Dynamic Simulation of an Atmospheric Crude Tower

Damien Hocking[1], Geoff Roosen[2] and Dain Stone[2]

[1]Hyprotech Ltd.
Calgary, Alberta, Canada
www.hyprotech.com
[2]Plant Solutions Pty. Ltd.
Melbourne, Victoria, Australia

Steady-state simulation of chemical processes has been a standard process engineering activity for many years. Dynamic simulation has had a very restricted application. There are two main reasons for this. Firstly, the computing power required for a useful (*i.e.* fast) dynamic simulation is approximately an order of magnitude greater than the equivalent steady-state simulation. Secondly, general-purpose dynamic simulation software has not been available until recently and the development of a dynamic simulation model required considerable programming effort from engineers. In the last few years computing power and software design have improved to allow realistic dynamic simulation of chemical processes.

Dynamic simulation offers many benefits over steady-state simulation. As economic constraints on process engineering have tightened, the assumption of a steady-state with a margin for error has become uneconomical both from a design and operations perspective. Many plants do not ever reach a true long-term steady-state. This places a significant emphasis on control system design.

Oil refining is an example of a process frequently subject to operating changes. The feedstock and product demands change regularly. The first major unit of a refinery is the atmospheric crude tower and the operation of the tower has a dramatic effect on the rest of the plant. A model of a generic 40000 BPD atmospheric crude tower has been developed in HYSYS.Plant to demonstrate the benefits of dynamic simulation applied to operational factors and implementation of advanced control. The model includes the preheat train and control system but remains straightforward enough to be readily understood by process engineers and operators.

INTRODUCTION

An atmospheric crude tower is the first major process unit of a refinery. Its purpose is to separate the crude feed into main fractions for further processing. The standard tower configuration is based on a main column with three or four sidestripper columns. In a four-sidestripper configuration, the products are light Naphtha from the condenser, and heavy Naphtha, Kerosene, Diesel and an Oil product from the four sidestrippers. A three-sidestripper configuration is similar except that all the Naphtha is drawn from the condenser.

Steady-state simulation of a crude tower will determine the operating conditions required for different product demands. There are many critical quality variables for refining; for an atmospheric tower the main ones are the cutpoints for each of the drawn products. A cutpoint defines the temperature at which a certain percentage of a particular product has boiled off. Usually the tower products are controlled for 5% and 95% cutpoints. Naphtha has a 5% cutpoint starting as low as 50 F and a 95% cutpoint of 356 F. This means that at 50 F, 5% by volume of the Naphtha will have boiled off and at 356 F, 95% of the Naphtha will have boiled off, at atmospheric pressure. The Kerosene cutpoints are about 360 F and 464 F, Diesel is about 470 F and 644 F and the Oil cut has a 5% cutpoint of about 650 F and a 95% cutpoint from 700 F upwards. The 95% cutpoint of a particular product is close to the 5% cutpoint of the next heaviest product because as one product finishes boiling, the next heaviest starts to boil. A D86 boiling curve analysis of a particular crude oil will provide a nominal product split based on the cutpoints. The demand and value of each product stream and the cost of utilities will determine the final optimum split for the refinery. For example, if Kerosene has a very high value, an optimisation could be performed to maximise the Kerosene return from the crude without unreasonably affecting the rates and qualities of the other products.

The optimum product split during a campaign will change if the values of the product streams change. Steady-state simulation will provide operating conditions for the plant but it does not provide any information about how the tower will behave at those conditions or how it can be transitioned from one regime to another. It is obviously beneficial to move to the new operating point as quickly as possible to minimise the production of lower-value products. Thorough investigation of stability and transitions requires a dynamic simulation.

Steady-state optimisation is a relatively mature technology and this paper focuses on the dynamic simulation and control aspects of a generic small-scale crude tower. The PID inventory control system, a Model Predictive Controller (MPC) implementation and dynamic simulation results are presented. The HYSYS.Plant MPC is based on the public-domain Dynamic Matrix Control (DMC) algorithm. The model with the MPC active runs at 25 – 30 times real time on a PIII 700 MHz PC. Numerical values are presented in imperial refining units.

SIMULATION MODEL

The model simulates a three-sidestripper tower processing 40000 BPD of crude oil. The objective of the model is to realistically simulate an atmospheric crude unit with rigorous thermodynamics while running considerably faster than real time. To meet this objective only the major process equipment is included and equilibrium is assumed for the distillation units. The main tower PID/PFD is in Figure 1 below.

