COMPARISON OF ADVANCED DISTILLATION CONTROL METHODS

Final Technical Report
April 1994 - March 1999

By

Dr. James B. Riggs

November 2000

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By
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Lubbock, Texas

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ABSTRACT

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

Proportional Integral (PI) control performance was used to evaluate the configuration selection problem. For single ended control, the energy balance configuration was found to yield the best performance. For dual composition control, nine configurations were considered. It was determined that the use of dynamic simulations is required in order to identify the optimum configuration from among the nine possible choices.

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

MPC was applied to three industrial columns: one at Phillips Petroleum and two at Union Carbide. In each case, MPC was found to significantly outperform PI controls. The major advantage of the MPC controller is its ability to effectively handle a complex set of constraints and control objectives.

PREFACE

The objective of this project was to evaluate the benefits of advanced distillation column control strategies in comparison with conventional control strategies. Simulations of several industrial columns were studied and used as the basis for comparison between conventional and advanced control during Phase III. In addition, the effect of the control configuration on control performance was tested. Phase IV involved the industrial comparison between advanced control and conventional control. Earlier work is described in four prior "Comparison of Advanced Distillation Control Methods, Technical Annual Report(s)", numbered DOE/AL/98747-1, -2, -3, and -4, respectively. These prior reports are recommended reading for those desiring a somewhat more detailed treatise of the technology described herein.

ACKNOWLEDGMENTS

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1. <u>INTRODUCTION</u>

1.1 <u>Background</u>

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, frequently industry over-refluxes their columns in order to ensure that the product purity specifications are met. That is, they use 30 to 50% more energy than actually necessary to meet the product specifications and 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 options. As a result, whether or not to apply advanced distillation control, what type of advanced control to apply, and how to apply it are usually determined based upon internal company politics and informal 'word-of-mouth'. In fact, when industry discusses advanced control, they refer to taking a "leap-of-faith". Because advanced control is not well 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.

1.2 Distillation Column Nomenclature

1.2.1 Nomenclature for a two product column

Figure 1 shows a schematic of a two-product distillation column. The feed (F) enters the column with a feed composition z, and the column separates the feed into a lighter boiling overhead product (D) which has a composition y, and a heavier boiling bottoms product (B) which has a composition x. In order to effect a separation between the light component in the feed and the heavy component, vapor (V) is boiled up the column from the reboiler, and liquid reflux (L) is returned to the top of the column. Vapor and liquid are contacted on each of the trays that compose the interior of the column, concentrating the light components in the vapor

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leaving the top of the column and concentrating the heavy components in the liquid leaving the reboiler.

Steam (S) provides the heat to boilup the vapor (V) from the reboiler. Cooling water (CW) removes heat from the overhead vapor in order to condense it into a liquid that is collected in the overhead accumulator. The liquid that is withdrawn from the accumulator is returned to the column as reflux (L), or leaves as the overhead product (D).

The instrumentation for this column includes level transmitters (LT) for the reboiler and the accumulator that indicate the measured level, composition analyzers (AT) on the overhead and bottoms product that provide online product composition analysis. The overhead of the column also has a pressure transmitter (PT) that indicates the overhead column pressure. The valves that are located on the steam, bottoms product, reflux, and overhead product represent flow control loops that maintain flow rates at specified levels. That is, the column controllers specify these four flow rates to control the product compositions and the levels in the accumulator and the reboiler.
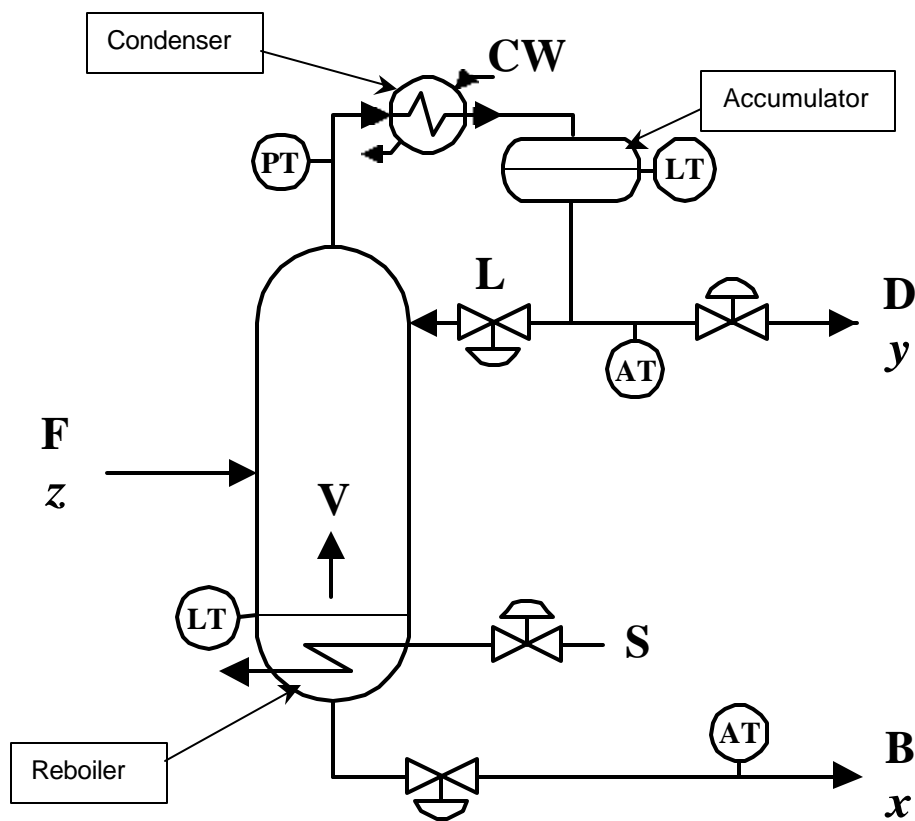
Figure 1 Schematic of a two-product distillation column.
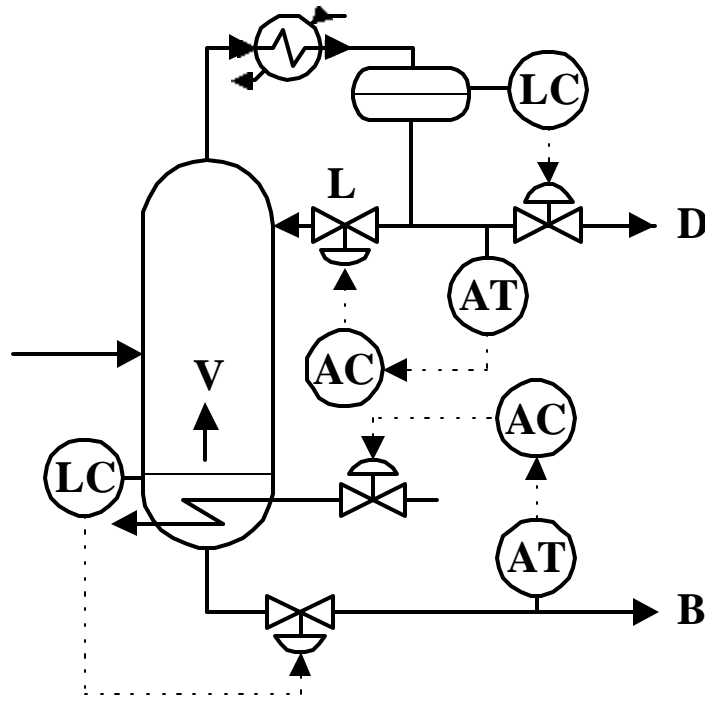
Figure 2 Schematic of the (L,V) configuration

