was able to handle up to a 10% reduction in reboiler duty while the setpoint tuned controller could handle up to a 15% reduction and remain stable. The two-way decoupling was tested for robustness and found to be stable up to a 35% reduction in reboiler duty. Once again, the feed composition tuned controller exhibited ringing for a 1% reduction in reboiler duty while the setpoint tuned controller exhibited ringing only near the stability limit.

DEPROPANIZER

Following is an analysis of the configuration selection process for the depropanizer for single and dual composition control as well as an analysis of feedforward control, decoupling control, and controller robustness.

RGA Analysis. The steady-state RGA results (λ_{11} values) for eight of possible configurations are listed in Table 4. Note that the (L/D, V/B), (L, V/B), and (L/D, V) configurations appear to be the most promising configurations based upon steady-state RGA values.

Sensitivity Feed Composition Changes. Table 5 shows the average percentage change in the overhead and bottoms impurities for a +5% change in the light key. These results are based upon using a steady-state column simulator while maintaining product purity at their base case values. The (L, V) structure is clearly the least sensitive configuration to feed composition changes.

Dynamic Analysis. Open loop responses using positive and negative 1% changes in each manipulated variable were used to assess the dynamic performance of each configuration. First consider the diagonal responses. Once again the (L, V) configuration was the fastest responding structure while the slowest responding were the (D, V) and (D, V/B) configurations. The (L/D, V/B) configuration and the (L, V/B) configuration were generally three times slower than the (L, V) structure.

The material balance configurations [(L, B), (L/D, B), (D, V) and (D, V/B)] each showed overshoot when the manipulated variable opposite D or B was changed. The remainder of the diagonal responses were generally first order-like. The steady-state gains and time constants for the overhead product were much more linear than for the bottoms product. The (L/D, V/B) configuration has the most linear responses while the (L, B) structure was the most nonlinear.

Single-Composition Control Result. Single composition control 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. Each controller was tuned for a 25% impurity setpoint change, and was tested for a step change in feed composition from 31.54 to 26.54 mole% propane while isobutane increased from 8.44 to 13.44 mole %. Table 6 lists the IAE's for each of these tests. Note that L outperformed

L/D and D for the overhead composition control loop and V performed best for the bottom composition control loop.

Dual-Composition Control Results. Each of the nine configurations was tuned using 25% impurity setpoint changes in the overhead and bottoms impurity levels. Once tuned, each configuration was tested using a step change in propane from 31.54 to 26.54 mole % while isobutane was increased from 8.44 to 13.44 mole%. Table 7 lists the control performance (IAE) for each product for each configuration. Note that the (L/D, V) configuration had the best overall performance closely followed by the (L, V) configuration. Figure 8 shows the control results for the (L/D, V), (L, V), and (L/D, V/B) configurations for the feed composition upset and Figure 9 shows the corresponding manipulated variable action. The close loop Bode plots were found to be consistent with the results in Table 7. Note that certain configurations perform quite well for one product but quite poorly for the other. For example, the (L/D, B) configuration had the best performance for bottom composition control, but was the worst for the overhead composition control.

Feedforward Results. Feedforward controllers using an on-line feed composition measurement were developed from open loop step tests and implemented as lead/lag elements with the existing feedback controllers.

Feedforward controllers were added to the (L/D, V) feedback controller and tested for a 5 mole % decrease in propane in the feed. The resulting IAE's for both products for feedforward control to the overhead control loop only, feedforward control to the bottoms control loop only, feedforward control to both control loops, and feedback-only control are shown in Table 8. Note that feedforward to the overhead product composition controller resulted in significant control performance improvement while feedforward to both loop significantly worsened the control performance of both loops.

Decoupler Results. Simple decouplers identical to the ones applied to the xylene/toluene column were applied to the depropanizer. Table 9 compares the decoupler results with the feedback controller without decoupler for the (L, V) configuration. Note that each decoupler considered resulted in about a 50% reduction in IAE for the bottom product and a 25% reduction in IAE for the overhead product. Also note that each decoupler used a detuning factor of 0.9 compared to 1.4 for the (L, V) configuration without decoupler.

Robustness Results. The (L, V) configuration was used to evaluate the tuning procedure used here. It was found that the size of the setpoint change that was used to tune the (L, V) controller did not significantly affect the tuning results.

The Shinskey robustness test was used to evaluate the robustness for the controller tuned using a setpoint change and using a step change in the feedcomposition. The feed composition tuned controller was 55% more aggressive than the setpoint tuned controller. The feed composition tuned controller was able to handle up to only a 10% reduction in reboiler duty and remain stable while

the setpoint tuned controller could handle up to a 65% reduction. The (L,V) with the two-way decoupler, which was tuned with setpoint changes, used the same detuning factor (0.9) as the feed composition tuned controller, but was able to handle up to a 40% reduction in reboiler duty and remain stable.