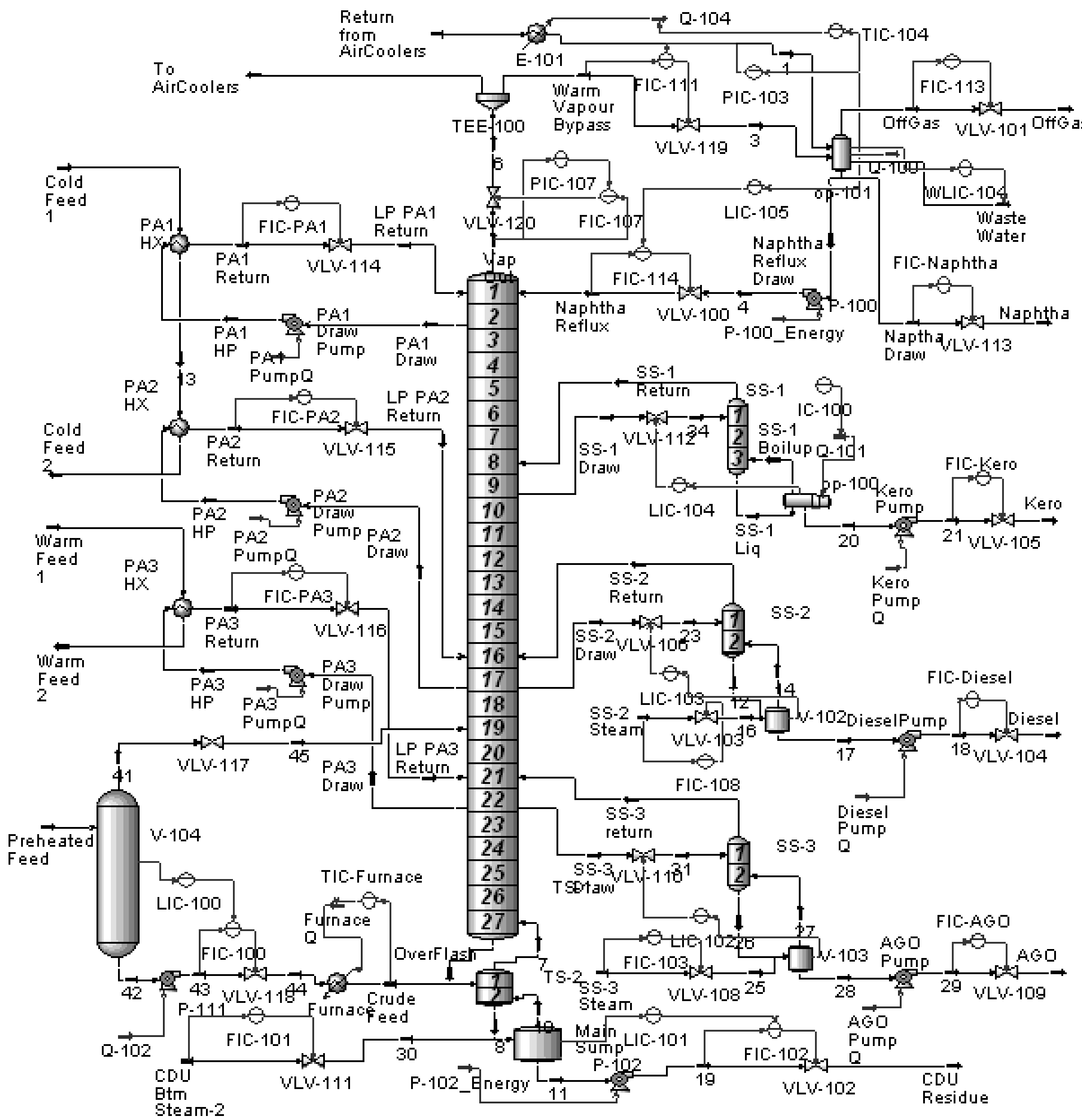

Figure 1: Tower PFD/PID structure

The crude feed passes through the preheat train and pumparound exchangers and into the preflash separator. The vapour from the preflash separator is fed to the main column and the liquid is fed to the furnace. The main column contains 29 equilibrium trays. The furnace product is fed to tray 28. The column sump is steam stripped. The pumparound cooling circuits draw from trays 2, 17 and 22 and return to trays 1, 16 and 21 respectively. The sidestrippers draw from trays 9, 17 and 22. The Diesel and AGO sidestrippers are steam stripped and the Kerosene sidestripper is reboiled. The Diesel and AGO sidestrippers have two equilibrium trays and a sump. The Kerosene sidestripper has three equilibrium trays and a sump. The main tray section has nominal top conditions of 310 F and 38 psia.

The overhead vapour is split and the majority of the vapour is partially condensed through three air coolers and then subcooled and fed to the condenser vessel. A small fraction of the tower vapour bypasses the overhead cooling system and is fed directly to the condenser vessel to maintain a warm vapour blanket for pressure control. Waste water, product Naphtha and reflux Naphtha are drawn from the condenser vessel. The condenser conditions are nominally 100 F and 25 psia.

The thermodynamics are modelled with the Peng-Robinson equation of state. The crude oil is characterised as a hydrocarbon mixture with water, C_1 – C_5 alkanes and 35 hypothetical oil fractions covering a boiling point range from 108 to 1336 F. The unit operation models are based on first-principles mass and energy balances and pressure-flow relationships. The tray models incorporate hydraulics and weir calculations. The dynamic model was initialised from the converged steady-state model.

