

# MULTIVARIABLE CONTROL OPTIMIZES CRUDE UNIT NAPHTHA SYSTEM

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## ABSTRACT

A constrained multivariable predictive controller using Dynamic Matrix Control (DMC) technology has been successfully applied to the overhead naphtha system of a crude oil distillation unit at Marathon Oil Company's Refinery in Garyville, Louisiana. A single controller was implemented for three columns in series: a naphtha splitter, a stabilizer, and a de-isopentanizer. The DMC controller has 19 dependent controlled variables, 13 independent manipulated variables, and 6 independent feedforward variables. The multivariable controller uses a linear program (LP) to optimize the operation of the columns and to maximize production of the most valuable product streams within the prevailing process constraints. These constraints include limits on the boiling point properties and concentrations of impurities in the product streams as well as physical equipment constraints. The product quality constraints include variables that are measured by on-line analyzers and those that are inferred using other available measurements.

The controller uses a novel technique for the control of an intermediate surge drum level, which provides for the best use of the available surge volume. The controller drives the level toward its high or low limit depending on whether the overall throughput constraint is upstream or downstream of the drum. This allows the distillation columns on both sides to be operated at maximum capacity for a higher percentage of the time when the throughput constraints change from column to column. The net effect is a higher overall throughput and smoother operation than by using the traditional level control technique.

Commissioning of the controller has resulted in increased and less variable throughput, improved control of product specifications and a reduced load on the console operator. Operator acceptance of the control system has been very good with high controller utilization being consistently achieved over a nine-month period to date.

## INTRODUCTION

This paper describes the implementation of a DMC multivariable control application to a typical refinery plant. Several practical issues relating to multivariable control implementation in the field are highlighted. Often these practical issues are dismissed in the literature as 'implementation details' and somewhat shunned. There is much truth in the adage, though, that 'the devil is in the details' and the real-world success of the best algorithms depends on addressing these issues. Multivariable control algorithms that are widely available today are capable of solving the vast majority of industrial control problems. Their widespread use and resultant benefit to industry is limited by the availability of competent engineers who can negotiate the minefield of implementation details and complete a successful project.

## PROCESS OVERVIEW

Figure 1 shows a simplified process flow diagram of the crude unit overhead naphtha system included in the scope of the control problem. The liquid product from an upstream atmospheric crude distillation column is combined with a light naphtha stream and fed to the splitter column. Here light straight-run (LSR) gasoline, consisting of C<sub>5</sub>- with a true boiling point end point of 185-190°F, is separated from the higher boiling naphtha cuts. The LSR product from the splitter overhead accumulator is further processed in a stabilizer column to remove butane and lighter hydrocarbons (C<sub>4</sub>-). These light hydrocarbons are passed to a saturated gas plant for further separation. The stabilizer bottoms are then fed to a de-isopentanizer (DIP) column where iC<sub>5</sub> is separated and sent via a sweetening unit to gasoline blending while the LSR from the bottom of the DIP is sent to an isomerization unit.

Several features of the process are notable for their impact on the design and implementation of the multivariable control system. These issues are highlighted here:

1. The splitter has no regulatory pressure control system because of physical problems with the condenser bypass system piping. The pressure control valves have been isolated by closing their associated block valves. The console operators must allow the pressure to float, monitoring it continually to ensure the value does not exceed normal limits. The continually varying pressure makes it difficult to maintain a constant composition profile in the column since temperature measurements alone are not meaningful. In summer, when ambient temperatures are high, the splitter feedrate often has to be reduced because of high column pressure, indicating that the condenser capacity limit has been reached.
2. The splitter reboiler is subject to gradual fouling and occasionally must be cleaned on-line. When the reboiler is most fouled, the console operators find it necessary to place the flow controller in manual and close the control valve completely to get sufficient duty from the reboiler. Note that since the control valve is an exchanger bypass, the valve must be closed to increase the heat supplied to the reboiler. Such saturation of a regulatory control valve can represent a non-linearity in the process behavior, and must be dealt with carefully.
3. The splitter overhead accumulator has a significant surge volume, indicating that it may represent an opportunity to reduce process variation by utilizing this volume effectively.
4. The de-isopentanizer column (DIP) has a regulatory bottoms level controller cascaded to the reboiler steam flow. This configuration was set up as part of the project to implement multivariable control on the process. The previous level controller, cascaded to the bottoms product flow, was ineffective in maintaining the level setpoint in response to disturbances. The small product flow relative to the column's large internal traffic required that the controller make very large moves in bottoms flow, upsetting the downstream isomerization unit. In practice, the console operators would leave the old level controller out of service, and try to keep the level within the instrument range by making manual adjustments. The level-to-reboiler cascade has been very effective, since by manipulating the column's internal traffic relatively smaller moves are made in this more powerful handle. This configuration also allows the isomerization unit's feedrate to be set constant without sacrificing regulatory control of the DIP level.
5. Only the DIP overhead product has a process analyzer installed to measure the controlled product composition. The measurements of interest here are the percent C<sub>4</sub> and nC<sub>5</sub> in the iC<sub>5</sub> stream. Other compositions of interest are only measured by laboratory test results, which are typically available twice per day. The other compositions of interest are splitter bottoms percent C<sub>6</sub>, splitter overhead 98% boiling point, percent C<sub>5</sub> in stabilizer overhead and percent iC<sub>5</sub> in DIP bottoms.

## DMC CONTROLLER DESIGN

The control system design requires selection of the manipulated and controlled variables for the multivariable control system. The number of separate multivariable controllers and the division of variables between them must also be decided. The selection of manipulated variables determines the scope of the control system and the regulatory control structure that will be used. The greater the number of manipulated variables in a single multivariable controller, the larger its scope or size. The size of each individual controller should be made as small as possible without sacrificing any potential economic benefit. Large controller size is not desirable for its own sake, but is frequently required to handle all constraints and dynamic interactions rigorously and therefore achieve maximum economic benefit. The

ultimate objective of multivariable control is to optimize process operations, and to achieve this goal all constraints to the plant's economic objectives must be considered as part of a single problem. This consideration defines the scope of a well-designed multivariable control system. If the capability of the multivariable control algorithm constrains the size of the controller then economic benefits have been sacrificed. In the case of the naphtha processing system under consideration the DIP is frequently the constraint to maximizing throughput of light naphtha. The controller scope must therefore span the process between the feed flow manipulated variable and the DIP constraints if it is to achieve maximum possible benefit.

Following the above principles, the variables included in the single controller are 13 manipulated variables, 6 disturbance variables and 19 controlled variables. The variables are listed in tables 1a through 1c. The use of the LP in the DMC controller means that each of the manipulated and controlled variables is given a high and low limit by the console operator. The LP then chooses the most economic operating point, passing control targets to the move calculation engine. Each controlled variable may therefore be assigned a setpoint or used only as a constraint without the need for two classes of variables.

The five process issues raised in the discussion above were each considered carefully for their impact on the multivariable control design. The individual issues were addressed as follows:

1. The lack of regulatory pressure control on the splitter was clearly not desirable, but it was also clear that the console operators could successfully run the column in this mode. The multivariable control system therefore could operate in this configuration too. The major difficulty was that the varying pressure made composition difficult to control by maintaining column temperatures. The DMC controller includes the pressure as a controlled variable with limits placed so as to maintain an acceptable range of operation. The DMC controller functions as the column's pressure controller, using all the available handles for multivariable pressure control. The upper pressure limit is determined by the column pressure relief valve setting, this ensures that maximum feedrate is attained when the condenser capacity is limiting. The control of composition is facilitated by the use of inferred properties using pressure compensated temperatures as described below.