1.2.2   Nomenclature for distillation control configurations

The nomenclature used here refers to a particular configuration as (c1, c2) where c1 is assumed to be the controller output that is used to control the overhead composition, and c2 is the controller output that is used to control the bottoms composition. If we limit ourselves to controlling the overhead composition with L, D, or L/D (the reflux ratio) and the bottoms composition with V, B, or V/B (the boilup ratio), there are a total of nine possible configurations: (L,V), (L,B), (L,V/B); (D,V), (D,B), (D,V/B); and (L/D,V), (L/D,B), (L/D,V/B).

As an example, Figure 2 shows the (L,V) configuration. Note that the reflux flow rate is used to control (AC) the overhead composition and the boilup rate (i.e. the heat duty to the reboiler) is used to control (AC) the bottom product composition. As a result, the distillate product rate (D) is used to maintain the level in the accumulator, and the bottoms product rate (B) is used to control the liquid level in the reboiler. Therefore, the manipulated variables used to control the compositions determine the remaining manipulated variables that are used to control levels (LC).

1.2.3　Nomenclature for the main fractionator

A schematic of a main fractionator is shown in Figure 3. Feed enters near the bottom of the column and decant oil, light cycle oil, heavy naphtha, light gas liquids, and light gases are removed as products. Heat is recovered from this column from the HCN (heavy cycle naphtha) pumparound, the LCO (light cycle oil) pumparound, and the HCO (heavy cycle oil) pumparound. That is, other process steams are heated using these heat exchangers. Products are withdrawn from the naphtha and the light cycle oil strippers and vapor is returned to the main column. The slurry pumparound vaporizes the decant oil providing vapor for the column. Steam is injected at the bottom of the column to strip out light gases from the feed.
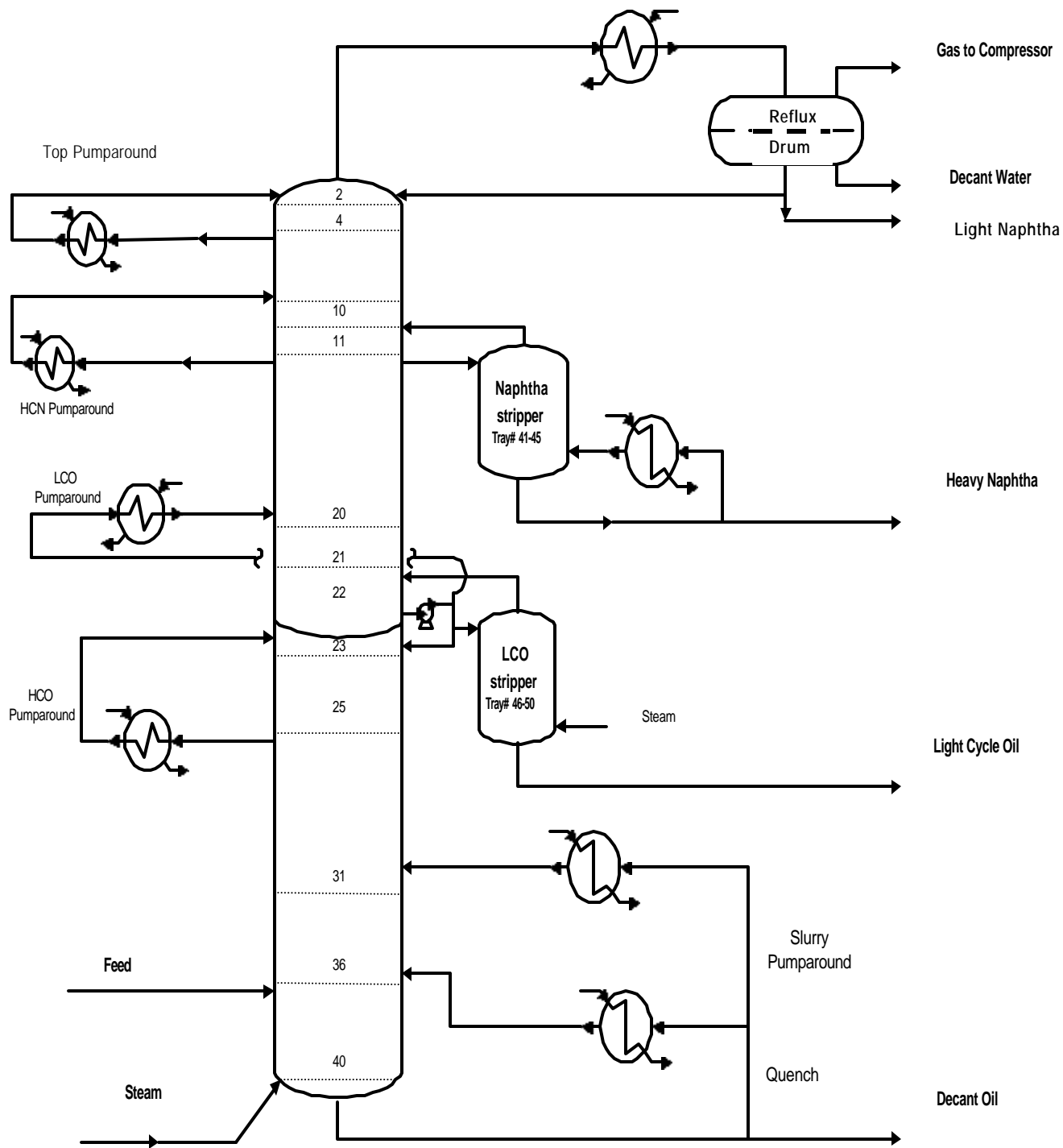
Figure 3 Schematic of a main fractionator.

2. <u>CASE STUDIES AND DYNAMIC MODELS</u>

2.1 Overview of the columns studied

Four different columns are studied: a propylene/propane column (a C_3 splitter), a xylene/toluene column, a depropanizer, and a main fractionator. The specifications for the first three columns are listed in Table 1, while the specifications for the main fractionator 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). The main fractionator has one feed, five products, and six pumparounds (Figure 3). Fractionators are a type of distillation column used to separate a wide boiling mixture of components into a number of separate products. This group of four columns should be representative of 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 the first three dynamic column models. Table 4 list the assumptions used for the model of the main fractionator.

