COMPARISON OF ADVANCED DISTILLATION CONTROL METHODS

Second Annual Report

By
James B. Riggs

November 1996

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

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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 1995 through March 1996, and is the second annual progress report for the project. This report and the previous report can be obtained as indicated by the notice inside the front inside cover. The description of the initial report is similar to this report except, the report number is DOE/AL/98747-1 (DE98000103).

ABSTRACT

Detailed dynamic simulations of two industrial distillation columns (a propylene/propane splitter and a xylene/toluene column) have been used to study the issue of configuration selection for diagonal PI dual composition controls. Auto Tune Variation (ATV) identification (Astrom and Hagglund, 1988) with on-line detuning was used for tuning the diagonal proportional integral (PI) composition controls. Each configuration was evaluated with respect to steady-state relative gain array (RGA) values, sensitivity to feed composition changes, and open loop dynamic performance. Each configuration was tuned using setpoint changes over a wider range of operation for robustness and tested for feed composition upsets. Overall, configuration selection was shown to have a dominant effect upon control performance.

Configuration analysis tools (e.g., RGA, condition number, disturbance sensitivity) were found to reject configuration choices that are obviously poor choices, but were unable to critically differentiate between the remaining viable choices. Configuration selection guidelines are given although it is demonstrated that the most reliable configuration selection approach is based upon testing the viable configurations using dynamic column simulators.

ACKNOWLEDGMENTS

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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 control options selected for the representative control configuration 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 creditability. 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

Shinskey (1984) presented a thorough coverage of steady-state relative gain array (RGA) analysis applied to eight sample columns considering ten different configurations. Shinskey concluded that the (L/D, V/B) configuration was 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.

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 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

Two different columns are studied: a propylene/propane column (a C_3 splitter), and a xylene/toluene column. 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.

Each column model assumed that the product composition analyzers had an analyzer delay of five minutes. Tray temperatures were found to correlate well with product compositions for some columns while that was not the case for the C_3 splitter and the xylene/toluene column.

The C_3 splitter used an Euler integrator (Riggs, 1994) with a step size of 0.3 seconds. 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. 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 hour 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) 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) 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 enough to be accurately measured.

If ZN controller settings are applied to a multivariable problem, such as a distillation column, poor controller performance will usually result due to the coupling present in the process. 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). 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} by trial-and-error in order to optimize tuning performance.

Two PI controllers were applied to the xylene/toluene simulator. 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 2.0. The integral absolute error (IAE) of the controller tuned using the BLT method was four times as large as the IAE of the controller tuned using online detuning for the bottom composition and was 50% larger for the overhead composition. 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, would clearly 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, because the user only has to decide if the controller is too aggressive or too sluggish. Moreover, the value of F_{DT} typically ranges between 2 and 6. 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 ZN settings, 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.

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 based upon diagonal PI controllers 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 inserts 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). 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).

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 results 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.

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 will be 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. 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 will be 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). It will identify configurations that are clearly inferior but generally adds little information over steady-state RGA results [except for the (D, B) structure]. In addition, 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 (Shinskey 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 (u,y) configuration selection for the two case studies: a C_3 splitter, and a xylene/toluene vacuum column. 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 identification with optimally selected detuning factors (F_{DT}) based upon setpoint changes. Feed composition upsets were chosen because they represent the most difficult day-to-day operational upset. In addition, as a further test of robustness, each column is tested for a reboiler duty upset. In each case, the tuning was performed for setpoint changes involving 50% changes in the impurity levels, and was based upon minimum Integral Absolute Error (IAE). Setpoint changes using 50% reductions in impurity levels 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. Smaller per cent impurity setpoint changes could be chosen and would result in more aggressively tuned controllers. Clearly, the degree of aggressiveness is determined by the type and severity of the disturbance faced by the controller. The use of 50% impurity setpoint changes for tuning represents conservative tuning that should remain stable for very serve upsets.

C₃ SPLITTER

Following is an analysis of the configuration selection process for the C₃ splitter test case using steady-state RGA analysis, disturbance sensitivity, dynamic behavior, and PI control results.

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, positive and negative open loop step tests were generated for all nine configurations. 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 have the least direct effect due to level control in the accumulator and reboiler, respectively.