DISCUSSION OF RESULTS

This section discusses the various aspects considered for each of the three test cases studied.

RGA Analysis. The RGA values in Table 4 are quite consistent with the RGA results reported by Shinskey (1984). Moreover, the results of Table 4 show the same general trends: the (L/D, V/B) configuration always has the least coupling for configurations with RGA values greater than one and the (L, V) configuration shows the largest degree of steady-state coupling.

A comparison between the RGA values and the control results in Table 7 does not show a good correlation between steady-state RGA values and control performance. For example, the (L, V) configuration performed significantly better than the (L/D, V/B) configuration for the depropanizer. Moreover, the best performing configurations for each of the three case studies [i.e., C₃ splitter: (L, V/B); xylene/toluene: (L/D, V); depropanizer: (L/D, V)] had RGA values, in each case, that did not distinguish themselves from the other potential choices. Also, because the RGA is a single number, it cannot be expected to identify cases, such as, the (L/D, B) configuration for the xylene/toluene column for which the overhead control performance was quite good while the bottom control performance was the worst observed.

Overall, The RGA indicates steady-state coupling, and therefore, by itself, cannot be expected to critically differentiate between potentially viable control configurations. Even for configurations that should be ruled out based upon RGA values [e.g., the (D, V) configurations for all three cases] were found to perform moderately well.

Sensitivity to Feed Composition Changes. In each case, reflux and boilup required the least relative change for feed composition changes. Thus, the (L, V) configuration would be expected to be least affected by feed composition changes.

Dynamic Analysis. Manipulation of reflux flow rate had the fastest response for the overhead product composition followed by the reflux ratio and the distillate product flow rates. Likewise, the boilup rate had the fastest response for the bottoms followed by the boilup ratio and the bottoms product rate. Whenever B or D is manipulated, complex coupling dynamics can result. The (L, V) configuration was observed to have fast acting, well-behaved dynamics in all cases.

Single-Composition Control. For single-composition control of the overhead product composition, reflux was identified as the best manipulated variable for top composition control while boilup rate

was found to be best for bottom composition control. The (L, V) configuration would be expected to be the best for single composition control because coupling is not an issue, and the (L, V) has been shown to be superior with regard to sensitivity to feed composition changes and with regard to dynamic behavior.

Dual-Composition Control. The (L,V) configuration was shown to be superior with regard to rejecting feed composition upsets and dynamic behavior while, with regard to coupling, the (L, V) configuration is one of the most highly coupled configurations. The (L/D, V/B) configuration is one of the least coupled configurations, but it is very sensitive to feed composition upsets. Therefore, for dual composition control, there is no obvious choice for the best configuration which makes configuration selection an interesting and challenging problem.

For the C_3 splitter (i.e., high reflux ratio case), it was found that the material balance configurations generally out performed the energy balance configurations. For the lower reflux ratio cases (i.e., xylene/toluene column and depropanizer), the energy balance configuration performed generally better than the material balance configurations. Identifying *a priori* the best configuration in either case is a nontrivial problem. At this point, detailed dynamic column simulation and control configuration testing is the only reliable means to identify the optimum configuration.

Feedforward Control. Feedforward control is shown to offer significant control performance enhancement in certain cases. When feedforward is applied to an end of the column for which reflux is used as the manipulated variable (e.g., see Table 8 for the C_3 splitter), or when feedforward is applied to an end of the column for which boilup rate is used as the manipulated variable (e.g., see Table 8 for the depropanizer), degradation in control performance actually results. This is probably related to the fact that reflux and boilup, used as manipulated variables, are least sensitive to feed composition upsets and each also promotes coupling.

On the other hand, the results in Table 4 demonstrate that feedforward applied when L/D, V/B, B, or D is used as a manipulated variable can be quite effective. Moreover, when feedforward is applied to each of these manipulated variables, the best control performance results. For example, feedforward to the bottoms control loop-only provided the best performance for the (L, B) configuration applied to the C_3 splitter. Feedforward to both control loops provided significant performance improvement for the (L/D, V/B) configuration applied to the xylene/toluene column. And finally, feedforward to the overhead loop-only resulted in a significant performance enhancement for the (L/D, V) configuration applied to the depropanizer.