PID Control System

The basic PID loops control inventory and energy. The flow controllers are mass flow-based. The inventory control is based on the assumption that the column is driven by product demand. Therefore the Naphtha, Kero, Diesel and AGO product streams are mass flow controlled directly. Level control on the condenser is through the Naphtha reflux. The sidestripper sump level controllers act directly on the sidestripper draw valves. The draw valves are driven by the low hydraulic pressure drop from the pressure difference between the draw and return trays. Depending on the column this pressure drop is typically less than 1 psi. A flow control loop would not respond well with such a low pressure drop, so the valves are controlled directly and the level control is allowed to be sluggish. This also constrains the setpoint changes on the product flow controllers. The vapour flow from the top stage controls the column top-stage pressure. A warm-vapour bypass controls the condenser pressure. Condenser vessel temperature is controlled with the final subcooler.

The steam rates are controlled at 8 lb steam/bbl of stripped residue, and 3.5 lb steam/bbl of Diesel and AGO product. The Kerosene stripper reboiler is controlled at a boilup ratio of 0.5.

MODEL PREDICTIVE CONTROL

Overview

MPC will not be described in great detail; complete summaries with mathematical derivations are available in any modern control textbook. MPC uses a simplified process model to perform two tasks:

1) Predict future plant response.
2) Determine appropriate control action to drive the predicted response as close as possible to the controller setpoints.

The Dynamic Matrix Control (DMC) algorithm was pioneered industrially by Shell in the late 1970s. The fundamental principle of DMC is to use a matrix of process models to approximate the response of specific plant process variables to changes in specific manipulated variables. The matrix of process models is constructed by step-testing the real process. A manipulated variable is step-changed by say 5% of its control span, and the response of the appropriate process variables is recorded. A first-order-plus-deadtime model is then fitted for each process variable response. The step-tests are conducted for each manipulated variable. A DMC controller employs a multivariable optimisation algorithm to determine the new manipulated variable values at each controller cycle, based on minimising the deviations of the process variables from their setpoints over a control time horizon. The great advantage of DMC is that beneficial and adverse interactions between variables are managed by the multivariable optimiser and the controller output is optimised for the best overall response. A group of SISO controllers cannot do this. The main disadvantage of DMC is that it requires a significant amount of time to step-test a real plant, at considerable cost and risk of producing low-quality products. The HYSYS.Plant MPC employs the standard public-domain DMC algorithm with a quadratic objective function.

An atmospheric crude tower is an excellent candidate for MPC. A modern atmospheric tower is strongly heat-integrated through the preheat train and there are strong interactions between the product draws. Consider the effect of changing the mass draw rate of a product on the cutpoints of the product streams. A product flow increase will remove more material in that product's boiling range, leaving less material in that range for the heavier products. For example, if the Naphtha draw rate is increased, more material that boils around 356F will be drawn into the Naphtha product. This will increase the 95% cutpoint of the Naphtha. The Naphtha also increases in density, which changes the volume flowrate of the Naphtha. The volume flow is another important control variable. Removing more 356 F-range hydrocarbons in the Naphtha means that there is less going into the light 5% end of the Kerosene, so the Kerosene cutpoints increase and the volume flow changes. There will be a similar but milder effect on the Diesel and AGO products as well.

DMC Implementation

There are a large number of process variables that could be controlled by an MPC on an atmospheric tower. There are two cutpoints and a volumetric draw rate for each product. There are other quality constraints on specific products, such as the pour point for Kerosene or the cetane index for Diesel. A critical economic variable is the tower overflash. Usually measured as a percentage of the total volumetric feed rate, the overflash determines how much of the vapourised feed from the furnace is condensed and returns to the bottom of the tower. The more overflash, the more energy is wasted. In an ideal crude tower, the overflash would be zero because just enough crude would be vapourised to be drawn off as products. Operating goals for overflash have steadily decreased over the years as economic constraints have tightened. Many years ago 4-5% overflash was considered acceptable, whereas many refiners now have 1-2% as a goal.

With the objective of a more straightforward dynamic example, 8 variables are selected for the DMC controller: the 95% cutpoints for Naphtha, Kerosene and Diesel, the volumetric flowrates of Naphtha, Kerosene, Diesel and Atmospheric Gas Oil (AGO), and the overflash flow. The AGO cutpoint is not controlled. The oil cut from a crude tower is generally a "sloppy cut" and not as strictly controlled as the other products. The 5% cutpoints are not controlled. The objective of the DMC controller is to satisfy product demand while maintaning quality. There are only 5 manipulated variables: the mass flowrate setpoints of the Naphtha, Kerosene, Diesel and AGO controllers and the furnace temperature controller setpoint. Inclusion of the stripping steam flowrates would permit control of the 5% cutpoints as well but they are omitted for simplicity. There are widely varying time constants in this system, the mass draw rate of Kerosene will have an instantaneous effect on the Kerosene volumetric flowrate, but a slower effect on the Naphtha, Diesel and AGO rates. The slower density change of the Kerosene will ultimately effect the Kerosene volumetric flowrate as well. This is a difficult optimisation problem and any solution will be a compromise between the product streams. This reflects what happens in the real refining process, a D86 analysis indicates suitable product flowrates based on ideal cutpoints, but economic factors influence the final demands and the split must be changed within acceptable quality limits.

The overflash flow is a difficult value to measure directly; it is the weir liquid flow of the tray above the feed tray. It may be inferred in a real plant by measuring the tray pressure drop. For the purposes of this example we have artificially created a directly measurable weir flow by separating the tray section above the feed tray and inserting a stream. This is visible in Figure 1; the stream is named Overflash and drops from Tray 27 of the main column. On-line analysis of cutpoints is an established technology in use in many refineries.