2.2 Vapor/liquid equilibrium models used

Vapor/liquid equilibrium (VLE) models describe the tendency of the lighter components in the liquid to concentrate in the vapor, and are used to model the separation provided by each tray in the column. The VLE description for each case study was different. For the C_3 splitter, the VLE was described using the relative volatility, which is a measure of how easily the light component separates from the heavy component; i.e. the greater the relative volatility is above unity, the easier the mixture is to separate. The VLE for the C_3 splitter 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 volatility was 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-Redlich-Kwong model (SRK; Soave, 1972) for the component K-values. (The K-value is the ratio of the mole fraction of a component in the vapor phase to the mole fraction of the same component in the liquid phase, and is a measure of the tendency of the component to vaporize.) Because the SRK method requires an iterative solution procedure, an empirical correlation for the K-values (Boston and Sullivan, 1974) was used in order to reduce the computational overhead. The empirical correlation for the K-values was reparameterized using the SRK model every 10 seconds of simulation time, or if a tray temperature changed by more than 1.0°C since the last time it was reparameterized. The relative volatility for the depropanizer was observed to range from 1.5 at the top to 1.9 at the bottom.

2.3 Composition measurements and estimates

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 that were used to estimate the product composition for the depropanizer used the following functional form

$$\ln x = a + b/T \qquad\qquad (1)$$

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.

2.4 Numerical integration and computational efficiency

The C_3 splitter and the depropanizer used a Euler integrator (Riggs, 1994) with step sizes of 0.3 seconds and 0.2 seconds, respectively. The ratios of simulated time to CPU time for a 66 MHz 486 PC (using Microsoft FORTRAN 5.1) were 50:1 for the C_3 splitter and 15:1 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 ratio of 7:1.

2.5 Benchmarking of the models

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 qualitatively to check the model against the industrial data. Next, estimated industrial response times (an eight hour response time for the overhead composition for a 0.5% step 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 constant of 3.0 seconds provided the best overall fit. Finally, the xylene/toluene model was found to match the results presented by Choe and Luyben (1987).

2.6 Main fractionator VLE and numerical integration

For the main fractionator, the VLE K-values and enthalpies were calculated via an approach similar to that for the depropanizer; i.e. the SRK equation was used to update an empirical correlation. The empirical correlation was updated every five minutes, or if a tray temperature changed by more than 1.0°C. An algorithm called the *dynamic stagewise adiabatic flash* (DSAF) algorithm was used to solve the dynamic main fractionator model (Chung and Riggs, 1995). At each time step, an implicit numerical integrator (Newton-Raphson method) was applied to each stage separately while the latest conditions for the liquid and vapor streams entering the stage was used. The time step was 1.8 seconds, and the simulator ran more than 3 times faster than real time on a Pentium 300 MHz machine.

2.7 Product composition estimates for the main fractionator

Inferential models are used extensively in industry for the endpoints and API gravity of products from the main fractionator. They are normally rigorous models based on temperature, flow, and pressure measurements, and are fairly accurate. To avoid the complexity of modeling, a perfect inferential model was assumed in this study. However, to model the dynamics of the temperature measurements, a first-order filter delays the inferred properties before they are used as measurements for control. The time constant for the filter was 15 seconds.

3. <u>IMPLEMENTATION APPROACH FOR EACH CONTROLLER</u>

3.1 General discussion

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

3.2 Conventional controls

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

Figure 4 on page 10 graphically demonstrates the ATV method. The user must select h, the relay height used or the change in the manipulated variable that is to be applied. h should be small enough that the process is not unnecessarily upset, yet large enough that the resulting amplitude, a, can be accurately measured. Then each time controlled variable y_S crosses y_0 (the initial value of y), the controller output is switched from $c_0 + h$ to $c_0 - h$, or from $c_0 - h$ to $c_0 + h$ (where c_0 is the initial value of the controller output c. A controlled variable constant (unity) amplitude ratio is established after 3 to 4 cycles; thereafter, the values of a and the ultimate period, P_u , can be measured directly and the ATV test is concluded. The ultimate gain, K_u , is calculated by

$$K_u = \frac{4h}{\pi a} \qquad\qquad (2)$$

K_u and P_u can be used in the Ziegler-Nichols (ZN) ultimate settings:

$$K_c^{ZN} = 0.45 \ K_u$$
$$t_I^{ZN} = P_u / 1.2$$

(3)

where t_I is the reset time.
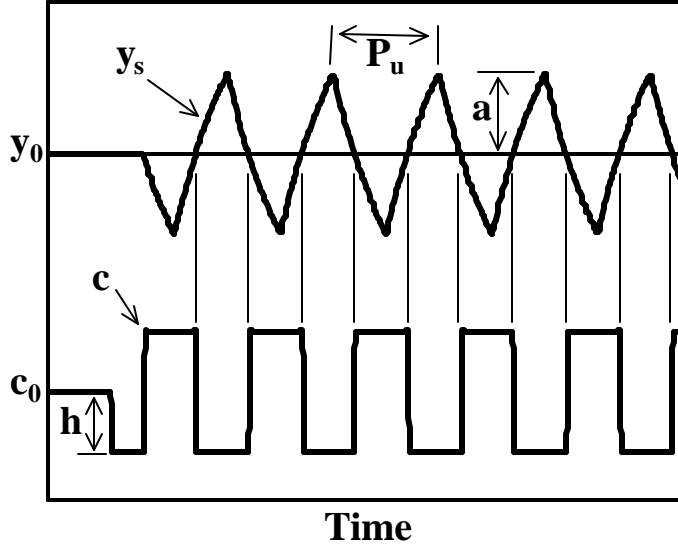
Figure 4 Graphical representation of an ATV test

After the ZN settings are calculated, they may require online tuning, particularly for the settings required to achieve the desired dynamic response. For example, the ZN settings would be tuned online as follows:

$$K_c = K_c^{ZN} / F_T$$
$$t_I = t_I^{ZN} \times F_T$$

(4)

by adjusting tuning factor F_T. Note that as F_T is increased, gain K_c decreases while time t_I increases by the same proportion (i.e., detuning). F_T can be adjusted to meet the performance requirements for each individual application. Therefore, online tuning is reduced to a one-dimensional search for the proper level of PI controller aggressiveness. If the controller is too aggressive, F_T is increased.   If the controller is too sluggish, F_T is decreased. In this work, the performance specification selected for tuning was the minimum absolute value of the integral of the error from setpoint (i.e., the minimum IAE).

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3.3 Model Predictive Control

The MPC applications for all the columns were done using Dynamic Matrix Control (DMC), which is an industrially popular version of MPC. The MPC applications for the C_3 splitter, the xylene/toluene column, and the depropanizer were done using DMC 5.0, which was provided to us by DMC Corporation. In 1996, AspenTech purchased DMC Corporation. When we began the main fractionator studies, AspenTech gave us a site license for DMCplus, which is their commercial MPC controller and an updated version of DMC 5.0.

