The Dawn of a New Era of Model Predictive Control
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Constrained Multivariable Predictive Control (CMPC) has a long successful track record of industrial applications in the process industry. Since the inception of Dynamic Matrix Control in Shell in 1970s, the number of installations reaping advanced control benefits have steadily increased as evidenced by the repeat business and revenue of companies that offer CMPC software and services.

Two years ago, Solutia started a program to implement CMPC resulting in about a dozen controllers all of which have met or exceeded the expected benefits. In contrast, longer efforts to use other advanced control tools have yielded a total of only 4 ongoing applications. Most of the success of CMPC can be attributed to the fact it is a closed loop technology based on extensive systematic process testing that is tied directly to economic goals. Advanced control technologies that are more advisory and theoretical, and less focused on benefits and testing than CMPC, do not go online or stay online as often.

The implementation of traditional CMPC typically takes a lot of effort and expertise. Most customers rely upon the use of outside consultants for part or all of the implementation. The success of a CMPC company often depends more on the quality of its services than features of the software. The entry level cost is also prohibitive to all but the few who are convinced that large benefits can be achieved. Management is often accustomed to thinking that additional capacity or yield required the installation of new equipment. Convincing management that these results can be achieved by just applying control techniques is normally a hard sell. This is particularly the case when the cost of engineering to implement CMPC controllers for a medium application (50 manipulated variables) can range anywhere from $200K to $2M. While, the price of the software is typically 10% or less of a project's budget, this in itself is a hurdle because each purchase of software requires significant justification, approval and implementation procedures.

While Solutia is addressing these conventional CMPC opportunities, there are as many or more non-traditional potential applications especially for batch and sheet control. If a small CMPC could readily replace the single loop advanced regulatory applications such as feed forward and dead time compensation, the number of new applications would be enormous. If the access to CMPC could be improved and the cost and difficulty of implementation greatly reduced, the rate of application could greatly accelerate.

A closer look at the CMPC implementation shows that the greatest project cost and effort is for testing the process (bump, PRBS, and step testing). Frequently, customers depend upon consultants to do all of the testing. Since the testing is typically manual and extends into the night and weekends, the hourly rates are higher than initial estimates with the actual costs overrunning by as much as 35%. Also, instrumentation and control valve problems, slow process responses, and frequent startups and shutdowns can more than double the time needed for testing. The List of typical CMPC Project Activities is shown below and assumes the software of the engineering tools and the online controller is already installed and an interface for both the CMPC and the data historian already exists.

List of CMPC Project Activities

Historical percentage of effort and desired allocation of levels of expertise (1=generalist 2=specialist and 3=consultant) are shown in parentheses.

- ***Process Analysis (10%) (3)***
- ***Bump Testing (10%) (2)***
- ***Regulatory System Improvements (10%) (2)***
- ***Data historian access (5%) (2)***
- ***PRBS and Step Testing (25%) (1)***
  - *testing time increases dramatically (2x) for slow and difficult units*
- ***Dynamic Modeling (15%) (2)***
- ***Controller Design (10%) (3)***
- ***Operator Interface (5%) (2)***
- ***Commissioning (10%) (2)***

Fortunately, a new approach to CMPC implementation is now available that promises to revolutionize this whole process. It offers an easy addition and configuration of a model predictive control (MPC) as part of a Field Based System. Today, a CMPC application has been pre-engineered into a single control block available as part of a standard automation system configuration. This MPC block is another choice of standard configuration blocks like the Fieldbus PID or fuzzy logic control block. This allows strategies that include the MPC block to be quickly constructed using dragged and dropped into a configuration graphic style sheet. An example configuration of the bottom composition control for a distillation column is shown in figure 1.