Step Tests

The Naphtha mass draw rate step tests are presented below in Figures 2a and 2b.

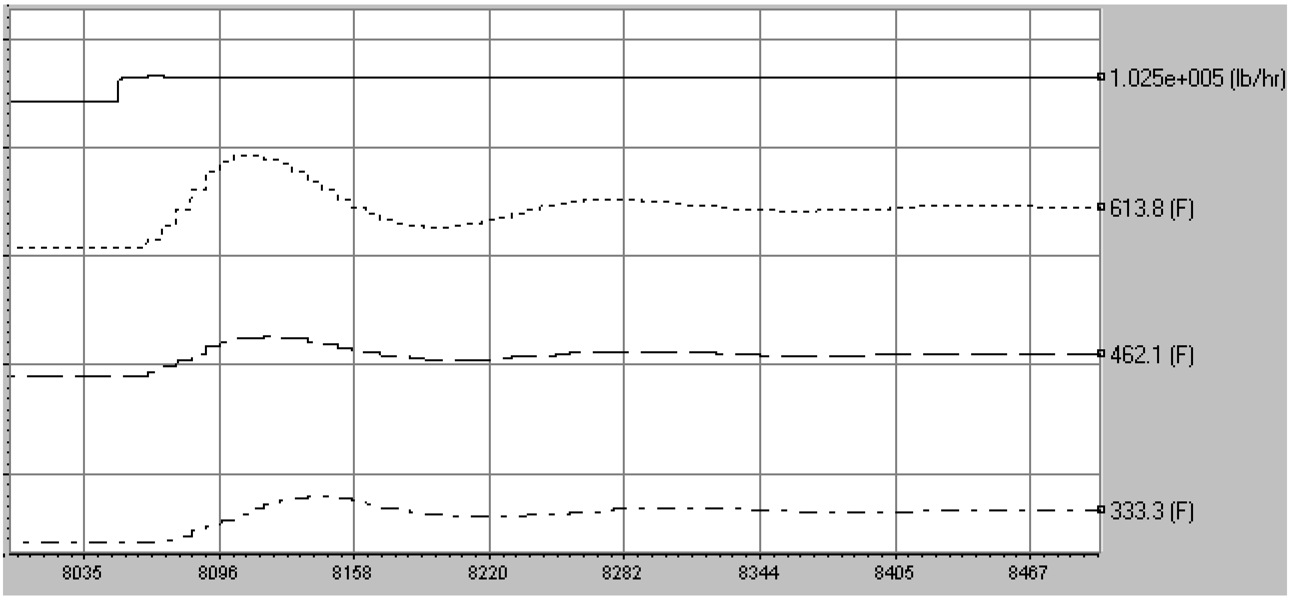

Figure 2a: Naphtha Step Test Cutpoint Response

Figure 3a above presents the 95% cutpoint responses of the Naphtha, Kerosene and Diesel. Each cutpoint is displayed on a separate 100 F scale and the responses are relative to each other. The total time span on the plot is 500 minutes and the step was initiated at 8050 minutes. The solid line at the top is the Naphtha mass step, which was 2.5% of the mass flow span. The dotted line is the Diesel 95% cutpoint, the dashed line is the 95% Kerosene cutpoint and the dot-dashed line is the Naphtha 95% cutpoint. The Kerosene and Diesel cutpoints actually respond faster than the Naphtha cutpoint. The deadtimes are approximately 12 minutes for the heavier cuts and about 30 minutes for the Naphtha. The effect of increased cutpoints in all the products from the one flow change is obvious. The shapes of the responses do not accurately fit a typical first-order-plus-deadtime model. This is a significant issue for implementing a DMC controller and is discussed in more detail with reference to Figure 2b below.

