MPC in Column control

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Abstract: This paper presents the use of a multivariable model predictive controller from ABB in a column control application. Some of its more important features are that it has three degrees of freedom and that it is based on discrete time state-space models obtained from a truly multivariable identification procedure. Control performance and handling issues of the mainly two-component distillation column is discussed in some detail. Control performance is documented via quality indexes both prior and after the changed control solution.

Keyword: multi variable control, state-space models, three degrees of tuning freedom, column control, MPC, loop performance indexes, loop diagnostics, quality indexes

Background

Distillation columns are probably the most common process unit within the chemical industry. In some instances it has been considered to be a well know area from control engineering point of view while information from production units has shown that performance do not meet performance expectations from production. Making the regulatory control application solutions more complex could reduce some of the performance problems but would then instead supply maintenance problem in the long run.

Looking at it this way supplies a new challenge, i.e. highest possible performance while still keeping the application solution as simple as ever possible. The trade off between performance and smooth integration in the overall existing site conditions will mean that the new solution will have to be adapted for a number of different soft and often unspecified customer conditions.

This includes not only functions with a dependency on the physical process and equipment but also on the people available at site. Another condition commonly put forward has been the possibility to run exactly as done before without having to do more then just to push a button on the traditional operator workplace, used by the production people.

With those conditions as a new main goal when implementing Model Predictive Control solutions means that the existing equipment and installation basically will have to be kept “as is”.

Reaching the very highest performance requirements in combination with a solution using MPC controller, normally requires installation also of new and advanced measurement devices. In some cases the number of new measurement devices have been fairly high. The trade of between the payback via production improvements and the cost of the additional measurement devices with continuous maintenance could be hard to motivate when the additional cost of maintaining the MPC solution comes into the picture.

The distillation column control solution, put forward here, has been done without any changes at all in the site equipment and the possibility to run “as is” from before remains.

Process overview

The distillation column is designed for a binary component mixture. Via the additional supply lines also other components could be added from other production parts. The principal main binary component feed mixture then becomes a feed with up to 6 different components. The basic components should be BUOH (Butanol) and water. The bottom product should be the 100% pure Butanol and the top product is a mixture of water and left over butanol in an azeotropic mixture.

The temperature difference over the column is around 25 degree C with a currently selected target “mid” temperature of 103 degree C.

Typical problem areas for this column have been the feed changes with associated changes in the feed composition mixture. The level in the buffer tank is used as reservoir and the level is pushed low if supply disturbances are known to come later on. There is no required target level for the buffer tank which means that the feed changes could be required to change in a more or less active manner to keep the level not too high or too low, depending on later production situations. The buffer tank is feed with mixtures of different density. Via the level in the tank the residence time change and thereby also the aggressiveness of the component mixture changes in the feed.