Decoupler. Table 9 shows the results for the decouplers studied. Note that the decouplers were applied to the (L, V) configuration for the xylene/toluene column and depropanizer. This was done because the (L, V) configuration suffers from the largest degree of coupling. In both cases, each of the decouplers considered resulted in a detuning factor, F_{DT} , that was significantly more aggressive than the feedback controllers without a decoupler. As a result, the control performance for feed composition upsets was significantly improved with the decoupler. For the depropanizer, each decoupler considered performed equivalently while the two-way decoupler performed best for the

xylene/toluene column.

For the C_3 splitter, only the decoupler to the overhead resulted in a lower detuning factor and improved control performance.

Robustness Analysis. Using the robustness analysis procedure described earlier, it was shown in all three cases that tuning based up-set point changes appears to a reasonable compromise between performance and reliability. Tuning based upon disturbance rejection is clearly too aggressive. The two-way decoupler applied to the depropanizer and xylene/toluene columns had the aggressiveness of a disturbance-tuned controller but with superior robustness.

CONCLUSIONS

The control performance of a particular distillation column configuration is affected by its coupling, sensitivity to disturbances, and dynamic behavior. For single-composition control, the (L,V) configuration is preferred because coupling is not an issue, and the (L, V) configuration has superior characteristics with regard to rejection of feed composition upsets and with regard to dynamic behavior.

For dual composition control, the optimum configuration is not generally obvious. It requires simulated control studies in order to identify the best configurations. Generally, for high reflux ratio columns, material balance configurations are preferred while for low to moderate reflux ratios, energy balance configurations were shown to be better. But, the variations in control performance among these groups of configurations is substantial. In fact, analysis tools, such as RGA, dynamic RGA, condition number, etc, are not reliable for differentiating the configuration choices within the energy and material balance configurations.

Feedforward using a feed compositions analyzer has been shown to provide significant performance improvement when the manipulated variable is L/D, D, V/D, or B. A decoupler is beneficial when either or both L and V are manipulated variables. These results are consistent with the coupling characteristics and the sensitivity to feed composition upsets for the various manipulated variable choices.

Using a novel robustness analysis procedure, it has been shown that tuning composition controllers, using setpoint changes in product compositions, provides a good compromise between reliability and performance.

NOMENCLATURE

a	-	amplitude of the measured variable in an ATV test
B	-	bottoms flowrate
D	-	distillate flow rate
F	-	column feed rate
F_C	-	tuning parameter for simple decoupler
F_{DT}	-	the detuning factor
K_c	-	proportional gain
K_{cu}	-	the ultimate controller gain
K_c^{ZN}	-	the ZN settings for the proportional gain
L	-	reflux flow rate
L_{AC}	-	accumulator level
L_{RB}	-	reboiler level
P	-	column pressure
P_u	-	ultimate period
Q_{COND}	-	condenser duty
u	-	manipulated variable
V	-	vapor boilup rate
x	-	impurity in bottoms product
y	-	controlled variable
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
ff	-	feedforward
o	-	initial value
RB	-	reboiler
sp	-	setpoint

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Table 1
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 2

Design Specifications for Each Case Study

	C ₃ 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 (2332,007#/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 3

Assumptions for Each Column Model

	C₃ Splitter	Xylene/Toluene	Depropanizer
Liquid Dynamic	Hydraulic time	Frances Weir	Hydraulic Time
	Constant	Formula	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 4

Steady - State RGA Values (λ_{11} 's) for the Three Case Studies

<u>Configuration</u>	<u>C₃ Splitter</u>	<u>Xylene/Toluene</u>	<u>Depropanizer Column</u>
(L, V)	2.53	3.37	3.48
(L, B)	0.94	0.92	0.93
(L, V/B)	1.70	1.37	2.09
(D, V)	0.06	0.10	0.07
(D, B)	*	*	*
(D, V/B)	0.12	0.23	0.24
(L/D, V)	1.91	1.54	2.14
(L/D, B)	0.97	0.96	0.91
(L/D, V/B)	1.26	1.16	1.64

*(D, B) structure has infinite steady state RGA since it is open loop unstable.

Table 5

Sensitivity of Manipulated Variable to Feed Composition Changes

<u>Manipulated Variables</u>	<u>Per Cent Changes from Base Case</u>		
	<u>C₃ Splitter:</u> <u>$\Delta Z = +5$ mole%</u>	<u>Xylene/Toluene:</u> <u>$\Delta Z = +5$ mole%</u>	<u>Depropanize:</u> <u>$\Delta Z_{LK} = +5\%$; $\Delta Z_{HK} = -5\%$</u>
L	2.0	0.5	0.6
D	3.7	7.5	8.1
L/D	-1.5	-6.5	-7.9
V	2.1	3.2	3.9
B	-8.4	-15	-19.4
V/B	15.8	22	25.6