2. Since the reboiler control valve position is often a limiting variable for the splitter performance, this valve is used directly as a manipulated variable. If the flow setpoint were manipulated then the saturation of the valve would have to be avoided as an unacceptable non-linearity. This would result in the necessity to leave a small amount of opening on the valve at all times, giving up some capacity. The direct manipulation of the valve allows it to be closed entirely without invalidating the process model. This principle is encountered in DMC controller design for many processes, and important regulatory loops are often broken to obtain the maximum possible performance from the plant.

3. The surge volume in the splitter overhead accumulator is exploited by the multivariable controller in a novel way. It is designed to manipulate the outflow from the vessel and control the level, thus breaking the regulatory level controller. This allows the application of an interesting feature available in the DMC algorithm used here. The feature is known as a 'programmed imbalance ramp', referring to the fact that the controller can deliberately maintain an imbalance in the level when it is desirable to do so. This goes beyond traditional averaging level control by consciously building or depleting the level depending on the currently active process constraints. The aim of the imbalance is to maximize economic benefit to the process. If the constraint to plant throughput lies in the DIP column, traditional control techniques would reduce the splitter feed to balance the accumulator level. The programmed imbalance ramp recognizes the opportunity to maintain the splitter at its maximum capability and store the excess product in the drum to be used when the splitter becomes limiting. In this splitter-limited case, the controller will consciously deplete the level by utilizing all the available DIP column capacity. These imbalances are maintained until the level reaches its high or low limit at which point the level is balanced and controlled at the limiting value. In this plant, the primary constraint to throughput alternates between the splitter and the DIP. When the accumulator level is between its upper and lower limits each column will be pushed to its capacity limit. The application of the programmed imbalance feature therefore increases the average utilization of each column and therefore the total plant throughput.

4. Establishing functional level control on the DIP greatly eased the implementation of DMC multivariable control on the column. This level control configuration results in a mass balance control scheme on the column, the compositions being adjusted primarily by manipulating the bottoms flow or column feed flow. Care must also be taken to ensure that the DMC controller respects the column's mass balance; not, for example drawing more from the bottom

of the column than is being fed to it. This objective is achieved by enforcing both high and low constraints on the overhead and bottoms impurities.

5. Where critical stream compositions are not measured, inferred properties have been developed for use as controlled variables. For this controller, all of the inferred properties are based on statistical regression of empirical data gathered during a controlled plant test. This data, gathered at the same time as the dynamic response testing described below, involved moving the compositions by a significant amount and then taking laboratory samples with the process as close to steady state as practical. The process conditions were recorded at the time of each lab sample and various combinations of process variables were regressed against the lab data. Good fits were achieved and the best relationships in each case were implemented as controlled variables for the DMC controller. Most of these inferred properties use pressure compensated temperatures as their independent variables, with some adding internal reflux flows or other predictors. When used on-line, these inferred properties have bias updates implemented using feedback from lab results. This corrects for any prediction error in the inferred property calculations and is essential to their success. The calculations have performed well in practice, resulting in significant reduction in the variability of the product qualities from the plant.

## RESPONSE TESTING AND DYNAMIC MODELLING

The most critical element in ensuring the success of a predictive multivariable controller is obtaining a sufficiently accurate dynamic model of the process. The key to obtaining a good model is high quality data obtained from a well-executed plant response test. There are many pitfalls for the unwary in the gathering of plant data to generate empirical models. Correlation between independent variables, feedback introduced by control systems or by operator intervention and insufficient perturbation of the system are only a few of the possible causes of erroneous models. It is critical to plan and execute a carefully supervised test during which each of the independent variables is perturbed in sequence. Great care must be taken to ensure independence of the perturbation sequences and sufficient signal to noise ratio in the dependent variables to be modeled. The best instincts of a good console operator are to resist change in important controlled variables. Active supervision of the test is necessary to avoid these instincts creating feedback or correlation in the data. Other duties of the supervising engineer include ensuring that no changes in regulatory configuration or control system saturations occur during the testing, and watching the process for disturbances so as to identify suitable feedforward variables.