MPC controllers use step response models that define how each input affects each output of the process. The step response models for the MPC controllers were developed for each input [i.e. mole fraction light component in the feed (z), feed flow rate to the column (F), reflux flow rate (L), and bottom product flow rate (B)]/output pair (x,y). The output for the high purity products were log transformed in an effort to linearize the overall process behavior; e.g.

$$y' = \log(1 - y) \tag{5}$$

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

3.4 Nonlinear Process Model-Based Control

The nonlinear PMBC controller using tray-to-tray models was applied using the approach presented by Riggs et al. (1993). The control law calculates target setpoints (x_{ss}, y_{ss}) based upon proportional and integral feedback. That is, the target setpoints are calculated according to the following equations (Riggs et al., 1993):

$$x_{ss} = x + K_{11}[x_{sp} - x] + K_{21}\int[x_{sp} - x]dt \tag{6}$$

$$y_{ss} = y + K_{12}[y_{sp} - y] + K_{22}\int[y_{sp} - y]dt \tag{7}$$

Then, the values of x_{ss} and y_{ss} are used by the tray-to-tray model to calculate the energy input to the column. Because Equations (6) and (7) can result in values of x_{ss} that are negative and values of y_{ss} that are greater than 1.0, limits are used to restrict the maximum and minimum values of x_{ss} and y_{ss}.

An overall material balance was used to calculate the value of the bottoms flow rate. Since the column responds much faster to energy changes (e.g., reflux flow) than to material balance changes (e.g. bottoms flow), the target product compositions (x_{ss} and y_{ss}) were applied by

using K_{xMB} (an additional tuning parameter) according to the following equations in an effort to improve the response of the bottom product composition (Riggs et al., 1993).

$$x_{MB} = x_{sp} + K_{xMB}[x_{ss} - x_{sp}] \tag{8}$$

$$y_{MB} = y_{sp} + K_{yMB}[y_{ss} - y_{sp}] \tag{9}$$

Then

$$B/F = \frac{y_{MB} - z}{y_{MB} - x_{MB}} \tag{10}$$

The feed rate and the feed composition used by the model were each dynamically compensated using a first order lag and a dead time. When a feed compensation analyzer is not available, the product purities and product flow rate are used to calculate an estimate of the feed composition. This back-calculated feed composition was filtered, and the filtered value was used as the feed composition by the tray-to-tray steady-state model.

When Equations (6) and (7) are used for setpoint changes, the resulting changes in the manipulated variables are much too sharp; therefore, a filter on the setpoint changes was used to stabilize the controller for setpoint changes. The nonlinear PMBC controller was tuned for setpoint changes based upon minimizing the IAE for the overhead product.

The tray-to-tray steady-state controller model used by the nonlinear PMBC controller used the relative volatility modeled as a function of liquid composition and pressure, but used a stage wise tray efficiency, while the dynamic simulator used a Murphree tray efficiency.

3.5 Neural network-based control

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

4. CONFIGURATION SELECTION

4.1 General discussion

A major degree of freedom (DOF) in designing a distillation control system is the choice of the manipulated/controlled variable pairings. For a two product column, there are, in general, five choices of controlled variables (x, y, L_{AC}, L_{RB}, and P), and five manipulated variables (D, L, V, B, and 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 numerous possible manipulated/controlled variable pairings.

In practice, the choice of manipulated/controlled variable pairings for distillation control is much more limited. First, condenser duty (Q_{COND}) is usually set and not directly manipulated (e.g. refinery columns typically operate at maximum condenser duty to achieve 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, however, exceptions to this rule; e.g. Shinskey (1984) recommends such an arrangement [i.e. (L/D,D)] for a very special class of columns.

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 and disturbance sensitivity characteristics that represent a major factor in the configuration selection process.

4.2 Implementation issues

At this point, it should be emphasized that the four manipulated variables (L, D, V, and B) should be implemented as ratios to the feed flow rate (i.e. L/F, D/F, V/F, and B/F). This is because, for a column operating at a constant overall tray efficiency, L, D, V, and B will scale exactly with feed flow rate. Note that in each case, the feed rate used is dynamically compensated. This approach will greatly reduce the size of the upsets caused by feed flow rate changes. Skogestad et al. (1990) 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, L, set as a reflux ratio times D. However, 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 summation of L and D [i.e. (L+D)$_{sum}$]; then D and L can be calculated as

$$D = \frac{1}{L/D+1}[L+D]_{sum} \tag{11}$$

$$L = \frac{L/D}{L/D + 1}[L + D]_{sum} \qquad (12)$$

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. As a result, changes in the reflux ratio, L/D, called for by the composition controller do not significantly affect the level in the accumulator. We have observed superior composition control performance for this approach to reflux ratio control when compared with the previous approach. V/B control can be implemented in a similar manner in order to decouple the bottom 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.

4.3 Configuration selection results

Here the configuration selection problem was evaluated based upon the performance of each configuration using a conventional PI controller. For both the single-ended and dual composition control cases, the composition controllers were tuned for setpoint changes in the impurity levels in the products. More specifically, each of the product composition controllers was tuned for the minimum in the absolute value of the integral error from setpoint (i.e. minimum IAE). After the composition controller was tuned, the control performance was determined by testing the controller response to a step change in the feed composition. Summarizing, each configuration was tuned for product composition setpoint changes and tested for feed composition upsets.

4.3.1 Single-ended control

The configuration selection problem for single composition control (i.e. controlling the composition of only one product) was evaluated by comparing L, D, and L/D for overheard composition control, and by comparing V, B, and V/B for bottoms composition control. For example, when L is used to control y, V is fixed, and when B is used to control x, L is fixed. In each case, the controllers were tuned for setpoint changes in the product impurity level and tested for feed composition upsets. Table 5 lists the IAE's in units of mole fraction-minutes for the overhead and bottoms product composition for each of the configurations and for each of the three columns considered. Note that the reflux, L, consistently provided the best control performance for the overhead product while the boilup rate, V, provided the best overall performance for the bottom product. That is, the (L,V) configuration performed best for single-ended control.

4.3.2 Dual composition control

Table 6 shows the control performance for each of the nine configurations considered for each of the three columns in response to a step change in the feed composition (F_T). For the C_3 splitter, the (L,B) and (L,V/B) configurations provided the best overall performance. For the xylene/toluene column, the (L/D,V) and the (L,V/B) configurations provided the best overall performance. For the depropanizer, the (L,V/B), (L/D,V), and (L/D,V/B) configurations provided the best overall performance. While guidelines can be given for general selection

procedures resulting in <u>reasonable</u> configuration selections, there are currently no reliable procedures to determine the <u>optimum</u> configurations without resorting to complex control comparisons using detailed dynamic simulators as was done in the study. This shortcoming is somewhat disturbing, as there is usually considerable difference between a reasonable configuration and an optimum one. For example, the depropanizer column (L,V) configuration is a reasonable choice and is usually employed industrially, but its performance is considerably poorer than the optimum configurations. Accordingly, the difference between a reasonable and optimum configuration selection may be manifested as lost process efficiency, with the attendant wasted energy and sub-optimal product purity.

4.4 Configuration selection for the main fractionator

As shown in Table 7, the main fractionator has a large number of manipulated variables (MVs) and controlled variables (CVs), and a very large number of combinations of control configurations including ratio schemes. While it is prohibitive to examine all possible configurations, industrial practice and results of previous studies can provide guidelines to select the most reasonable pairings.