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 of five to one occurred for the (L/D, V), (D,V), (D,V/B) and (L,V) configurations while the time constants for the overhead were quite linear for the (L,B) and (L/D,B) configurations. For the bottom, the time constants for positive and negative changes were relatively linear for most configurations.

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. That is, step changes in V and V/B result in a quick high gain change in the bottom composition. But the change in vapor will result in an eventual change in the reflux by the level controller in the accumulator because D is fixed. As the change in reflux works its way through the accumulator and down the column, it will counteract most of the initial change in the bottoms composition; therefore, the overshoot results.

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 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. Likewise, when either B or D is decreased, the impurities decrease.

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.

Control Results. Tables 6 and 7 show the control results for the nine control configurations considered. Table 7 shows the performance index (IAE) for each configuration for a series of setpoint changes in the overhead impurity level. Table 6 also shows the bottoms control performance for each configuration for the setpoint changes in the overhead composition. 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. The detuning factor F_{DT} for each configuration was selected to minimize the IAE for the overhead alone because the overhead product is a feedstock for a polymer reactor and the bottoms product setpoint is based upon economic trade-offs. Note that the (L, B) configuration has the best overall performance while (L, V) has clearly the worst. Note that (D,B), (L/D, V/B), and (L/D, B) have similar overhead and bottoms performance. Figure 1 shows the overhead impurity level for this test for the (L,B) and the (L/D, V/B) configuration while Figure 2 shows the bottom composition for the same test. Figures 3 and 4 show the manipulated variable action for this test. Note the shorter settling time for the (L,B) configuration. Also note that the changes in the reflux flow caused by the setpoint changes in y result in a sharp upset in the bottom product composition. This was due to the high gain inverse

response of the bottom for reflux rate changes in the (L, B) configuration.

The same controller settings developed from the setpoint changes were used to evaluate disturbance rejection performance of each configuration. Table 6 shows the performance of each controller for the overhead and bottoms composition control loops for a ramp change in pressure from 211 to 226 psia over a one hour period starting at $t = 100$ minutes. This was then followed by a step change in feed composition from 70 to 65 mole % propylene at $t = 2000$ minutes. Figure 5 shows the overhead composition for this disturbance test for the (L, B) and (L/D, V/B) configurations. Note that the (L,B) configuration shows the shortest settling time value for the overhead and bottoms products.

The results shown so far are for relatively loose level control, but level control was found to have a significant effect upon control performance for material balance configurations, which is consistent with earlier work (Yang, et. al, 1990). Three different degrees of tightness in level control were considered: (1) perfect level control and (2) a maximum deviation of 2% and (3) a maximum deviation of 6% in the accumulator and reboiler levels (used for the previous results). Figure 6 shows the (D,B) configuration performance for the three degrees of level control for setpoint changes in the overhead composition. It should be pointed out that for perfect level control, the (D,B) performance was essentially equivalent to the (L,B) structure.

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 consistent with the performance indices for the setpoint changes and the disturbance rejection cases. It should be remembered that the model for the C_3 splitter had loosely tuned level controllers for the accumulator and reboiler. If very tight level control is possible, the (D,B) configuration was shown to be as effective as the (L,B) structure.

XYLENE/TOLUENE COLUMN

Following is an analysis of the configuration selection process for the xylene/toluene column using steady-state RGA analysis, disturbance sensitivity, dynamic behavior, and PI control results.

RGA Analysis. The steady-state RGA results (λ_{11} values) for the eight of the possible configurations are listed in Table 3. 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 while 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. It is interesting to note that the (L/D, V/B) configuration tends to exhibit the least coupling while the (L, V) configuration is least affected by feed composition changes which is generally consistent with other lower reflux ratio columns.

Dynamic Analysis. For the dynamic analysis, positive and negative open loop tests were generated 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.

Control Results. Table 6 shows the control performance index (IAE) for each configuration for setpoint changes in the overhead composition. The controllers for each configuration were tuned (i.e., the value of F_{DT} was selected) using setpoint changes for the overhead and bottoms. The (L/D, V/B) and (L, V/B) configurations had the best overall performance. In particular, these configurations greatly outperformed the other configurations with regard to the bottom composition control performance. Figures 7 thru 10 shows the results of the setpoint test for the (L/D, V/B) and (L, V/B) configurations. Figure 7 shows the overhead impurity level and Figure 8 shows the impurity in the bottom product. Figure 9 shows the reflux and distillate flow rates and Figure 10 shows the boilup and bottoms flow for the setpoint tests.