The plant response test of this plant was carried out round the clock for approximately two weeks. The engineers adopted twelve-hour shifts alongside the plant operators for the duration of the testing. Each manipulated variable was perturbed in sequence, groups of four to six step changes being made in each. Each step move was held for a duration between one to four thirds of the settling time of the process. This sequence was then repeated several times until a total of twelve to twenty moves had been made in each manipulated variable. Separation of each manipulated variable's moves into groups reduces the likelihood of an unmeasured disturbance distorting one variable's behavior. It also has the benefit of allowing preliminary regression cases to be done in the plant after one rotation of moves has been made. These preliminary analyses allow the quality of the data to be evaluated so that, before finishing the test, the data is known to be good enough to develop an acceptable model.

The plant data were analyzed using a multivariable identification algorithm, which uses the inverse of the dynamic matrix control equations (Cutler and Johnston, 1985). An advantage of this technique is that the model form describing the dynamics of the system is not assumed. Instead, the numerical representation of the dynamic response curves allows the model to be of any order. Also, because all interactions between variables are taken into account more than one independent variable at a time can be moved during the plant testing without invalidating the analysis. Further, the system is not required to be at steady state to start the analysis or to come to steady state during the analysis. These attributes permit the operator to make changes to keep the process on specification without disrupting the test,

Of the 361 (19 X 19) possible independent/dependent variable response curves, the controller contains 118 responses as shown in Figure 2. The missing pairs occur where the response is either zero or insignificant.

## DMC CONTROLLER COMMISSIONING

The controller was commissioned at the Marathon refinery over a two-week period during which control engineers again provided continuous coverage. Following initial off-line tuning and simulation the controller was turned on, intensive monitoring and further tuning adjustments ensured the proper functioning of the controller. Operator training is a major objective of the commissioning period; extensive time was spent with each of the console operators explaining the controller's functions and behavior.

Two types of parameters are used to tune the controller; these are the Equal Concern Errors (ECE) for dependent variables, and the Move Suppression Factors (MSF) for manipulated variables. The ECE provides a mechanism for weighing the dependent variables relative to one another. Various ECEs determine the negative slack penalties in the Linear Program (LP) portion of the DMC controller and the weighting for each variable in the overall DMC calculations. The MSF tuning parameters allow trade-offs between movement of the manipulated variables and error in the dependent variables. A high MSF will result in sluggish movement in the manipulated variable and more error in the dependent variable. The ECE and MSF are tuned based on knowledge of the process and the relative importance of the control objectives. Trial and error adjustment during commissioning finalizes the tuning process.

The overall operating objective of the controller is to maximize the feed rate and the recovery of the most profitable products subject to column constraints and product specifications. The LP objective function is configured such that the controller will always strive to increase light naphtha feed up to the high limits on column constraints.

The splitter column has five controlled variables associated with it including two inferred composition values: the LSR 98% point and the splitter bottoms %C<sub>6</sub> and both are constrained between high and low limits. Isomerization unit charge is the most valuable stream so the controller tends to maximize the overhead product rate thus maximizing LSR 98% point. When the DIP column is limited or the maximum isomerization charge is already being produced, then the LSR 98% point will be reduced in order to run more light naphtha for naphtha production. Thus an incentive is always present to maximize light naphtha feed. The splitter DP is a limit to the light naphtha feed rate to prevent column flooding. The overhead pressure may also limit light naphtha feed if the splitter becomes condenser capacity limited.

The splitter overhead accumulator level is operated as a surge volume as described above. In the situation where the splitter is the limit to throughput, then the controller will drive the accumulator level down to the low limit in order to keep the isomerization charge maximized for as long as possible. Only when the level is at its low limit will the isomerization charge be reduced. If the constraint to throughput lies downstream of the splitter then the controller will continue to run as much light naphtha as possible and drive the level in the accumulator up to the high limit, thus accumulating material for periods when the splitter capacity may be lower. This intelligent use of the surge volume results in more average capacity than would a more traditional level control strategy.