In most industrial cases, the middle pumparound duties are set by a higher level optimizer, which considers both the main fractionator and downstream units that use these pumparounds as reboiling media. Stripping steam flows have no significant effect on product quality as long as they are large enough; therefore, they are typically controlled manually by an operator. The top reflux is normally too small to use it as a manipulated variable, while the vapor distillate is normally set by the maximum compressor capacity.

Table 7 details the MV/CV pairings used by this study, which is a popular configuration used by industry, although it is not the only one. (Note that, with the exception of the 'Bottom Temperature' CV, the controlled variable and manipulated variable that are on the same line indicate pairing between them). In this configuration, whenever an energy balance manipulated variable is available, it is used to control the product quality; e.g. LCO reflux (L_{22}) and quench pumparound duty (Qp_6) control the separation between the LCO and the slurry. This is consistent with results of previous studies on single columns; i.e. an energy balance type configuration should be used for low reflux (high relative volatility) columns.

Conventional PI control, PI with a simple decoupler, and MPC control were applied to the main fractionator. All three controllers used the same configuration shown in Table 7. The simple decoupler uses the total flow of the overhead liquid product and the HCN ($D+ F_{P1}$) to control the endpoint of the HCN (EP_3). Each controller was tuned for setpoint changes of 5°F in the overhead liquid product and the HCN endpoints (EP_1 and EP_3), then tested with step changes to lighter and heavier feeds. The lighter feed was simulated by an increase of 1% in the mole fractions of components which makeup the lighter half of the feed, and by decreasing the fractions of heavier components by the same magnitude. The heavier feed was simulated in a similar manner.

ATV tests with online tuning were applied to tune the PI controller and the PI controller with the simple decoupler in the same manner as was applied to the other columns.

An interface was developed to connect the simulator with the MPC controller. The MPC controller is a 4x5 system with 4 MVs (Qp_1, Fp_1, L_{22}, and Qp_6) and 5 CVs (EP_1, API_2, EP_3, EP_4, and T_{40}). The step response models used in the MPC controller have a time to steady state of 3 hours with 150 coefficients. The simulator runs with MPC software at a speed of 3:1 simulation to real time ratio on a Pentium 300 MHz machine.

5. COMPARISON OF CONVENTIONAL AND ADVANCED CONTROLLERS

Table 8 lists the control performance comparisons for the conventional PI controller, the MPC controller, and the PMBC controller. For the C_3 splitter, the MPC controller provides the best control performance for the overhead product, which is the most important control objective, but generally poorer control performance for the less important bottom product. The PMBC controller is not quite as effective as the MPC controller for the overhead product, but does a much better job on the bottom product.

For the xylene/toluene and depropanizer cases, the PI controller clearly outperforms the MPC and PMBC controllers. For the xylene/toluene case, the impurity level is 0.1% in both products, and it was assumed that both products are equally important. For the depropanizer, the impurity level is 0.5% in both products, and once again it was assumed that both products are equally important. It appears the reason for this behavior is that the MPC and the PMBC controllers use multivariable models of the process to calculate control action. Due to the nonlinearity of high purity distillation columns, such as the xylene/toluene and depropanizer columns, the process gain and dynamic behavior of the process can change significantly due to changes in the operating conditions. As a result, when MPC and PMBC are applied to such a process, the errors in the models become amplified since all the process models are used to calculate the control action. On the other hand, for a PI controller, while it too is affected by process nonlinearity, it is less susceptible than the multivariable controllers because its control action is based on the error from setpoint and not a multivariable model of the process.

Another design of the C_3 splitter was studied in which both impurity levels were set at 2.0% and both products were equally weighted. The comparison between PI and MPC are shown in Table 9. Note for this case, PI performs better than MPC.

Table 10 shows the results for a different xylene/toluene column with the overhead impurity set at 2.0% and the bottom impurity at 0.1% with control of the bottom selected as the highest priority. Note that in this case, the MPC controller provides the best control of the bottom product, which was the primary control objective in this case.

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

The PMBC controller did not appear to offer advantages over the PI and MPC controllers. Even for the high purity case, for which one would have expected the PMBC controller to perform best, PMBC was generally inferior to PI. Because the ANN controller is simply a PMBC controller with a neural network steady-state model, ANN also did not offer any advantages over PI or MPC.

Table 11 shows the results for the control studies for MPC and PI controls applied to the main fractionator. Both PI controls and PI control with a simple decoupler were tested. The

MPC controller outperformed the PI controller by a wide margin with up to a 4:1 reduction in the IAE for EP_1. Note that since EP_1 and EP_3 were the most important product specifications, the MPC controller showed a better relative performance compared to the PI controllers. The PI controllers with a simple decoupler showed a performance improvement over the conventional PI controllers.

6.0 INDUSTRIAL DEMONSTRATIONS (PHASE IV)

6.1 Phillips Petroleum propane/isobutane splitter

We participated in the application of MPC to a propane/isobutane splitter (C_3/iC_4 splitter) in Phillips' Borger (Texas) refinery. The original PI controls were implemented and maintained by the same technical staff that participated on the MPC project.

6.1.1 Propane/isobutane process description

The C_3/iC_4 splitter is in the portion of the Borger refinery that is fed by natural gas liquids (NGL). The feed to the splitter is mostly isobutane, with 5% to 20% propane, at a total feed rate of 4,000 to 11,000 barrels per day. The major disturbances to this column are feed composition upsets (i.e. the 5% to 20% variation in propane content) and feed flow changes (i.e. the 4,000 to 11,000 barrels per day variation). The primary control objective is to keep the propane content in the isobutane bottoms product below 1%, and the secondary objective is to keep the isobutane content in the propane overhead product below 5%.   In addition, due to the variations in the propane content in the feed and the subsequent variation in the distillate flow rate, maintaining the accumulator level between appropriate limits was a serious operational challenge for the original PI control system. The bottom product is equipped with an online gas chromatograph for measurement of the propane content in the isobutane product. The overhead composition was inferred from the temperature of the 10th tray from the top of the column. The overhead condenser became a constraint during the summer season.

6.1.2 Propane/isobutane project summary

During the implementation of the MPC controller, some of the identified models were determined to be too inaccurate for control purposes. Further evaluation and testing was required, but finally sufficiently accurate models were obtained. The temperature controller for the tray temperature in the rectifying section of the column was left in service, and the MPC controller was allowed to select its setpoint. The implementation of the MPC controller resulted in a 20% reduction in energy usage for this column due to improved composition control, which, in turn, allowed greater processing rates.

6.2 Union Carbide precooler

We participated in the application of MPC to a precooler distillation column in a Union Carbide chemical plant. The original PI controls were first checked and verified to be in proper operating condition. Then, MPC was applied and tested.

6.2.1 Precooler process description

In this particular plant, there are 20 furnaces that take raw materials (e.g. ethane and propane) and recycled streams and crack them into ethylene and other by-products such as hydrogen, acetylene, propylene, and C_{4+} materials. The separation section consists of a series of distillation columns to separate these products.