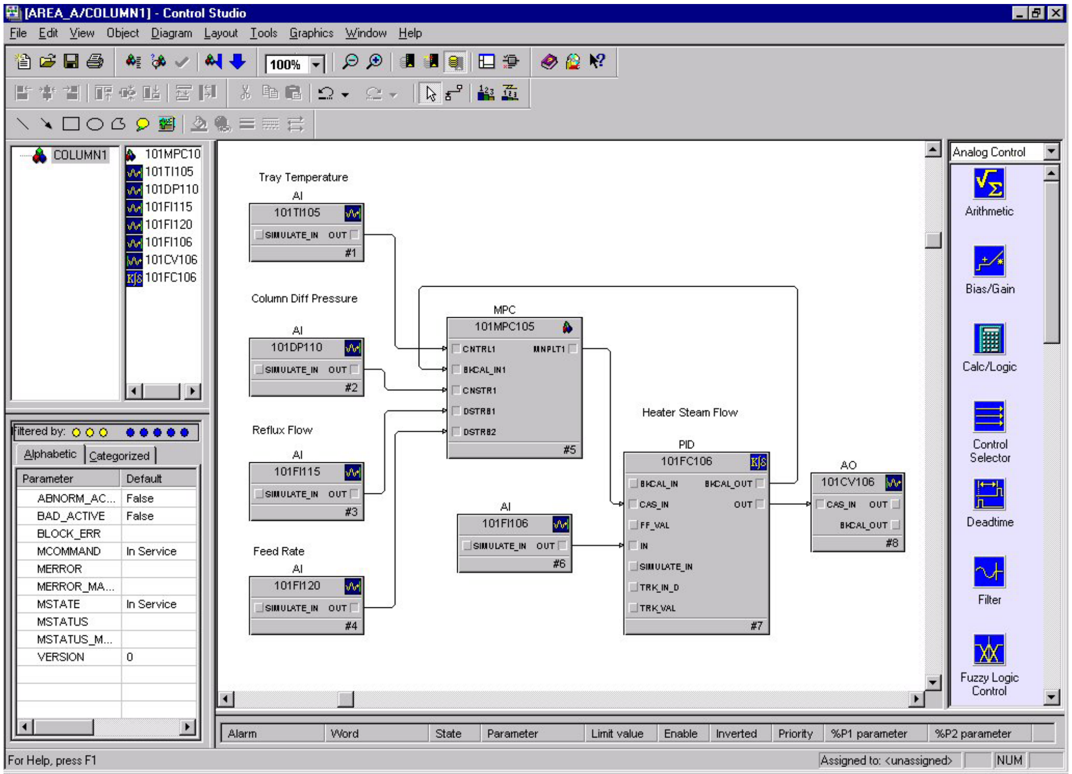

Figure 1
Application of the MPC Block in Distillation Column Control
(Courtesy of Fisher-Rosemount)

One of the first applications of the MPC block was a distillation column. It was found during a review of the process control system that the tray temperature and sump level loops were oscillating. Additionally, for an increase in feed rate, by an increase in steam flow to the feed vaporizer, the column could flood or the temperature gets too high. For decreases in feed rate or temperature, the column temperature might drop so low it would have to go on recirculation. During bump testing, it was found that the level loop, which manipulated the bottom flow, was fast, noisy, and upset by changes in the changes in steam flow made by the temperature loop. It was also discovered that the temperature loop had a very high and nonlinear process gain besides a slow response. The plot of temperature versus steam to feed ratio from a high fidelity simulation model resembled a strong-acid base titration curve. The control point was just below the steepest portion of the curve. An increase of just 1% in steam flow would cause the temperature to sky rocket and end up hung up on the upper flat portion of the curve. Furthermore, since it was a thermosyphon reboiler, changes in sump level could affect the circulation and liquid in the reboiler.

The solution was to add a 9 second measurement filter, increase the reset time from 60 seconds to 300 seconds, and increase the gain from 2.0 to 4.0 in the PID loop for sump level. This eliminated the reset cycle and provided tight level control. A filter was also added to the differential pressure across the column that was an indication of flooding. The Model predictive control application was then able to focus on the prevention of flooding and off spec product. The MPC block was configured within a few minutes for temperature as a controlled variable, steam as a manipulated variable, feed and reflux flows as disturbance variables, and differential pressure as a constraint variable. The engineering interface shown in Figure 2 was used to test and build models. With just the temperature loop in manual, very small steps of 0.25 to 0.5% were made in the steam flow.

All of the model's parameters were identified automatically and rapidly, without disturbing normal operations. This automatic model extraction eliminates the need for the expensive consulting services normally required with classical CMPC applications. It also enables the control engineer to insure the integrity of the sampled data used to determine the model parameters.