Table 7 shows the performance index for each of the nine configurations for a step change in feed composition from 67% to 62% toluene at time equal 10 minutes. The (L, V/B) configuration had the best overall performance. The (L/D, V/B) and (L, V) configurations did well for the overhead composition but exhibited poorer performance for the bottom product. Control performance plots for the (L, V/B) and (L, V) configurations for the overhead and bottoms products are shown in Figures 11 and 12, respectively.

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. The (L, V) configuration was found to be the least sensitive to feed composition changes and to be the fastest responding configuration. Based upon the feed composition upset test, the (L, V/B) configuration had the best overall performance closely followed by the (L, V), and (L/D, V) configurations.

(L, V) WITH DECOUPLER

The (L, V) configuration has the advantages of being best dynamically and having a relatively low sensitivity to feed composition changes. Unfortunately, it suffers from severe coupling as evidenced by its consistently high RGA values (also see Shinskey, 1984). The (L/D, V/B) configuration has superior RGA values particularly for reflux ratios less than five. It would be highly desirable, therefore, to have an (L, V) configuration with the decoupling characteristics of the (L/D, V/B) configuration.

For example, first consider the effect of boilup rate changes on the overhead. When the boilup rate is increased for the (L, V) configuration, the extra vapor that reaches the top of the column and leaves as distillate shifting the overall material balance of the column. For the (L/D,

V/B) configuration, part of the extra vapor reaching the top of the column would be passed back down the column as reflux. One way to modify the (L, V) structure would be to increase the reflux rate by a fixed fraction of any boilup rate change called for by the bottom composition controller. Note that no dynamic compensation is required in this case because the vapor travels through the column at a relatively fast rate. Moreover, by implementing these reflux rate changes directly, one eliminates the dynamic lag caused by the level controller on the accumulator.

Likewise, adjustments to the boilup rate can be made based upon a fraction of reflux rate changes made by the overhead composition controller. But in this case, dynamic compensation is required due to the liquid dynamic lag down the column. This simple two way decoupler was suggested to us by Beauford (1993) and is a special case of a generalized decoupler proposed by Waller (1992).

Table 8 shows the performance index (IAE) for the setpoint and feed composition changes for the (L, V) configuration with and without the decoupler on the xylene/toluene column. For both the setpoint changes and the feed composition changes, the decoupler had the most dramatic effect upon the control of the bottom product. This decoupler is quite simple to tune. Under open loop conditions, a step change in the boilup rate is made and the fraction of the boilup rate change is adjusted to minimize the effect upon the overhead composition. A factor of 55% was used in this case. For changes in the reflux, it was found that 85% of the reflux change should be added to the vapor rate, and a five minute delay was used for dynamic compensation.

(L/D, V/B) WITH FEEDFORWARD FOR FEED COMPOSITION CHANGES

Xylene/Toluene Columns

The (L/D, V/B) configuration suffers from its sensitivity to feed composition changes. Even though the (L/D, V/B) structure was characterized as the most highly decoupled configuration, it did not perform well in this study for feed composition changes. As a result, the performance of the (L/D, V/B) configuration can be enhanced by the addition of feedforward correction for feed composition changes.

The feedforward approach used here is based upon having the (L/D, V/B) configuration respond to feed composition changes as if it were the (L, V) structure. For example, if the feed composition were to increase in lights, the (L,V) structure would simply pass the extra lights out the overhead as distillate flow while the (L/D, V/B) configuration would put part of the extra overhead

vapor flow back down the columns as reflux to maintain a constant reflux ratio. Therefore, for the (L/D, V/B) structure to maintain a constant reflux flow, the reflux ratio would have to be decreased. For a binary column, the feedforward correction for the reflux ratio is given as where $(L/D)_0$ and z_0 are the initialized values of reflux ratio and mole fraction of light in the feed, z is the current measured value feed composition, and x_{sp} is the setpoint for the bottom composition of the light. This feedforward correction will require dynamic compensation. A similar expression can be developed for a multicomponent column.

