409g Implementation of Coordinator Mpc on a Large-Scale Gas Plant

Elvira M. B Aske, Dept. of Chemical Engineering, Norwegian Univ of Sci & Tech (NTNU), Trondheim, N7491, Norway, Stig Strand, Process Control, Statoil, Trondheim, Norway, and **Sigurd Skogestad**, Chemical Engineering, Norwegian Univeristy of Science and Technology, Sem Sælandsvei 4, Trondheim, 7491, Norway.

In many cases economic optimal operation is the same as maximum plant throughput. This corresponds to maximum flow through the bottleneck(s). This insight may greatly simplify implementation and a real-time optimization (RTO) based on a nonlinear model is not necessary. The key issue is to identify the active "bottleneck" constraint and a coordinator MPC based on a linear model with constraints is proposed (Aske *et.al*., 2008)

The main objective is to maximize the throughput (feeds) of the plant, subject to achieving feasible operation(satisfying operational constraints in all units). With the assumption that the flow through the network is represented by a set of units with linear flow connections, the maximum throughput problem is then a linear programming (LP) problem. The use of MPC to solve the LP has the benefit of allowing for a coordinated dynamic implementation. The constraints for the coordinator MPC are the maximum flows through the individual units. These may change with time and a key idea is that they can be obtained with almost no extra effort using the models in the existing local MPCs. The coordinator MPC receives a remaining capacity measure from each of the local MPC applications (each unit) and thereby tries to maximize throughput by adjusting feed rates and crossovers (between parallel trains). In this paper, an actual implementation of coordinator MPC at the Kårstø gas plant is described.

The plant

The Kårstø gas processing plant plays a key role in the transport and treatment of gas and condensate from central parts of the Norwegian continental shelf. The plant receives rich gas and unstabilized condensate through pipelines and separates the feeds into its various components. More than 30 fields export its gas through the Kårstø plant, and this set high demands to the plant efficiency and regularity. If the gas plant stops or limits the export, the one or several fields must reduce, or in worst-case, shout down oil- and gas export since the fields can not dispose the gas elsewhere (burning the gas offshore is prohibited).

Calculating remaining capacity

The constraints for the coordinator MPC are non-negative remaining capacities ($R \geq b \geq 0$) in all units, where R is the remaining capacity in the unit and b denotes the back off. The remaining capacity for unit k is given by R_k and is the distance between the current feed and maximum feed, $R_k = F^l_{k,max} - F^l_k$, where F^l_k is the current feed to the local unit and $F^l_{k,max}$ is the maximum feed the unit can receive and still operate within the operation constraints.

Most of the distillation columns at the Kårstø gas plant have already MPC installed with two-point composition control. These local (unit) MPC applications are used to calculate the maximum feed rate each unit can receive, subject to given controlled variable (CV) and manipulated variable (MV) constraints. The maximum feed to the unit k is then obtained by solving the

additional steady-state problem:

$F^l_{k,max} = max\ F_k$

subject to the linear model equations and constraints of the local MPC, and this is a LP problem. Here u^l_k is the vector of manipulated variables in the local MPC, and the optimization is subject to satisfying the linear constraints for the unit. To include past MV moves and disturbances, the end predictions of the variables should be used instead of the present values.

Coordinator MPC design

For the total plant, the coordinator MPC consist of approximately 36 CVs, 12 MVs and 3 DVs. However, a sub-application is implemented first, because it is easier and we want to demonstrate the coordinator ability. Another reason is that parts of the plant will be rebuilt shortly and therefore no local MPC application is implemented on this part.

The coordinator MPC is implemented in SEPTIC MPC (Strand and Sagli, 2003). The sub-application consist of 22 CVs (14 remaining capacities, 7 other constraints and total plant feed), 6 MVs (4 feed rates, 1 crossover and 1 feed split) and 5 DVs (2 feed compositions, 1 feed rate, 1 crossover and 1 feed split). The CVs are constraints with low- and/or high limits, except for the total plant feed, which is a set point controlled CV. The total plant feed is

the sum of the feeds rates, which should be maximized. To maximize throughput, the total plant feed set point is a high, unreachable set point with lower priority than the constrained CVs.