The stabilizer has five controlled variables including one inferred composition for the stabilizer overhead %C<sub>5</sub>. The stabilizer DP is maximized by pushing reboiler duty up to the column flooding limit. The two valve position controlled variables in the pressure control system ensure that the controller maintains the stabilizer pressure on control at all times and allows a pressure minimization strategy to be followed.

The DIP column has eight controlled variables including two predicted analyzer values for %C<sub>4</sub> and %nC<sub>5</sub> and one inferred composition for the DIP bottoms %iC<sub>5</sub>. When the constraint to throughput is not in the DIP then the overhead normal composition is reduced to the minimum limit before isomerization charge is reduced. If the DIP column is the overall constraint, then the overhead nC<sub>5</sub> content will be maximized to the high limit before overall charge is cut. At times the overhead product hydraulic capacity may become a constraint before the composition does, in which case overall throughput is reduced, resulting in a lower light naphtha charge.

Overall throughput has been increased considerably since commissioning of the controller, bringing significant economic benefit. The controller has been well accepted by the console operators and remained in service continuously during periods of normal operation since commissioning. It has made a major reduction in the attention required to maintain smooth operations of the plant.

## CONCLUSIONS

A single DMC controller has increased throughput and stabilized operation of a crude distillation unit's overhead naphtha system. The single controller was specified because of its ability to rigorously handle all of the process constraints and thus achieve the optimum economic operation for the process. A novel level control technique used on an accumulator vessel provided further benefit in additional throughput and smoother operation. Inferred properties developed for unmeasured stream compositions were successful in predicting lab results and facilitated control of the compositions.

## REFERENCES

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**Table 1a. Manipulated Variables**

	Description	Typical Units
1.	Light naphtha feed flow	MBPD
2.	Splitter reboiler valve	%
3.	Splitter overhead temperature	Deg. F
4.	Stabilizer feed flow	MBPD
5.	Stabilizer reboiler flow	MBPD
6.	Stabilizer tray 5 temperature	Deg. F
7.	Stabilizer pressure	PSIG
8.	Stabilizer pressure valve controller	%
9.	Stabilizer condenser DP	PSI
10.	DIP reflux	MBPD
11.	DIP column pressure	PSIG
12.	DIP bottoms flow to isomerization	MBPD
13.	DIP hot vapor bypass valve	%