The feedforward correction to V/B for feed composition changes based upon keeping the boilup rate constant is given by where $(V/B)_0$ is the initialized values of boilup ratio and y_{sp} is the setpoint for the lights in the overheads.

The results for the (L/D, V/B) configuration on the xylene/toluene column with and without the feedforward compensation are shown in Table 9. Both feedforward compensations were dynamically compensated using a first order exponential filter. The filter factor for the feedforward correction for the reflux ratio was 0.15 and for the feedforward correction for the boilup ratio was 0.95. This results because the saturated liquid feed affects the bottom of the column before the overhead. Note that there is a marked improvement in performance for the overhead and bottoms products. In fact, these results are superior to the results for the (L,V) with decoupler (e.g, the IAE's for the overhead and bottoms are reduced by a factor of two). But it should be remembered that the feedforward compensation requires a feed composition analyzer which many times is not available.

(D/zF, B/[1-z]F) CONFIGURATION

C₃ Splitter

The (D, B) structure was found to provide excellent performance on the C₃ splitter when tight level control was used. The (D, B) structure must make significant changes to D and B by feedback action for feed composition changes. Because for constant product purities, the steady-state ratio of D to B will change significantly for feed composition changes. Assuming that the impurity levels in both products are small, the distillate flow rate is given by zF and the bottom flow rate would be $(1-z)F$; therefore, as a first approximation, normalizing D by zF and B by $(1-z)F$ should provide the majority of the adjustment required in D and B for feed composition changes. As with the (L/D, V/D) with feedforward, dynamic compensation on z would be required in both cases.

Table 9 lists the performance indices for the (D, B) and (D/zF, B/(1-z)F) configuration for

a 5 mole % decrease in feed composition on the C_3 splitter with perfect level control in the reboiler and accumulator. The value of z used for the overhead was filtered with a first order exponential filter value of 0.042 while the bottoms used a filter value of 0.125. Note that the IAE was reduced by a factor of three to five using the $(D/zF, B/(1-z)F)$ configuration. This configuration also requires a feed composition analyzer which is not always available.

CONCLUSIONS

Configuration selection was shown to be a critical element to the control performance for diagonal PI distillation dual composition controls. Moreover, configuration analysis tools (e.g., RGA, condition number, disturbance sensitivity) were found to reject configuration choices that are obviously poor choices, but were unable to critically differentiate between the remaining viable choices. For example, it is clearly demonstrated that although steady-state RGA's are routinely used in industry for configuration screening, RGA's do not correlate well with control performance. Dynamic column simulations represent the best available means of identifying the optimal configuration from among the group of viable configuration choices. When it is not feasible to test configuration choices using a dynamics simulator, the following approach is recommended: when $L/D < 5-8$ use (L, V) with decouplers; when $L/D > 5-8$ use (D, B) with tight level control. When on-line feed composition measurements are available, feedforward compensation can result in substantial control performance enhancements during feed composition upsets.

NOMENCLATURE

a	-	amplitude of the measured variable in an ATV test
B	-	bottoms flowrate
D	-	distillate flow rate
F	-	column feed rate
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

	C₃ Splitter	Xylene/Toluene
Number of Trays	232	49
Feed Tray Location (From Bottom)	64	24
Feed Flow rate	13.44 kg/sec (106,400#/HR)	29.03 kg/sec (232,007#/HR)
Feed Comp. (mole %)		
Lighter than the Light Key		
Light Key	C ₃ - 70	Tol - 67
Heavy Key	C ₃ - 30	Xyl - 33
Heavier than Heavy Key 1		
Heavier than Heavy Key 2		
Heavier than Heavy Key 3		
Factor times minimum reflux	1.3	1.1
Column Diameter	3.96 m (13ft)	3.96 m (13ft)
Overhead Pressure	15.0 atm (221 psia)	0.12 atm (90 mm Hg)
Overhead Product Impurity	C ₂ - 0.3 mole %	Xyl - 0.1 mole %
Bottoms Product Impurity	C ₃ - 2.0 mole %	Tol - 0.1 mole %
Overhead Flow rate	9.21 kg/sec (73,100 #/HR)	18.53 kg/sec (148,091 #/HR)
Overhead Temperature	34.7°C (94.4 °F)	49.5°C (121.1°F)
Bottom Flow rate	4.21 kg/sec (33,400 #/HR)	10.5 kg/sec (83,916 #/HR)
Bottom Temperature	42.3°C (108.1 °F)	118.7°C (245.7°F)
Reboiler Vapor flow rate	131.24 kg/sec (1,041,165 #/HR)	41.87 (334,625 #/HR)
Reflux Ratio	12.6	1.21
Feed Quality	Saturated	Saturated