The furnace effluent is a hot vaporous stream that is first combined with a number of refinery gas streams, then compressed to a pressure that is suitable for condensation by a refrigeration system. The combined stream then passes through a pre-condenser en route to a distillation column, called the precooler. So, the precooler is the first separation unit after the ethylene furnaces, and separates C_{4-} materials from C_{5+} materials.

The column has an internal condenser at the top section. The overhead product passes through a flash tank. The liquid from the flash tank, which is manipulated by the tank level controller, goes back to the column top as reflux. The vapor goes to a dryer downstream for further processing. The composition of the vapor is measured every 5 minutes using an analyzer. The cooling duty to the condenser is used by the operators as the primary manipulated variable to control the C_5 in the vapor stream. An external reflux, which is drawn from a propylene product stream at the far end of the separation section, was used as the secondary manipulated variable (but the use of this stream should be minimized due to economic considerations).

The bottom stream of the column passes through two flash tanks at different pressures to recover some C_{4-} materials. These recovery streams are recycled back to the feed and combined with the furnace gas after the compressor.

The major disturbances encountered by the operators include:

1. Composition changes in refinery gases. The refinery gas streams contains mostly C_{3-} materials, and account for a significant fraction in the feed. Therefore, refinery gas composition changes are severe disturbances to the precooler operation.

2. Among these refinery gas streams, one stream is called a gypsy flow, which exhibits intermittent behavior. This can have intermittent adverse impact on the column.

3. The pre-condenser cooling capacity is limited. Therefore, the feed temperature is not fully controlled all the time.

4. Other disturbances include raw material and recycle flow changes to the furnaces.

Due to these disturbances, PI controllers were not able to fully stabilize the system, and the process equilibrium relies mainly on operator expertise.

6.2.2 Precooler project summary

A MPC controller with 10 independent variables and 8 dependent variables was designed and implemented to stabilize the precooler operation. There are only two manipulated variables: the external reflux and the condenser coolant side pressure setpoint. Other independent variables are all measured disturbances. The following instrument modifications or calculations were made to handle the disturbances:

1.   Variation in the composition of the refinery gas, especially the C_3 components, represents a significant disturbance to this column. However, the analyzer used to measure this composition was shared with other streams, and the C_3 mole fraction was only available every 8 hours. This period rendered the composition measurement essentially useless for feedforward control. Hence, the analyzer was rescheduled to have the C_3 in the refinery gas measured every 15 minutes, and this measurement and the refinery gas flow rate were used in MPC controller as feedforward variables.

2.   The gypsy flow could not be directly measured due to various operational factors. Therefore, this flow was inferred from the measured control valve position and the valve design flow characteristics. The calculated flow was then used as a feedforward variable in the MPC controller.

3.   Since the feed temperature controller was saturated 90% of the time, the actual temperature measurement was used as a feedforward variable instead of the temperature setpoint.

4.   A tray temperature above the feed tray was included in the MPC controller as a controlled variable since it is a fast indicator of flooding and C_5 composition in the overhead product, as suggested by the operators. High and low limits were specified for this temperature.

Except for the above changes, all regulatory loops remained unchanged. Other major controlled variables include the C_5 composition in the overhead product, pressure drop across the column (an indicator of flooding), bottom tray temperature, valve positions of the condenser coolant outlet and inlet, as well as the reflux valve position.

The step tests for the controller were finished in two weeks. About another two weeks were spent on building the models, and offline and online tuning and adjustments.

The MPC controller was successfully commissioned, and was able to stabilize the column and maintain the C_5 composition in the overhead product in the range between 300PPM and 500PPM. This variation range had previously been from 200PPM to 800PPM before the MPC controller was implemented, representing a 400PPM improvement in C_5 impurity variation, and a 100PPM improvement in average C_5 impurity level. The major benefit was derived from the feedforward power of MPC.

6.3 Union Carbide prefractionator

We participated in the application of MPC to a prefractionator column in a Union Carbide chemical plant. The original PI controls were first checked and verified to be in proper operating condition. Then MPC was applied and tested.

6.3.1 Prefractionator process description

As discussed in Section 6.2.1, the Union Carbide facility operates 20 cracking furnaces that produce a variety of light hydrocarbon species. These species are first directed to the precooler for separation of the C_{4-} and C_{5+} fractions. The precooler overhead product (C_{4-}) components, ranging from methane to butane, are then fed to the prefractionator, the second column in the separation train.

The prefractionator produces an overhead product that contains mostly methane, ethane, ethylene, and small amounts of propylene. The bottom product contains the heavier components, with a small amount of methane. The overall objective of this column is to maintain a specified upper limit on methane in the bottoms product, while minimizing the propylene fraction in the overhead product. In order to minimize the propylene concentration in the overhead product, the column must be operated at its maximum processing rate. Column flooding is the operational constraint for this column, and a flooding condition is inferred from the pressure drop across the column. Maintaining the column differential pressure just below the onset of flooding corresponds to the maximum separating power for this column.

The controlled variables for this column include the methane concentration on the 24th tray, the propylene concentration in the overhead product, the column side temperature, and the column differential pressure. The manipulated variables include the steam pressure to the reboiler, the column feed temperature, the condenser pressure, and the column pressure. There are two feedforward variables for this column, each of which represents a parameter considered to exhibit the potential for major system disturbances. Those variables are the feed rate to the precooler that precedes the prefractionator, and the C_5's concentration in the feed to the precooler. Since the prefractionator is susceptible to flooding due to feed flow and feed composition changes, and since it takes some time to prepare the prefractionator for an increase in load in order to prevent flooding, the feed rate and composition to the precooler, instead of the feed rate and composition to the prefractionator, were used as a feedforward variables in order to give the control system extra time to prepare the column for a load change.

6.3.2 Prefractionator project summary

After testing and evaluation of the existing PI controls, it was determined that the standard deviation of the methane concentration on the 24th tray was 0.35 mass fraction, and that the average mass fraction of propylene in the overhead was 0.19%.

A MPC controller with 6 dependent variables and 6 independent variables was designed and implemented on the prefractionator. After implementation of the MPC controller, the standard deviation in the methane concentration on the 24th tray was reduced from 0.35 to 0.17,

and the average mass fraction of propylene in the overhead was reduced from 0.19% to 0.12%. At the time of this writing, the MPC controller had been in service for 10 months with excellent control performance and a 98% service factor.

6.4 Energy and waste benefits

The following is a discussion of the estimated energy savings and waste reduction for each of the industrial columns considered in the industrial demonstration phase.