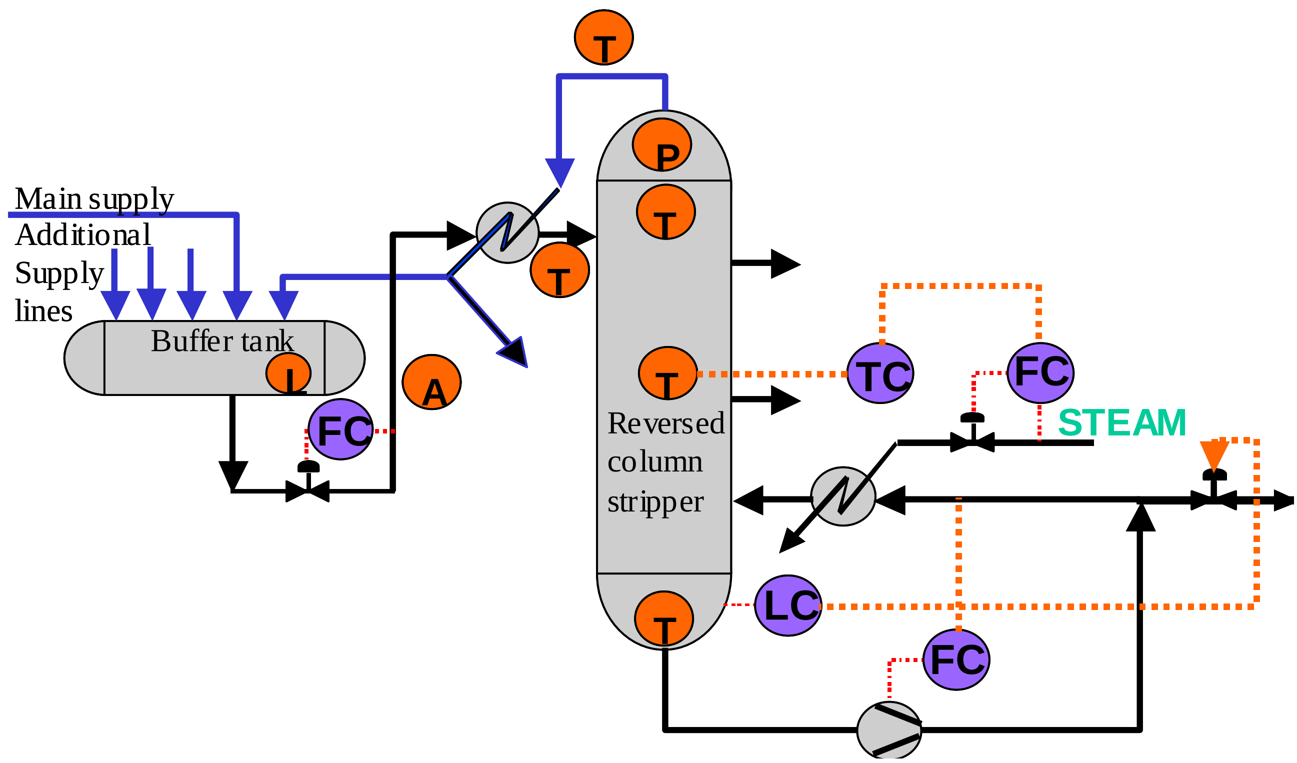

Figure 1.0: BASF C-vitamin production in Denmark, Principle process diagram

Downstream from the column the heavy component must be 100% pure if it should be accepted in the following production steps. If the down-stream production steps have problems more re-circulation material is also feed back which in itself also supplies production problems. Re-circulation is also dependant on the level in the bottom of the column and conditions based upon temperatures in the stripper column.

Principle column solution

The initial solution for the column included use of all available signals, i.e. 12 measurement devices. These signals would then be used for feedback and feed-forward purposes. The aim was to make use of the available 5 PID controllers and set their setpoints from the MPC controller.

A detailed test pattern for each of the manipulated variable was created. The test sequence was discussed with personal from the production and instrumentation side. Based on information from conventional loop tuning steps, acceptable sizing was defined on the changes and included in the test pattern. Signals were there was no direct possibility to influence the process variables a similar test pattern was done to supply the necessary process dynamic information. Those changes were done further up-stream from the stripper column.

Process excitations tests were run during two main test periods. Data and results from those tests were then evaluated from different aspects, such as;

- Acceptable operational conditions
- Physical product conditions
- Equipment usability and functionality
- Process model accuracy
- Process model sensitivity and robustness

Above steps are necessary if the model predictions should be useful and supply a reliable multivariable process model.

With above as a base, the controller design calculations were done to create a controller with robustness for process dynamic changes and at the same time more aggressive for disturbances.

Based upon the state space model building technology the number of signals finally used for the application was reduced as they supplied too little improvements in process model accuracy or predictions. Further reductions in the number of signals were done were the improvements could be “discussed” but then from the point of view to make the process more easily maintained and handled.

Measured performance and results

Performance analysis has been done using data from production during at least 6 weeks as a reference period. With those data as comparison a new period was acquired but then using the new application solution with the 3*d*MPC controller.

Performance comparison was done such that session lengths of 5 hours were selected as a representative period for performance index calculations using data acquired each 30 seconds. This thereby supplies approximately 210 sessions over the reference period. Each session supplies individual performance index values of different kind. Trend curves of those calculated quality indexes are displayed in below graphs. A comparison is then done between the two periods for some of the quality indexes supplied by the PCT Loop Optimizer Suite program used at site.

One very essential index is the standard deviation of the loop deviation on the “mid” column temperature. This index supplied a significant change. The ratio in improvement was calculated to a factor of 2.9 compared with the reference period. This ratio was calculated as the ratio of the means between the standard deviation of the reference period divided with the new period. The Ruled Based Knowledge Database from the evaluation tool indicated a change in the “Acceptance” level of the mid temperature loop to increase from 91% to 99%. From this it is evident that the column mid temperature was controlled fairly well before the change over, i.e. when the MPC controller was put into operation.