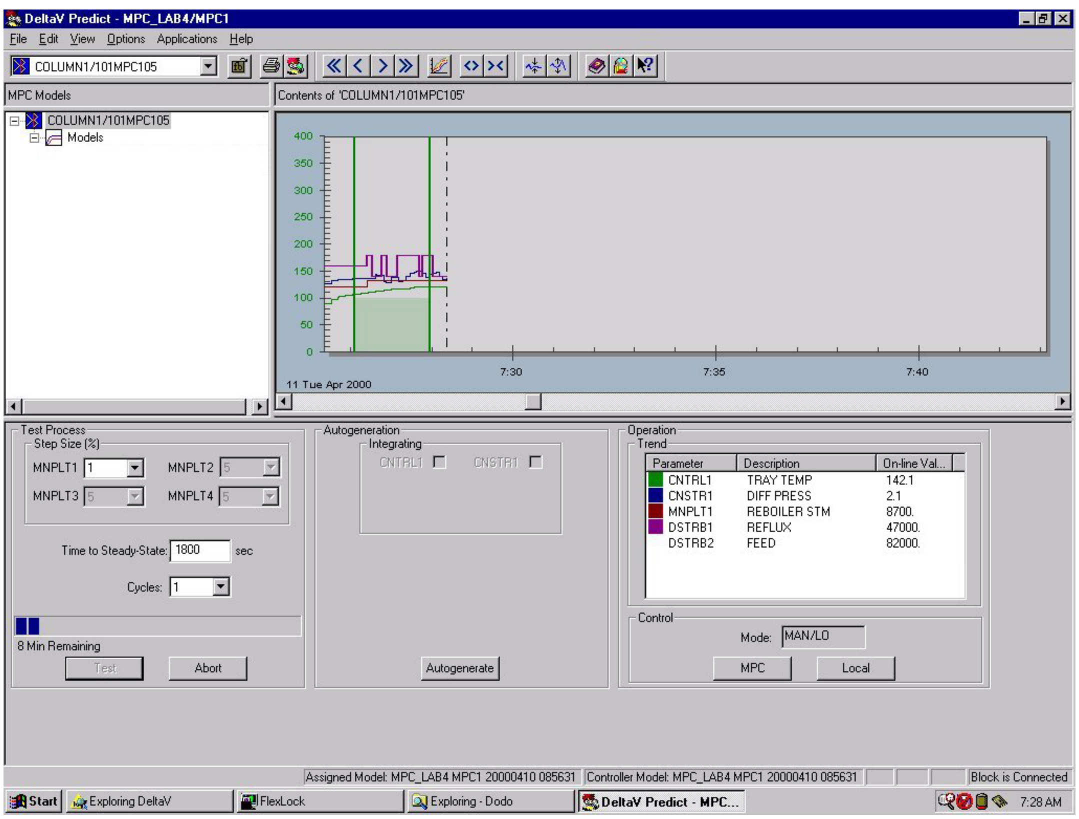

Figure 2
Engineering Interface for Model and Controller Building and Testing
(Courtesy of Fisher-Rosemount)

The operator interface as shown in Figure 3 can be launched from a graphic operation display or through the engineering display. It automatically has the faceplates and trends set up for the MPC block. The modes in the model can be changed individually or as a group. The limits for the controlled variables, constraint variables, and manipulated variables are shown graphically in the "boxes" in the graphic display. This display is particularly useful during the commissioning phase. For maintenance and troubleshooting, the percentage of time each variable is at its limit would be displayed. Additionally, an analysis tool built into the system can automatically keep track of inputs and outputs at their limits, uncertain or bad inputs, standard deviation, performance compared to minimum variance control, and utilization (per cent of loops in highest design mode) for the control system. This tool along with the built-in auto tuner is important for improving and establishing a baseline for the regulatory control system before the MPC is added.

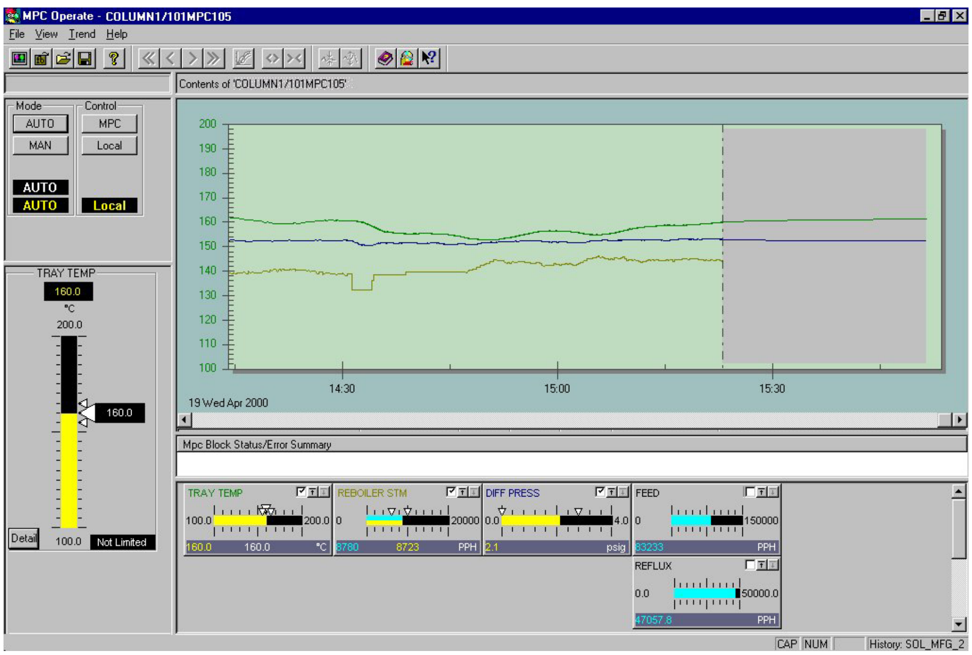

Figure 3
Operator Interface for a Model Predictive Controller
(Courtesy of Fisher-Rosemount)