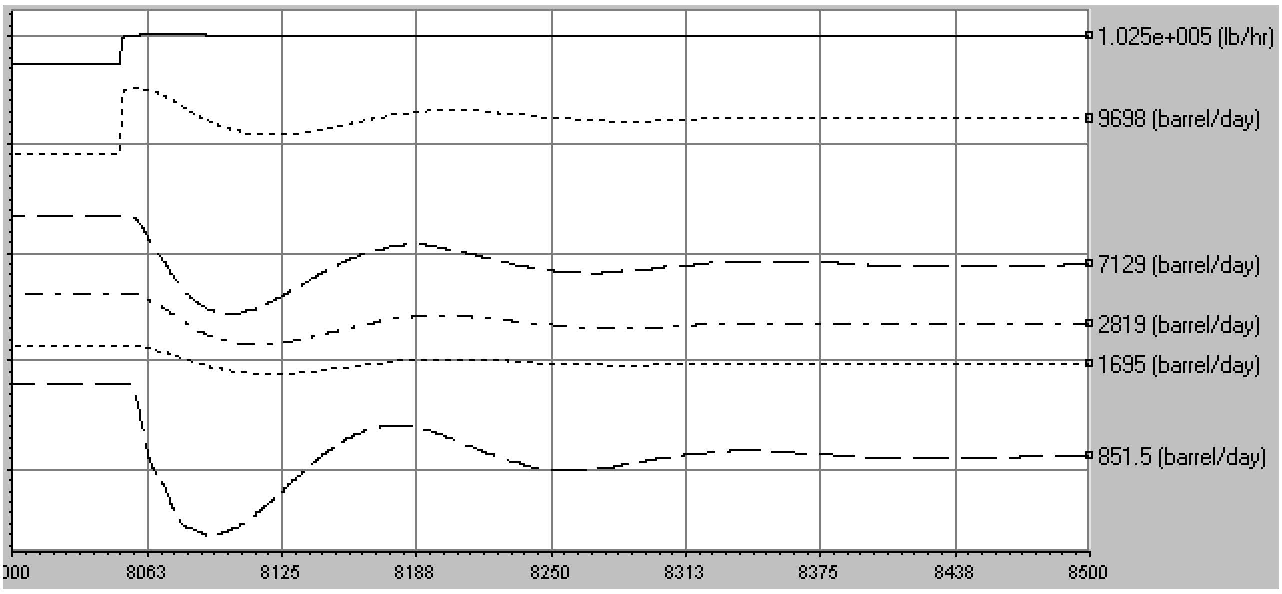

Figure 2b: Naphtha Step Test Flowrate Response

Figure 2b presents the volumetric flowrate changes that result from the Naphtha step change. Each flow is on a separate 1000 bbl/day scale and the responses are again relative. The timescale is 500 minutes. The solid line at the top is the Naphtha step. The dotted line under the Naphtha mass flow curve is the Naphtha volumetric flowrate response. Moving down the plot, the Diesel response is next, followed by the Kerosene, AGO and overflash flows. The Naptha volumetric flowrate responds immediately. The overflash has a 4 minute deadtime and the other flowrates begin responding after between 10 and 21 minutes. None of the responses fits a first-order-plus-deadtime model particularly well. The AGO response is the closest. They all demonstrate a high initial deviation with a decaying oscillation to a final value, much like an underdamped second-order system. This is reasonable; a distillation column is essentially a collection of series-connected first-order mixing processes, one for each tray.

Oscillatory responses like these present a challenge. There are several options available for the selection of the gain, dead time and time constant. One option is to assume a larger deadtime that would bypass the majority of the oscillations and then calculate the gain based on the final value of the process variable, with a time constant roughly equivalent to 1/3 of the oscillation period. This was the first option attempted for this DMC controller and resulted in a matrix with models containing very large deadtimes, of the order of several hours. To correctly function with deadtimes, the DMC control horizon must extend past the deadtime of the process models. The DMC controller required a four-minute sample time to account for the long deadtimes. While the controller would calculate some control action, the true response of the column meant that it would start to swing almost immediately and the response was more oscillatory and destabilised the column.
A second option is to use the true deadtime, final value of the process variable and 1/3 the value of the oscillation period. This produced a stable, oscillatory response but the oscillations continued for several hours and in extreme cases dried out the lower sections of the tower. The reason for this behaviour is that the model gain reflects the final settled value of each process variable and not the true response as it settles. The DMC controller calculates the manipulated variables based on a lower gain than is really present.

The best performance was obtained with parameters calculated by assuming that the initial response provided the true process deadtime, gain and time constant. The ultimate gain is incorrect with these assumptions, but during transitions the response predicted by the controller over a short time horizon is much closer to reality. In addition, the controller will be less aggressive. The process model gains are larger, so the optimiser will calculate a smaller change in the manipulated variables.

The responses to the other mass flow step-tests were similar, but with fewer oscillations. The furnace temperature step-test significantly affected the overflash flow. This is to be expected; an increase in furnace temperature increases the vapour traffic in the column, which in turn increases the condensing duty and internal reflux and the overflash flow will then increase. The other effects of a furnace temperature increase are to reduce the volumetric flowrates and cutpoints of all the products. More light hydrocarbons are vapourised, decreasing the average boiling point of each product and reducing the liquid density. The effect is more pronounced in the heavier products. Most of the lighter Naphtha components will already be vapourised at a furnace outlet temperature of 600+ F, whereas Diesel with a 95% cutpoint at 644 F is significantly affected by changes to the furnace temperature.

The 8 x 5 system contains 40 first-order-plus-deadtime process models. The advantage of a dynamic simulation is that the step tests are very rapid and economical. Many different steps may be investigated to provide the best step response for each variable. The completed DMC controller is unconstrained, with a 100 minute time horizon, a 50 minute prediction horizon, a one-step control horizon and a one-minute sample time.

CASE STUDIES

Three case studies are presented to demonstrate the MPC performance on the tower operation. The first is a large increase in demand for Kerosene. The second is a decrease in the overflash flow for more efficient operation. The third is a major disturbance where the furnace is suddenly constrained to a lower outlet temperature. All temperatures are on separate 50 F scales and all volumetric flowrates are on separate 1000 bbl/day scales so that observed changes are relative. The initial controller setpoints at lined-out conditions are presented in Table 1.