Table 3

Assumptions for Each Column Model

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

Table 4

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

<u>Configuration</u>	<u>C₃ Splitter</u>	<u>Xylene/Toluene</u>
(L, V)	2.53	3.37
(L, B)	0.94	0.92
(L, V/B)	1.70	1.37
(D, V)	0.06	0.10
(D, B)	*	*
(D, V/B)	0.12	0.23
(L/D, V)	1.91	1.54
(L/D, B)	0.97	0.96
(L/D, V/B)	1.26	1.16

*(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: $\Delta Z = +5$ mole%</u>	<u>Xylene/Toluene: $\Delta Z = +5$ mole%</u>
L	2.0	0.5
D	3.7	7.5
L/D	-1.5	-6.5
V	2.1	3.2
B	-8.4	-15
V/B	15.8	22

Table 6**Control Performance for Each Configuration for Setpoint Changes**

Configuration	<u>C₃ Splittler</u>			<u>Toluene/Xylene Column</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>
L, V	6.0	50.4	1,084	2.0	6.6	11.2
L, B	1.5	8.4	132	2.0	7.7	8.2
L, V/B	2.0	46.3	294	1.5	4.5	1.3
D, V	1.5	18.2	455	4.5	18.5	49.5
D, B	1.5	15.4	15	*	*	*
D, V/B	4.0	203.0	1,280	3.2	10.7	3.6
L/D, V	5.0	45.0	6,750	2.5	5.0	6.6
L/D, B	5.0	13.4	333	2.5	5.7	8.8
L/D, V/B	5.0	16.1	201	2.0	5.0	2.2

* 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>Overhead</u>	<u>Bottoms</u>	<u>Overhead x 10²</u>	<u>Balance x 10²</u>
L, V	20.8	505	3.0	12.1
L, B	2.2	47	7.8	21.8
L, V/B	18.5	124	2.4	7.5
D, V	5.4	150	28.1	31.1
D, B	3.4	75	*	*
D, V/B	61.6	399	34.7	21.5
L/D, V	15.4	373	3.7	6.6
L/D, B	3.1	161	4.0	23.5
L/D, V/B	3.5	115	2.0	15.5

* 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>	
Xylene/Toluene Column		E_{DT}	<u>Overhead</u>	<u>Bottoms</u>	<u>Overhead</u>	<u>Bottoms</u>
	Without Decoupler	2.0	0.066	0.112	0.03	0.121
	With Decoupler	2.0	.051	0.035	0.016	0.053

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>			
(D,B)	<u>E_{DT}</u>	<u>Overhead</u>	<u>Bottoms</u>
	1.5	50	117
(D/zF, B/(1-z)F)	1.5	1.6	22

Figure 1 Top Impurity for Setpoint Changes for C₃ Splitter

Figure 2 Bottom Impurity for Setpoint Change for C_3 Splitter

Figure 3 Reflux and Vapor Flow for Top Impurity Setpoint Changes for C_3 Splitter

Figure 4 Distillate and Bottom Flow for Top Impurity Setpoint for C₃ Splitter

Figure 5 Top Impurity for Pressure and Feed Composition Changes for C₃ Splitter

Figure 6 Top Impurity for Setpoint Change Using Different Degrees of Tightness in Level Control with the (D,B) Configuration

Figure 7 Top Impurity for Setpoint in Top Impurity for
for Xylene/Toluene Column

Figure 8 Bottom Impurity for Setpoint Change in Top Impurity for Xylene/Toluene Column

Figure 9 Reflux and Distillate Flow for Setpoint Change in Top Impurity for Xylene/Toluene Column

Figure 10 Reboiler Duty and Bottoms Flow for Setpoint Change in Top Impurity for Xylene/Toluene Column

Figure 11 Top Impurity for Negative 5% Step Change in Feed Composition for Xylene/Toluene Column

Figure 12 Bottom Impurity for Negative 5% Step Change in Composition for Xylene/Toluene Column