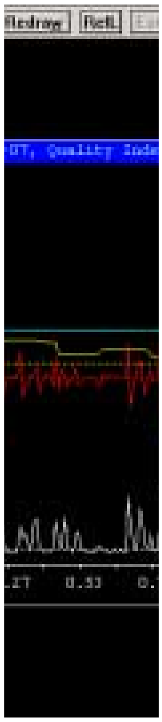

Figure 2.0: MPC based values of main quality indexes for column mid temperature, 2.1: PID based control

In above graphs the individual colures displays the following:

- Cyan line is work position equal to 103 degree C
- Red line is the average loop deviation
- White line is the standard deviation of the loop deviation.
- Yellow line is the average feed flow

The above 4 displayed indexes are just a small amount of the available quality aspects, made available by the loop diagnostics tool.

Via the smoother control achieved using the 3*d*MPC a more robust temperature gradient is achieved in the column. As a result of the better temperature gradient in the column, energy savings can be achieved by decreasing the column mid temperature. The temperature decrease can be based on the improvement factor of the standard deviation. The estimated temperature decrease would be in the range of 1.2 – 1.6 degrees. A lower mid temperature in the column creates a more pure azeotropic mixture in the top and thereby even less energy need for the reduced amount of Butanol in that mixture. This thereby creates secondary reductions of energy use. The decreased mid column temperature must be made with the constraint that the bottom product must remain 100% pure Butanol.

With the temperature reduction indicated above the steam demand is reduced in the order of 10%. Viewed in that perspective a reduction in column temperature could remove the need to re-build the steam boiler to meet increased steam demand from the other production units.

Via a more stable temperature profile in the stripper column all following process sections will get more robust and steady operational conditions. This is achieved in combination with a higher production volume and reductions in energy demand.

Model Predictive Control technology used

The 3dMPC controller uses a combination of feedback and feed-forward methods. The controller determines the manipulated variables (MV) based on actual measurements of the controlled variables (CV) and disturbance variables (DV). The process variables can be assigned set-points that are the target for the feedback control law or they could just be used in the state-observer to improve the state estimate. The feed forward signals are measurable disturbances acting on the process.

At each sample the manipulated outputs are calculated and combined from a sequence of constrained optimisations. The loss function in these problems penalizes a weighted sum of squared control-errors and moves in the manipulated outputs. Constraints specifications participating in the optimisation method comes from requirements on the manipulated outputs and the controlled variables.

The controller works with *three degrees of freedom* (3*d*). This has been the motivation for its name. Three degrees of freedom means that the controller can be configured to have different dynamic responses to:

- set-point changes
- changes in measurable disturbances
- other disturbances and to model mismatch.

The three different optimisation problems are solved with individual control error formulations even if they are formulated in a similar manner.

After the dynamic optimisation method has been performed, a *static optimisation* algorithm is employed to drive the manipulated outputs towards desired values. This algorithm will only have effect in cases where there is ambiguity in how the manipulated outputs are chosen in order to reach the desired targets for the controlled variables.

The inputs and outputs signals of the controller can operate in independent *operational modes*. This means that some signals in the controller can work in automatic mode while others can work in manual mode. The mode for a signal is determined either by the operator or by external inputs to the 3*d*MPC controller module.

A main feature of model predictive controllers is the ability to *handle constraints*, not only for manipulated outputs but also for the controlled variables. Constraints are ranked with individual priorities for controlled variables, deviations and manipulated outputs. These constraints can be violated under exceptional operating conditions. They are then violated according to their priorities. Hard constraints can be defined for manipulated outputs as well as rate of change constraints and they must always be fulfilled.

The 3dMPC is based on a *discrete time state-space model.* The active state vector is estimated/updated using Observer Based Technology. The parameters of the model are determined using truly multivariable identification methods provided by the modeling tools of the 3*d*MPC product.

State space models was selected as the preferred description as it provides some powerful features for the model predictive controller:

- Uniform treatment of stable, integrating, and unstable processes dynamics.
- Allows feed-forward from non-measurable disturbances using additional measurement signals
- Efficient methods to verify usefulness of signals to participate in the process model creation

The controller is able to *handle non-linear processes* in four manners:

- Static non-linearity can be defined for each signal to compensate for known characteristics of measurement and control devices.
- The controller can make use of external input signals for parameter scheduling. One of the four pre-defined sets of controller parameters is then selected. Such a set contains, e.g. a complete process model, weights on control-errors and moves and constraint definitions. This high level of controller parameter scheduling allows complete change of the controller.
- The controller can also use external inputs for known changes of the characteristics in the measurement and control devices depending on e.g. operational load, speed or any other relation.
- The controller can run having variable execution time interval. This makes it easy to handle processes with variable process deadtime, i.e. as found in paper machines, rolling mills and steam boilers.

If the four pre-defined sets that can be changed in an automatic manner are not sufficient, the user can manually select any available definition set as the active alternative.

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