330b Coordinator Mpc for Maximization of Plant Throughput

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In many cases the prices and market conditions are such that real-time optimization of the plant is the same as maximizing plant throughput. I this case optimal operation is the same as maintaining maximum flow through the bottleneck of the plant. Any back-off at the bottleneck is the same as lost production that can never be recovered.

One solution to this problem is to use real-time optimization (RTO) and identify the bottleneck based on a detailed steady-state model of the plant. However, the formulation of such a model is expensive and time consuming and the on-line solution is difficult.

Actually, a nonlinear model is not necessary in this simple case because the objective is to identify the active "bottleneck" constraint. Therefore, a simpler solution is to use a "coordinator MPC" based on a linear model with constraints. Here there are two possibilities. The first is to use a coordinator MPC that duplicates all the models and constraints in the the local MPCs for each unit. However, a duplication is undesirable, Therefore, we suggest a more decoupled solution strategy based on using estimates of available capacity in each unit in the plant. We assume that local MPC have already been implemented on each unit, and we then add a few lines of code in each local MPC application, so that it provides an estimate of the extra capacity available in the unit. This estimate is then send as a estimate to the coordinator MPC.

More precisely, the steady state part of the local MPC on unit k provides an estimate of the remaining (extra) capacity in the unit

Rk = Jk,max - Jk

Here Jk is the current throughput (feed) in unit k. The maximum feed rate Jk,max is obtained by solving a simple LP optimization problem withing the local MPC. The constraints in the LP problem are the same as in the existing steady state solver in the local MPC - only the objective function is different.

The LP calculation in each local MPC returns to the coordinator the remaining capacity Rk in each unit. The optimization at the coordinator level is to maximize the weighted overall feed rate:

max J subject to Rk > 0; Rk = G * MV

where J is a weighted sum of the feed rates to the plant. The manipulated variables (MVs) at the coordinator level are typically the external feed rates and crossovers in the plant. G represents the steady-state and dynamic influence from each MV to Rk.

The coordinator task is to maximize the plant throughput within feasible operation. Maximizing flow rate can be realized with a standard MPC quadratic objective function by using a total plant feed as a CV with a high (not reachable) set point with lower priority than the capacity constraints. Other solutions are possible, but this "trick" was found to work well in practice.

Implementation

The coordinator MPC approach has been tested with good results on a detailed dynamic simulator of the Karsto Gas Processing plant on the south western coast of Norway, The model includes two fractionation trains, T-100 and T-300, which both have a deethanizer, depropanizer, debutanizer and a butane splitter. In addition T-300 has two stabilizers in parallel. There are two separate train feeds, a liquid stream from a dew point control unit (DPCU) that is divided between the two trains, and a crossover. The five streams are MVs in the coordinator MPC. The local MPCs and the coordinator are implemented in Statoils SEPTIC MPC software. Work is underway to implement the coordinator MPC on the real plant.

Discussion

1. To estimate the remaining capacity in the distillation columns, we have so far used the column pressure drop as an indicator. However, preliminary testings shows that this is not a always a good measure. Thus a more detailed column capacity model may be need in some cases.

2) All flow changes (MV and Rk) should be expressed as relative (%) changes. In this case, most of the the steady-state gains in the model G are equal to 1, independent of overall feed composition.

3) From an economic point of view, the flow should always be maximum at the bottleneck. However, the flow through the bottleneck unit may not be available as a dynamic MV, and in this case a dynamic back-off (and loss in production) is unavoidable. The magnitude of the back-off depends on the expected size of the disturbances and the dynamic performance of the underlying control system. The use of back-off reduces the value of J, but makes the coordinator more robust. To reduce the back-off, one may introduce the level setpoints (e.g. condenser and reboiler levels) in all units as additional (dynamic) MVs for the coordinator MPC. The economic benefit of this strategy may be substantial, and would not be possible to realize with a standard steady-state real-time optimizer (RTO).