Table 1: Baseline DMC Setpoints

Variable	Setpoint	Variable	Setpoint
Naphtha volume flow	10120 bbl/day	Diesel volume flow	7850 bbl/day
Naphtha 95% cutpoint	356 F	Diesel 95% cutpoint	644 F
Kerosene volume flow	2300 bbl/day	AGO volume flow	1515 bbl/day
Kerosene 95% cutpoint	464 F	Overflash flow	500 bbl/day

Increased Kerosene Demand

This example demonstrates an increase in demand for Kerosene, from 2300 bbl/day to 2800 bbl/day, ramped up over two hours. The column responses are in Figures 3a and 3b below. The time axes are in hours.

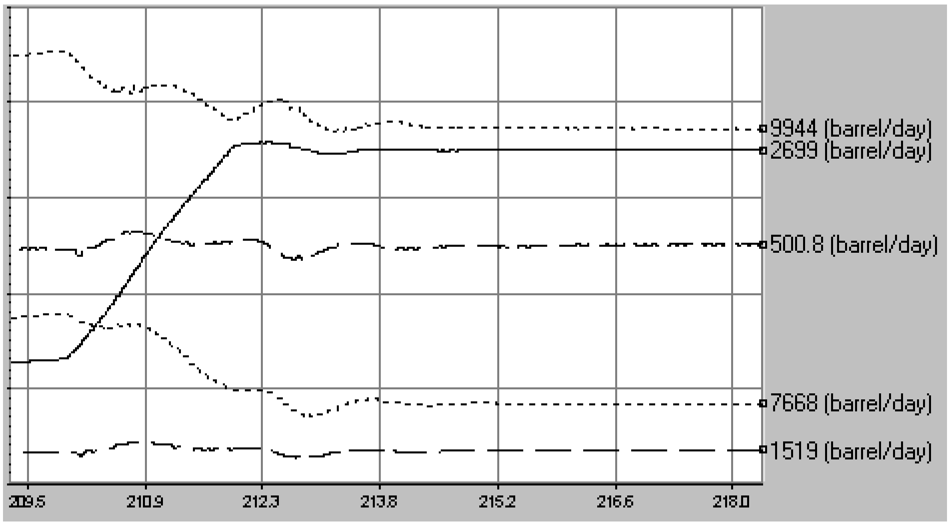

Figure 3a: Kerosene Increase Flowrate Responses

The Kerosene flow is the solid line and it linearly increases to about 2700 bbl/day. The Kerosene is unable to be maintained at 2800 bbl/day without compromising quality. The Naphtha flowrate is the upper dotted line and decreases from 10100 bbl/day to 9944 bbl/day. The Diesel is the lower dotted line and decreases from about 7850 bbl/day to 7668 bbl/day. The overflash is the dashed line in the centre of the plot and is controlled well. The AGO flowrate is the lower dashed line and is affected very little. The compromises made by the controller are clearly evident in the plot. There is a gain of 400 bbl/day on the Kerosene but a net loss of about 340 bbl/day on the Naphtha and Diesel products; the Kerosene value would have to be high to economically justify this change.

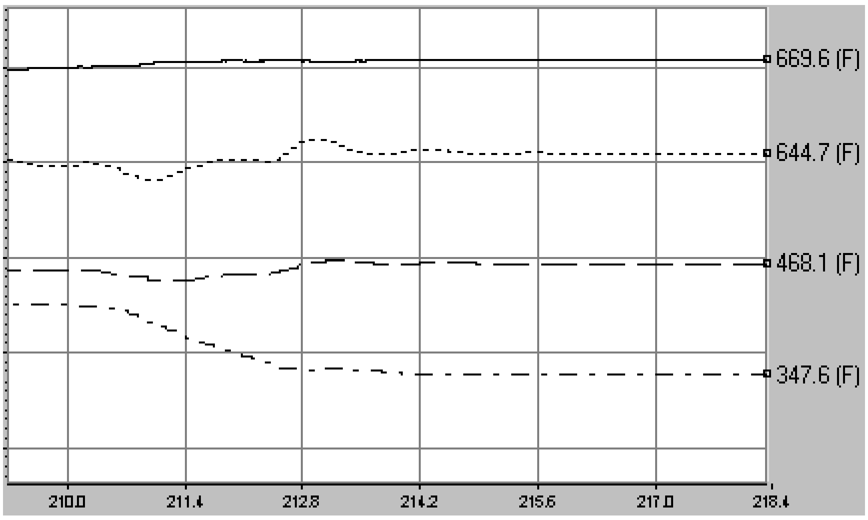

Figure 3b: Kerosene Increase Cutpoint Responses

Figure 3b indicates the temperature effects of the Kerosene increase. The furnace temperature is the solid line and increases from about 668 to 669.6 F. The Diesel 95% cutpoint is the dotted line and oscillates mildly before returning close to the 644F setpoint. The Kerosene 95% cutpoint is the dashed line and is controlled well. The Naphtha 95% cutpoint is the dot-dashed line and decreases significantly from 355 to 347.6 F. Depending on the refinery, a cutpoint decrease of this size is probably approaching the acceptable quality limit for the Naphtha draw.