The variability in the column was greatly reduced. Future possible improvements could use data from the HYSYS model for the column, signal characterization of the temperature measurement to linearize the temperature response, pressure compensation of the temperature set point, and the addition of the steam flow to the feed vaporizer as an optimization variable to push the production rate.

The use of a Windows NT environment for the Field Based System facilitated the development of a virtual plant where the entire system (controllers, operator interface, and advanced control tools) can run on a laptop. The actual system configuration and displays can be generated and exported or imported from a real system. Figure 4 shows the interface for the professional version where the controllers can be run 30 times slower or faster than real time and can be stopped. The blocks can also switched between simulation and normal modes. The signals can be connected to a dynamic model of the process such as HYSYS Plant via Object Link Embedding for Process Control (OPC). When the model is started, the analog outputs are initialized by the HYSYS Plant and the controller outputs and any intervening blocks used for signal selection, split ranging, or characterization are automatically initialized via the BCK-CAL function. In Solutia, we use this capability to test configurations before the hardware is ordered, try out new advanced control tools, conduct control studies, and train operators. We also use the virtual plant to provide initial and supplemental test data to develop and test models for model predictive controllers and online property estimators. The virtual plant is proving to be helpful in identifying the degree of variation in process gains and time constants.

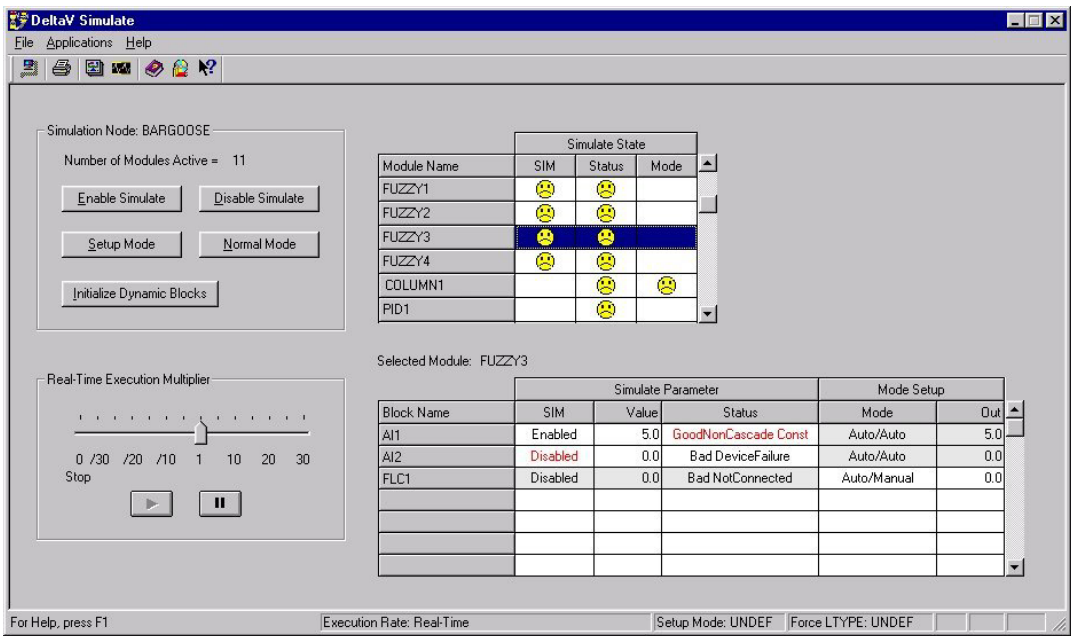

Figure 4
Engineering Interface for the Virtual Plant
(Courtesy of Fisher-Rosemount)

The easy addition and configuration of the MPC block and integration and automation of other implementation activities such as process testing and building data, engineering, and operator interfaces, reduces the initiation effort and opens the door for process control engineers to learn how to apply CMPC. It is acknowledged that process knowledge is still very important and an understanding of basic control essential. It is also recognized that today most of the application experience for CMPC resides with consultants. As with most other areas in industrial control, the rules for designing, building, and tuning these controllers are not often well documented but are learned the hard way on applications. Since the traditional CMPC implementation has been so time consuming and selective, there has be little opportunity for the process control engineer to participate. The application expertise and additional understanding of the process ends up with the consultants from the CMPC service and software companies. It is interesting to speculate that expanding the role of the user in implementing CMPC might actually benefit CMPC companies by increasing the number of applications and elevating the role of consultants to process analysis and controller design. This leads to optimum use of resources listed in the parentheses in the list of CMPC project activities that can reduce the cost of the project by 50%.