6.4.1 Phillips propane/isobutane column project benefits

Based on the 20% reduction in energy usage for this column, the application of MPC reduced the energy consumption by 1.6×10^{10} BTU per year. As a result, fly ash would be reduced by 50 tons per year, the SO_2 emissions were reduced by 13 tons/year, and the NOx emissions were reduced by 3 tons/year. The fly ash estimates were based on 75% of the energy coming from natural gas and 25% from coal, with the assumption that the coal used is 15% ash with a 9000 BTU/lb heating value. The SO_2 estimate is based on the national average of 1.73 lb SO_2/MMBTU. The NO_x estimate is based on the national average of 0.35 lb NO_x/MMBTU. In addition, further assuming that the energy saved is in the form of methane, the energy savings of this project would reduce CO_2 emissions by over 1000 tons/year.

6.4.2 Union Carbide precooler and prefractionator project benefits

Improved distillation control does not always result in readily observable reduced energy consumption. Sometimes improved process control is used to increase processing throughput, to increase recovery of a valuable component, to reduce the variability of the products, or simply to stabilize the operating conditions of the process. For each of these cases the energy consumption may appear to remain relatively constant.

Clearly industry will use improved control in the most economically advantageous manner. In certain cases, when improved control is used to increase the throughput of the process at constant energy usage, the energy consumption per pound of product produced is reduced, which can be considered as energy savings. In other cases, improved control can increase the recovery of a valuable component using approximately the same amount of energy, also effectively decreasing the specific energy consumption.

In the case of the precooler and the prefractionator, significantly improved control was clearly realized for these columns, although no direct energy savings were observed. That is, the variability of the C_5 impurity in the overhead from the precooler was reduced by a factor of three. For the prefractionator, the propylene recovery was increased and the variability in the overhead product was reduced by a factor of two. These results represent very significant improvements in control performance. (It should be noted that, in light of operational and economic considerations required to be accommodated by the host facility, quantification of column energy savings was not a matter of priority during our control system application efforts in the field.)

It is axiomatic that, all else being equal, diminished disturbances to column inputs from whatever cause will directly result in lower variability products. While control system improvements apparently did not significantly affect energy consumption for these two specific columns, it is probable that, by reducing the variability of the products from the precooler and prefractionator, the energy consumption of the downstream columns would be improved due to the reduction in the variation in the feed composition to those downstream columns. This results in a more stable, predictable, and productive process, all of which translate into more or better product yield per unit resource expended.

7. CONCLUSIONS

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

1. For single-ended control, apply the (L,V) configuration

2. For dual ended control, the optimum configuration will generally require simulated control studies.

3. PMBC and ANN did not offer any advantages over PI and MPC

4. MPC outperformed PI where the control performance of one product was valued more than the other, which is usually the case industrially.

5. MPC outperformed PI on the industrial columns largely due to its ability to effectively handle a complex set of constraints.

8. NOMENCLATURE

AC   -   analyzer controller, i.e. a composition controller
ANN   -   artificial neural network
AT   -   analyzer sensor transmitter, i.e. an online analyzer
ATV   -   auto tune variation test for tuning a PI controller
B   -   bottoms flow rate
BLT   -   Biggest Log-Modulus Tuning, a controller tuning procedure
BPD   -   barrels per day
CV   -   controlled variable
CW   -   cooling water
D   -   distillate flow rate
DMC   -   Dynamic Matrix Control, a type of model predictive controller
DOF   -   degrees of freedom
EP   -   the 90% TBP end point temperature
F   -   column feed rate
HCN   -   heavy cycle naphtha
HCO   -   heavy cycle oil
IAE   -   absolute value of the integral of the error from setpoint
L   -   reflux flow rate
LC   -   level controller
LCO   -   light cycle oil
LT   -   level sensor transmitter
L_{AC} -   accumulator level
L_{RB} -   reboiler level
L/D   -   reflux ratio
MPC   -   model predictive control, a type of controller that uses a linear empirical model for control
MV   -   manipulated variable
P   -   column pressure
PI   -   proportional and integral controller
PMBC   -   process model based control, control based on a nonlinear process model
PT   -   pressure sensor transmitter
Q_{COND} -   condenser duty
S   -   steam flow
SRK   -   Soave-Redlich-Kwong equation of state
T   -   tray temperature
TBP   -   total boiling point, the temperature of a liquid after a certain portion has been evaporated
V   -   vapor boilup rate

VLE	-	vapor/liquid equilibrium
V/B	-	boilup ratio
x	-	impurity in bottoms product
y	-	impurity in overhead product
z	-	mole factor of light component in the feed
ZN	-	J.G. Ziegler and N.B. Nichols, pioneers in the field of automatic controller tuning

Greek Symbols

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

Subscripts

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

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Table 1

Design Specifications for Each Column Case Study

	C₃ Splitter	Xylene/Toluene	Depropanizer
Number of Trays	232	49	50
Feed Tray Location (From Bottom) 64		24	25
Feed Flow rate	13.41 kg/sec(106,500 lb/h)	29.23 kg/sec (232,007 lb/h)	17.01 kg/sec [135,026 lb/h]
Feed Comp. (mole %)			
Lighter than the Light Key			C_2 – 0.0193
Light Key	$C_3^=$ – 0.70	Tol – 0.67	C_3 - 0.3154
Heavy Key	C_3 - 0.30	Xyl – 0.33	iC_4 - 0.0844
Heavier than Heavy Key 1			nC_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_3 - 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 lb/h)	18.65 kg/sec (148,091 lb/h)	4.01 kg/sec (31,800 lb/h)
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 lb/h) 10.6 kg/sec (83, 916 lb/h)		13.01 kg/sec (103,226 lb/h)
Bottom Temperature	42.3°C (108.1 °F)	118.7°C (245.7°F)	139.0 °C (282.2°F)
Reboiler Vapor flow rate	131.18 kg/sec (1,041,165 lb/h)	41.16 (334,625 lb/h)	25.36 kg/sec (201,290 lb/h)
Reflux Ratio	12.6	1.21	4.75
Feed Quality	Saturated	Saturated	Saturated

Table 2

Design Specifications for the Main Fractionator

Feed

Flow Rate	50,000 BPD
API	40
Temperature	950°F
Pressure	35 Psia
Phase	Superheated Vapor
Components	Hydrogen, water, light hydrocarbon to asphalt (36 pseudo-components and 9 defined components)

Main Column

Number of Trays	40
Feed Tray Location (from top)	36
Diameter	18 ft
Overhead Temperature	110.6°F
Overhead Pressure	30 Psia
Overhead Vapor Gas Flow	6,120 BPD
Overhead Liquid Gas Flow	9,679 BPD
Overhead Liquid 90% TBP Endpoint	331°F
Bottom Slurry Flow	4,986 BPD
Slurry API	7.3
Bottom Stripping Steam Flow	10,812 lb/h
Bottom Temperature	690°F

Heavy Cycle Naphtha (HCN) Stripper

Number of Trays	5
Draw Tray Location on Main Column	11
Diameter	6 ft
Product 90% TBP Endpoint	400°F
Product Flow	10,663 BPD
Stripping steam flow	5,406 lb/h
Bottom Temperature	422°F

Light Cycle Oil (LCO) Stripper

Number of Trays	5
Draw Tray Location on Main Column	22
Diameter	5 ft
Product 90% TBP Endpoint	675°F
Product Flow	18,590 BPD
Bottom Temperature	415°F
Reboiler Heat Duty	10 MMBTU/h