**Table 1b. Disturbance Variables**

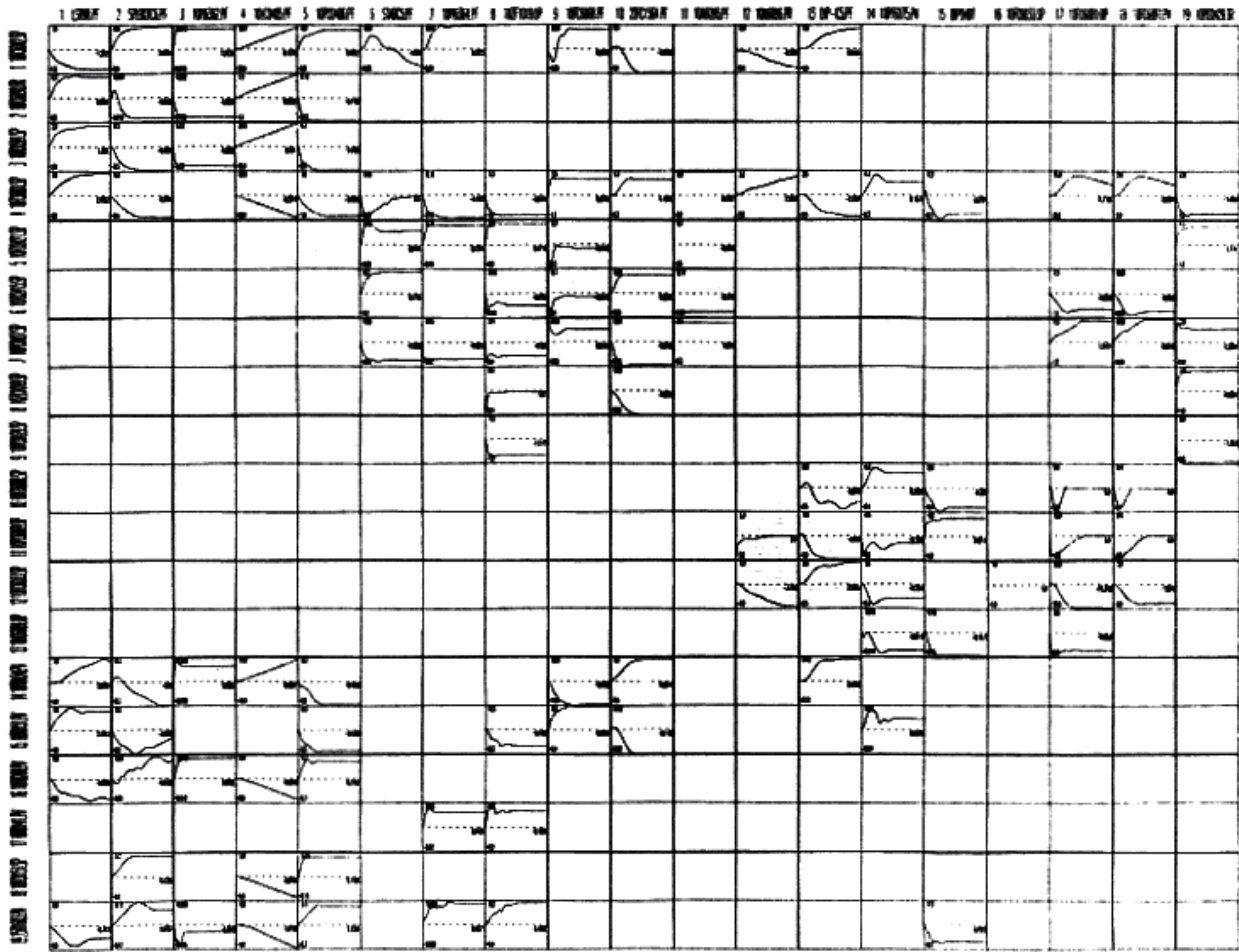
	Description	Typical Units
1.	Crude overhead product flowrate	MBPD
2.	Crude overhead receiver temperature	Deg. F
3.	Kero draw rate	MBPD
4.	Ambient temperature	Deg. F

**Table 1c. Controlled Variables**

	Description	Typical Units
1.	LSR 98% temperature	Deg. F
2.	Splitter bottom %C6	%
3.	Splitter DP	psi
4.	Splitter overhead accumulator level	%
5.	Splitter overhead pressure	PSI
6.	Stabilizer overhead %C5	%
7.	Stabilizer DP	PSI
8.	Stabilizer pressure valve	%
9.	Stabilizer bottoms flow	MBPD
10.	Sat gas absorber liquid flow	MBPD
11.	Stabilizer condenser bypass valve	%
12.	DIP overhead %C4	%
13.	DIP overhead %n-C5	%
14.	DIP bottoms %i-C5	%
15.	DIP column DP	PSI
16.	DIP overhead reflux DT	Deg. F
17.	DIP target product flow to isomerization	MBPD
18.	DIP overhead flow valve	%
19.	DIP overhead flow	MBPD

### DMC Controller Model

DMIPLT Version 3.38      Files DMC10GAS.MDL...      Period: 240.0 (mins)



**Figure 1. Crude Unit Overhead Naphtha System – Simplified Flowsheet**