Decreased Overflash Flow

This example demonstrates the effects of decreasing the overflash flow from 500 to 400 bbl/day, ramped down over two hours. This corresponds to an overflash percentage decrease from 1.25% to 1%. The column responses are in Figures 4a and 4b below. The time axes are in hours.

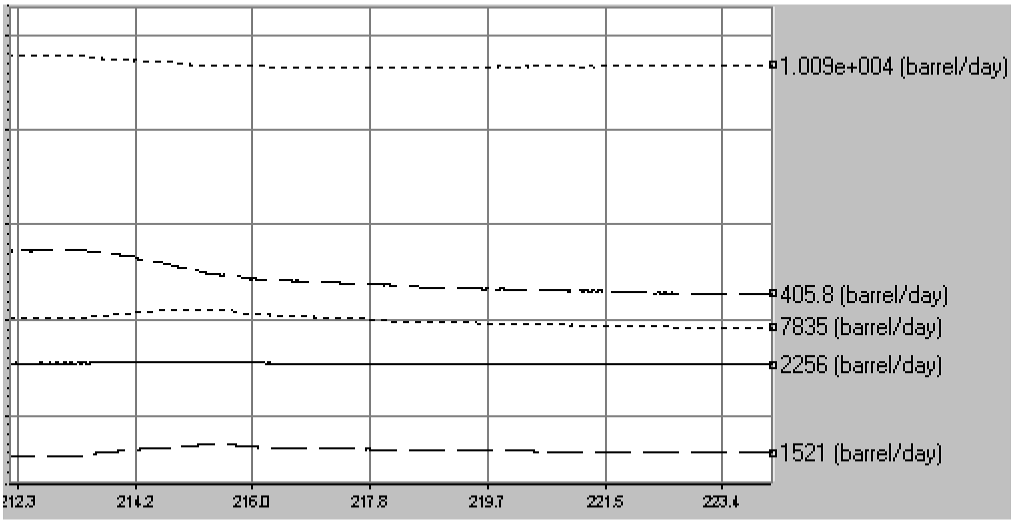

Figure 4a: Decreased Overflash Flowrate Response

This plot indicates a very slight decrease in production. The Naphtha flow is the upper dotted line and decreases from 10100 to 10090 bbl/day. The Diesel flowrate is the lower dotted line and decreases from 7850 to 7835 bbl/day. The Kerosene flow is the solid line and changes very little. The AGO is the lower dashed line and it deviates mildly before returning to the setpoint. The overflash flow response is very smooth.

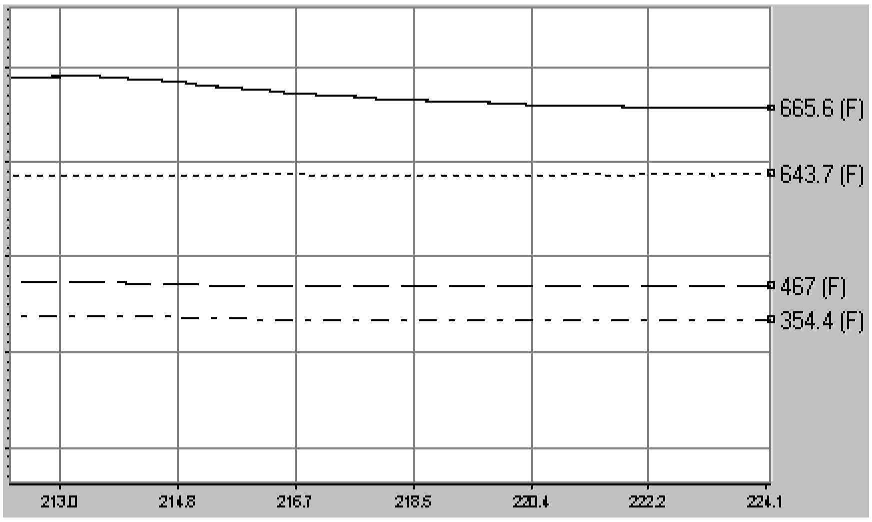

Figure 4b: Decreased Overflash Cutpoint Response

The benefits of decreased overflash flow are more apparent in Figure 4b. The 95% cutpoints of the products are essentially unaffected by the change, but the furnace temperature decreases from about 669.5 to 665.6 F. Over the long term this would probably provide a useful operating cost reduction.

Major Furnace Disturbance

This example demonstrates the effects of suddenly constraining the furnace outlet temperature to 660 F. The column responses are in Figures 5a and 5b below. The time axes are in hours.