Table 6

Control Performance (IAE) for Each Configuration for Setpoint Changes

	<u>C₃ Spittler</u>			<u>Toluene/Xylene Column</u>			<u>Depropanizer</u>		
<u>Configuration</u>	<u>E_{DT}</u>	<u>Overhead</u>	<u>Bottoms</u>	<u>E_{DT}</u>	<u>Overhead x 10²</u>	<u>Bottom x 10²</u>	<u>E_{DT}</u>	<u>Overhead</u>	<u>Bottoms</u>
L, V	6.0	50.4	1,084	2.0	6.6	11.2	3.0	8.5	5.1
L, B	1.5	8.4	132	2.0	7.7	8.2	3.5	10.0	8.9
L, V/B	2.0	46.3	294	1.5	4.5	1.3	3.0	8.3	5.9
D, V	1.5	18.2	455	4.5	18.5	49.5	2.5	150.5	0.3
D, B	1.5	15.4	15	*	*	*	7.0	103.0	8.0
D, V/B	4.0	203.0	1,280	3.2	10.7	3.6	2.5	91.0	1.3
L/D, V	5.0	45.0	6,750	2.5	5.0	6.6	2.0	88.2	0.3
L/D, B	5.0	13.4	333	2.5	5.7	8.8	4.0	70.7	5.1
L/D, V/B	5.0	16.1	201	2.0	5.0	2.2	3.0	39.3	1.3

* Due to poor performance we were unable to obtain results for this configuration.

Table 7

**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 x 10²</u>	<u>Balance x 10²</u>	<u>Overhead</u>	<u>Bottoms</u>
L, V	20.8	505	3.0	12.1	6.9	1.4
L, B	2.2	47	7.8	21.8	4.4	4.6
L, V/B	18.5	124	2.4	7.5	3.7	1.9
D, V	5.4	150	28.1	31.1	936.1	28.9
D, B	3.4	75	*	*	191	28.9
D, V/B	61.6	399	34.7	21.5	176	6.5
L/D, V	15.4	373	3.7	6.6	332	6.1
L/D, B	3.1	161	4.0	23.5	93.4	10.5
L/D, V/B	3.5	115	2.0	15.5	172	8.0

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

Table 8

**Control Performance (IAE) for (L,V) Configuration With and Without
Decoupler for Xylene/Toluene Column**

		<u>Setpoint Changes</u>			<u>Feed Composition Upsets</u>	
		<u>E_{DT}</u>	<u>Overhead</u>	<u>Bottoms</u>	<u>Overhead</u>	<u>Bottoms</u>
Xylene/Toluene Column	Without Decoupler	2.0	0.066	0.112	.03	0.121
	With Decoupler	2.0	0.051	0.035	0.016	0.053
Depropanizer	Without Decoupler	3.0	8.5	5.1	6.9	1.4
	With Decoupler	3.0	7.7	0.1	1.5	0.1

Table 9

**Control Performance (IAE) With and Without Feedforward Compensation
for Feed Composition Changes for C₃ Splitter Columns**

<u>Tylene/Toluene Column</u>	<u>E_{DT}</u>	<u>Overhead</u>	<u>Bottoms</u>
(L/D, V/B) without feedforward	2.0	0.02	0.155
(L/D, V/B) with feedforward	2.0	0.006	0.042
<u>C₃ Splitter with Perfect Level Control</u>			
	<u>E_{DT}</u>	<u>Overhead</u>	<u>Bottoms</u>
(D,B)	1.5	50	117
(D/zF, B/(1-z)F)	1.5	1.6	22

Figure 1 Comparison of open loops test with an ATV test

Figure 2 Single composition control results for overhead composition control for the C_3 splitter

Figure 3 Dual composition control results for C_3 splitter (A) overhead product composition (B) bottom product composition

Figure 4 Manipulated variable action for dual composition control for C_3 splitter (A) reflux flow rate (B) boilup ratio and bottoms product flow rate

Figure 5 Closed-loop Bode plot for C_3 splitter (A) overhead product composition control performance (B) bottoms product composition control performance

Figure 6 Dual composition control results for toluene/xylene column (A) overhead product composition (B) bottoms product composition

Figure 7 Manipulated variable action for dual composition control for xylene/toluene column (A) reflux and reflux ratio (B) reboiler duty and boilup ratio

— [L, V] [L/D, V] - - - - [L/D, V/B] — Setpoint

Figure 8 Dual composition control results for depropanizer
 (A) overhead product composition (B) bottoms product composition

Figure 9 Manipulated-variable action for dual composition control for depropanizer (A) reflux and reflux ratio (B) reboiler duty and boilup ratio