There are significant opportunities for small CMPC applications to replace PID controllers. Derivative is rarely used due to resolution limits, process and measurement noise, sample and hold, inverse response, and interactions. Many theoretical candidates have an irregular or abrupt response that rate action would amplify and pass on to the controller output. In temperature loops, the A/D dither from fast scan times and large calibration spans and in composition loops, the stair casing and scatter of readings from analyzers preclude the use of derivative action. A CMPC can out perform a PI controller and is more robust. If the dead time estimate is off, a CMPC will bias the trajectory so that current predicted reading is closer to the actual reading of the process response. Since the control action is based on minimizing the control error for the trajectory and not just the current control error, it is less sensitive to resolution limits, dither, and jitter. The change in PI or PID controller output is entirely based on what it sees at each scan whereas the move made by a CMPC is based on a process model. For these reasons and others, a CMPC can better withstand the nonlinear and erratic response of industrial process measurements.

PID controllers are rarely completely and accurately tuned. Even the computations from an auto tuner are normally based on just 3 to 5 steps or cycles of a relay method. A CMPC uses an order of magnitude more process testing and identification. Thus, even if derivative action can be used and the dead time is relatively constant, a CMPC will generally perform better because it employs more extensive and systematic testing and better process knowledge.

The implementation feed forward control is readily accomplished in CMPC where as in PID it is more ad hoc. Feed forward gains are often in error and dynamic compensation minimal. The process model built into a CMPC between a disturbance variable and each controlled and constraint variable offers an opportunity not realized in PID loops.

Dead time compensators such as the Smith Predictor only can look one dead time ahead and are very sensitive to over estimates of the process dead time. A CMPC looks as far out into the control horizon as desired and shifts the trajectory based on the actual dead time. Consequently, a CMPC can maintain a much higher level of performance.

Override control systems are effective for small applications of constraint control in both batch and continuous processes. In these systems a signal selector is used to choose the lowest or highest output of PID controllers whose process variable is a constraint variable and whose set point is set just inside the constraint limits. Table 1 shows the % of time an operator, override, and control system can hold coincident constraints. For two or more coincident, a CMPC is much more effective.

Table 1 How Well Can Coincident Constraints Be Handled?

Number of Coincident Constraints	% Time Operator Can Hold	% Time - Override Can Hold	% Time CMPC Can Hold
None	70%	10%	2%
One +	30%	90%	98%
Two +	20%	45%	90%
Three +	0%	30%	80%

Traditionally, a CMPC executes every 30 to 60 seconds. If this is more than 10% of the total dead time or largest process time constant, there is some appreciable degradation in capability of the control system due to the increase in dead time from the execution time. The minimum peak error is the extent of the process excursion during the dead time. The availability of an MPC block that can execute every second opens up much faster applications to the advantages of CMPC. Today, CMPC applications can be embedded directly in controller software and as a result can execute at these rates.

Control valves with a large dead band or a poor resolution are a problem for both PID and CMPC, particularly for sensitive processes. If the valve doesn't move for small changes in controller output, nothing happens. The minimum move size for a CMPC can be increased to exceed the dead band but the changes when the vale does move might be too large. If there is slip besides stick, the situation is even more difficult and the deterioration in control more significant.

Operating on the flat portion of a installed characteristic of a control valve is not desirable for either a PID or CMPC. Ideally, the control valve, pump, or piping system should be replaced. In practice, signal characterization is used to get the most out of the least. The signal characterizer increases the change in the valve signal for a change in the controller output per the slope of the flat part of the installed characteristic. It reduces the effect of valve dead band and dead time. Today it is possible to modify system configurations by inserting a signal characterization block between a MPC output and an analog output block.

From the above discussion, it is possible to see that a small CMPC application can start to replace PID applications since it is accessible and easy to implement and try out. This combined with the synergy offered by other advanced control tools such as performance analyzers, auto tuners, virtual plants, and field bus blocks, signals the dawn of a new era of model predictive control where its implementation is not relegated to a selective few conventional applications and consultants.