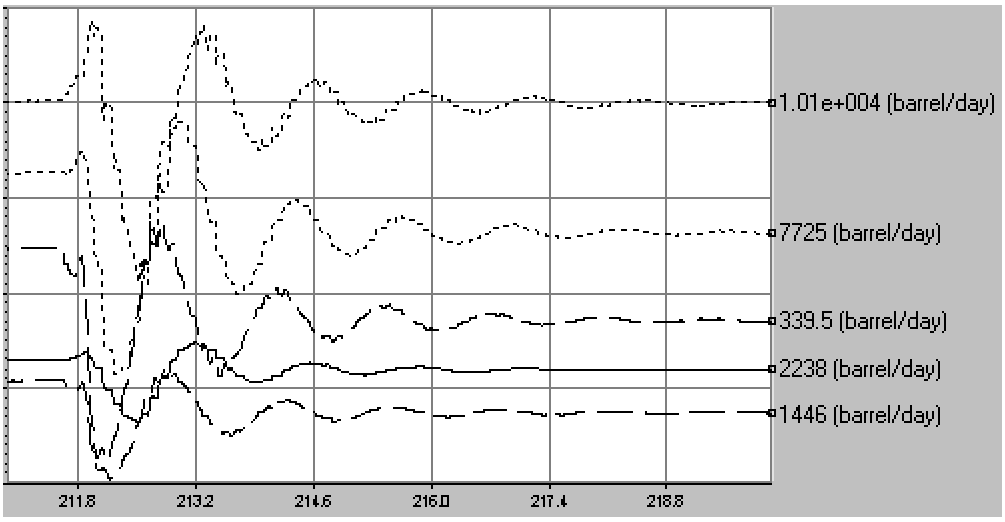

Figure 5a: Major Furnace Disturbance Flowrate Response

Constraining the furnace temperature to 660F effectively removes the furnace temperature setpoint from the DMC controller because the production setpoints require a furnace outlet temperature close to 670 F. This is a severe disturbance. The DMC controller is left with only the product flowrates to control the system. The Naphtha is the upper dotted line, the Diesel is the lower dotted line, the Kerosene is the lower solid line and the AGO is the lower dashed line. All the product flowrates are seriously disrupted by the furnace disturbance and at the same time the DMC controller is attempting to regulate the flows to keep the column under control. The overflash flow is the upper dashed line. The early oscillations are almost enough to dry the lower section of the column and after several hours the flow settles to around 340 bbl/day. There is some lost production on the Diesel and AGO streams. The Diesel flow decreases from 7850 to 7725 bbl/day and the AGO flow decreases from 1515 to 1446 bbl/day. This is to be expected; the heavier streams are more influenced by changes in furnace temperature.

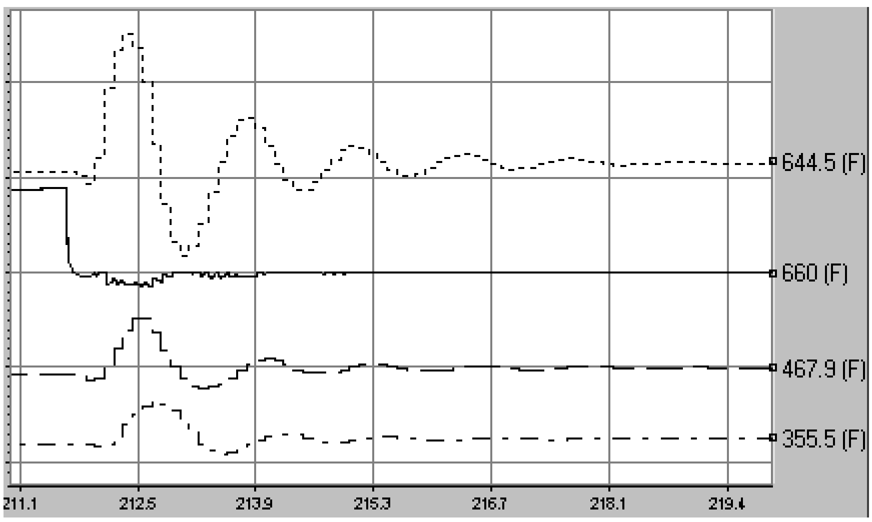

Figure 5b: Major Furnace Disturbance Cutpoint Response

Considering the magnitude of the disturbance, the 95% cutpoints are not seriously affected. The Diesel cutpoint is the dotted line and peaks at about 659 F before oscillating back to 644 F. The Kerosene is the dashed line and peaks at 473 F before settling. The Naphtha is the dot-dashed line and peaks at 359 F before settling.

The furnace disturbance demonstrates the robustness of the controller. It also indicates that while it may be possible to reduce the overflash below 400 bbl/day, production of the heavier streams is likely to become constrained as the furnace temperature decreases.

CONCLUDING REMARKS

A dynamic simulation of an atmospheric crude tower with a DMC controller has been developed in HYSYS.Plant. The model represents the general behaviour of a crude tower well and serves as a useful process trainer. The effects of production changes and the necessary compromises between quality and production on a crude tower are clearly demonstrated. The scope of the advanced control could be easily expanded to include the 5% product cutpoints and by making more manipulated variables such as the steam rates available to the DMC controller.

Dynamic simulation has considerable potential for increasing process understanding through rapid, cost-effective investigations of plant behaviour. For a specific plant and a complete DMC implementation, a higher level of detail is required. A typical large-scale operator trainer model will reproduce the plant behaviour within 10%. A simulation of this accuracy could be expected to generate very good process models for a DMC controller. Further advantages are the plant step-tests may be performed off-line, repeated if necessary and performed overnight, while PCs are usually idle. The prototype MPC may then be tested and fine-tuned off-line with no process impact.

ACKNOWLEDGEMENTS

The authors would like to thank Ron Chartres and Errol Goberdhansingh of Hyprotech Ltd. and Gloria Chukman of KBC for their assistance.