Pumparound Flows

Top(stage 2-4)	755,909 lb/h
HCN(stage 11-9)	345,790 lb/h
LCO (stage 22-20)	128,654 lb/h
HCO (stage 25-23)	277,500 lb/h
Slurry (stage 40-31)	120,000 lb/h
Quench (stage 40-36)	574,403 lb/h

Pumparound return temperatures

Top(stage 2-4)	140°F
HCN(stage 11-9)	240°F
LCO (stage 22-20)	240°F
HCO (stage 25-23)	350°F
Slurry (stage 40-31)	420°F
Quench (stage 40-36)	420°F

Table 3

Model Assumptions for the C$_3$ Splitter, Xylene/Toluene Column, and Depropanizer

	C$_3$ Splitter	Xylene/Toluene	Depropanizer
Liquid Dynamic	Hydraulic time constant	Frances Weir Formula	Hydraulic Time Constant
Negligible Vapor/Holdup	yes	yes	yes
Value dynamic on all flows	no	yes	yes
Accumulation and Reboiler level control	PI	P only	PI
Analyzer delays on product composition	5 minutes	5 minutes	5 minutes
Eqimolal overflow	yes	no	no
Residence time in reboiler	5 minutes	5 minutes	5 minutes
Residence time in accumulator	5 minutes	3 minutes	10 minutes
Heat transfer dynamics modeled	no	yes	yes
Saturated liquid feed	yes	yes	yes
Subcooled reflux	no	no	yes
Pressure dynamics modeled	no	yes	no
Perfect mixing of liquid on trays	yes	yes	yes
Ideal VLE	no	yes	no

Table 4

Modeling Assumptions for the Main Fractionator

Liquid Dynamics	Hydraulic Time Constant
Negligible Vapor/Holdup	Yes
Value Dynamic on all flows	No
Accumulation and Reboiler level controls	PI
Analyzer Delays on product composition	Perfect inferred properties with first order delay (15 seconds)*
Equimolal overflow	No
Residence time in reboilers/bottom sumps**	5 min
Residence time in accumulator**	5 min
Heat transfer dynamics modeled	No
Saturated liquid feed	No
Subcooled reflux	No
Pressure dynamics modeled	No
Perfect mixing of liquid on trays	Yes
Ideal VLE	No

Notes:

*The qualities are endpoints or API gravity for main fractionator products. These are usually measured by an off-line laboratory. Inferential control is normally used based on pressure, temperature, and flow measurements. To simplify the simulation, the inferential model is assumed perfect, but the inferred properties are delayed by a first-order filter before they are used for control.

**There was a bottom sump and an accumulator with water decant in the main column, and a reboiler in each of the strippers. The same residence time was used for all these levels.

Table 5

Single Composition Control Results (IAE) for Feed Composition Upsets

	Manipulated Variable	C$_3$ Splitter	Xylene/Toluene Column	Depropanizer
Control of Overhead Product Composition	L	2.1	2.3	51.5
	D	3.8	2.8	148.6
	L/D	19.2	3.0	78.3
Control of Bottoms Product Composition	V	44.4	7.2	14.7
	B	198	8.4	72.1
	V/B	114	7.8	31.1

Table 6

Dual Composition Control Results (IAE) for Feed Composition Upsets

Configuration	C$_3$ Splitter				Toluene/Xylene Column				Depropanizer	
	F$_T$	Overhead	Bottoms		F$_T$	Overhead	Bottoms	F$_T$	Overhead	Bottoms
L,V	1.5	0.25 13.3	1.3		0.027	0.16	0.6		0.75	0.42
L,B	0.8	0.07 1.5	1.1		0.038	0.15	1.0		3.33	1.66
L,V/B	0.8	0.06 0.3	0.7		0.010	0.05	0.8		0.49	0.24
D,V	0.8	0.10 2.5	1.5		0.110	0.11	0.6		1.33	0.37
D,B	1.6	0.18 5.9	*		*	*	0.8		1.42	1.19
D,V/B	1.4	0.24 1.9	1.5		0.290	0.16	0.6		0.98	0.19
L/D,V	3.0	0.09 21.0	0.9		0.029	0.04	0.6		0.51	0.22
L/D,B	3.0	0.14 26.0	1.3		0.027	0.18	2.4		1.94	0.88
L/D,V/B	2.5	0.10 2.0	1.0		0.016	0.11	0.8		0.46	0.18

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

Table 7

Typical Industrial MV and CV Pairing for a Main Fractionator

MVs	CVs
Liquid distillate flow (D)	Overhead accumulator level (M_1)
Decant water flow (W)	Overhead water decanter level (M_w)
Top pumparound duty (Qp_1)	Overhead liquid endpoint (EP_1)
HCN side draw flow (S_1)	HCN stripper bottom level (M_{45})
HCN product flow (Fp_1)	HCN endpoint (EP_3)
LCO side draw flow (S_2)*	LCO draw tray level (M_{22})
LCO product flow (Fp_2)	LCO side stripper bottom level (M_{50})
LCO reflux to lower section (L_{22})*	LCO endpoint (EP_4)
Quench pumparound duty (Qp_6)	Slurry API (API_2)
Slurry product flow (L_{40})	Bottom level (M_{40})
	Bottom temperature (T_{40})**

Other DOFs	
Vapor distillate flow (V)	Fixed for maximum compressor capacity
Top reflux flow (L1)	On flow control
Qp_2-Qp_5	Fixed and set by higher level optimizer
Stripping steam flows	Fixed

Notes:

* The LCO draw tray is a total draw tray (chimney tray). All the liquid is pumped outside the column and split three ways: LCO pumparound, LCO reflux to lower section (L_{22}), and LCO side draw (S_2).

** Bottom temperature is controlled by using an overrider with Qp_6 as the MV.

TABLE 8

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

Configuration	C_3 Splitter (L,B)		Xylene/Toluene (L,V/B)		Depropanizer (L/D,V/B)	
	x	y	x	y	x	y
PI	1.49	0.067	0.052	0.040	10.6	27.4
MPC	5.73	0.013	0.10	0.060	23.5	59.4
PMBC	1.61	0.021	0.739	0.051	47.7	46.9

Table 9

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

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

Table 10

**Comparison of Control Performance (IAE) Between MPC and PI for
the Vacuum Column with Priority Given to the Bottoms Product**

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

Table 11

Comparison of IAEs Among MPC, PI, and PI with a Simple Decoupler (DC) for the Main Fractionator

	EP_1(F-h)	EP_3(F-h)	EP_4(F-h)	API_2(API-h)
Heavier Feed				
PI	12.6402	2.5639	2.6983	0.029682
PI-DC	9.2862	1.9341	1.2142	0.126406
MPC	2.98	1.2213	1.7706	0.136884
Lighter Feed				
PI	6.7658	1.5753	1.8367	0.056049
PI-DC	5.7936	1.0356	1.7478	0.043366
MPC	1.4057	0.6866	0.9922	0.050474