All distillation columns in the sub-application (12 in total) have local MPC installed and the remaining capacity is calculated there. The compressors have not MPC control, but here are “dummy” MPC applications to calculate their remaining capacity. The two “dummy” applications consist only of CVs and DVs. The other CV constraints in the coordinator are measurements to avoid emptying or filling pipelines and tanks.

The coordinator MPC receives three values from each of the local MPC: the calculated remaining capacity, quality of the remaining capacity calculation(GOOD/BAD) and status of the local MPC application (ON/OFF). The latter value is used to avoid that the coordinator MPC is active without the local MPC applications is active.

Each variable (CV, MV and DV) is located in one or more sub-groups that will deactivate if one critical variable is deactivated. For instance, if a local MPC application is turned off, the corresponding remaining capacity is deactivated, which leads deactivation of the whole sub-group, including the correspondingfeed rate. The coordinator MPC can then operate even if parts of the plant is no running or not available for throughput maximizing for some reason. For the coordinator MPC, each MV defines a sub-group with corresponding CVs as members.

Coordinator MPC modelling and tuning

The models in the coordinator MPC are all SISO step-response models. The models are obtained from step tests and historical plant data. The steady-state gain in the models from the feed rates is calculated from a typical feed composition to validate step-tests. The prediction and control horizon is set to be 6 hours, whereas the longest models are approximately 4.5 hours. The coordinator has 4 integrating CVs, and for these the control horizon is a tuning parameter that can be set individually. Here they are selected to half the prediction horizon, which is 3 hours.

The tuning of the coordinator MPC is a trade-off between robustness and MV (e.g. feed) variations on one side and keeping the flows through the bottlenecks close to their maximum on the other side. A rational use of the MVs is prioritized to avoid unnecessary variations in throughput.

Experience from implementation

The performance of the local MPC applications are crucial for the coordinator MPC to work as intended. The weakness is the calculation of the remaining capacity in each unit. The goodness of the models and in particular the steady-state gain is important. Our experience is when a larger disturbance occurs, the predicted steady-state values may violate limits and if this violation is large enough, the LP optimization do not find a feasible solution, and the maximum capacity calculation (Flmax) turns bad. The end prediction value is in such cases often not reasonable because the MPC application assumes that the disturbance will persist constant (or with a low-pass filter)throughout the prediction horizon, which is seldom the case. To overcome this problem, several approaches are used:

With a known, measured, short-time disturbance: the maximum capacity (F_max) is held constant throughout the disturbance. This is used for the disturbance that occurs at each dryer exchange.

Include a minimum value of maximum capacity, Flmax, that the unit for sure is able to process.

Include gain scheduling in models to describe typically non-linear behavior.

The coordinator MPC is also vulnerable to model errors in the local MPC applications. The local MPCs consider in general product quality control within some operational constraints. However, due to feed back, acceptable product quality control can be achieved even with some model errors. On the other hand, the coordinator MPC uses the models in the local MPCs directly for estimation of remaining capacity. Thus, model errors leads to errors in the remaining capacity calculation. We have observed variations in the estimated capacity with periods of 1-2 hours. These variations are challenging because this is the bandwidth where we want the coordinator to operate; hence these variations can not be reduced by filter. The variations in the estimated capacities were usually due to model errors especially from the unit feed. A systematic treatment of the inferential models (estimators of product quality) and models in the local MPC applications is necessary to obtain satisfactory performance of the coordinator MPC.

The operators are familiar with the MPC interface after several years of implementing the local MPC applications at the plant and operating those. This is a clear advantage because the coordinator MPC has the same interface. Close dialog with both operator personnel and operator leader is crucial.

The coordinator MPC uses the feed valves (throughput rate) in closed-loop. This directly involves the operator leader because the leader receives guidelines from the booking responsible and operators of the gas pipeline network.