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Control Structure Selection with Regard to Stationary and Dynamic Performance with Application to A Ternary Distillation Column

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Abstract
In previous work, we extended the analysis of control structures based upon full process models to include the dynamic performance in a consistent manner. In our approach, NMPC-controllers are assumed to avoid the problem of comparison between control structures where the dynamic performance depends on the type and the parameterization of the controllers that are used. The weights of these controllers are optimized for each structure to yield an optimal economic performance for the disturbance scenarios considered. In order to avoid basing the decision on the worst-case disturbances only, the probability of the occurrence of the disturbances is taken into account, considering also small disturbances which happen more frequently. Thus the result is a good approximation of the performance that is attained for each structure in reality and the structures are compared on equal grounds. The approach is demonstrated for a ternary distillation problem.

Keywords: Control structure selection, economic performance, distillation control

1. Introduction
Control structure selection deals with the choice of manipulated and measured variables in feedback control. The performance of the closed-loop system is more affected by the choice of control structure than by the choice of control algorithms. Most papers from the control community deal only with the resulting tracking and regulation performance rather than the resulting performance of the plant in the presence of disturbances and parameter variations.
Several important aspects of the control of chemical processes from the point of view of plant performance were discussed by Morari et al. (1980). Morari's idea was followed by the so-called "self-optimizing" control introduced by Skogestad (2000), meaning that in the presence of disturbances, a well chosen control structure could be able to maintain the process at a close-to-optimal operating point. A step wise approach based on rigorous stationary analysis was proposed. Engell et al. (2005) refined this approach, using objective criteria and optimization to replace informed judgements taking into account the effect of measurement errors. Static performance indicators were used to assess the promising control structures. Pham and Engell (2009) extended the work to consider also the dynamic performance for time-varying disturbances to better judge the performance of the controlled plants. In this paper, results for a new case study of a ternary distillation column are shown for which realistic static and dynamic disturbances are delivered. The control structure selection is performed based upon a full nonlinear dynamic plant model.

2. Control Structure Selection Procedure

The purpose of automatic feedback control from a process engineering point of view is to establish the close-to-optimal process operation in the presence of disturbances and plant model mismatch. The effect of feedback control on the profit function J is the difference between the profit from keeping the manipulated variables at the nominal values with no disturbances affecting the plant and the profit from regulating the manipulated variables by the controller with disturbances affecting the plant. This can be expressed as (Engell et al., 2005, Engell, 2007):

$$\Delta J = \left(J(\underline{u}_{nom},0) - J(\underline{u}_{nom},\underline{d}_i)\right) + \left(J(\underline{u}_{nom},\underline{d}_i) - J(\underline{u}_{opt},\underline{d}_i)\right) + \left(J(\underline{u}_{opt},\underline{d}_i) - J(\underline{u}_{con},\underline{d}_i)\right).$$

The first term is the loss if disturbances occur and the manipulated variables are fixed at their nominal values. The second term is the effect of the optimal adaptation of the manipulated variables in the presence of disturbances. The third term is the difference between the optimal adaptation and the one realized by the chosen feedback control structure in the presence of disturbances. The overall performance of a control structure should be evaluated by the expected loss of profit (Engell et al., 2005):

$$\Delta J = \int_{-d_{1,\max}}^{d_{1,\max}} \cdots \int_{-d_{n,\max}}^{d_{n,\max}} w(\underline{d})(J(\underline{u}_{nom},\underline{d}_i) - J(\underline{u}_{con},\underline{d}_i))dd_1 \ldots dd_n,$$

where $w(\underline{d})$ denotes the probability of the occurrence of the disturbance $\underline{d}$ which is usually approximated by the weighted sum over a set of disturbance scenarios. Here we assume that the process is affected by two kinds of disturbances: slow and fast varying disturbances. The optimization of the steady-state performance is used to treat the former ones, analyzing the worst case performance of the regulatory control when keeping the controlled variables within a range around the set-points defined by the measurement errors (Engell et al., 2005). The latter ones are incorporated in a consistent manner introducing optimization-based control using nonlinear dynamic models to avoid the problem of a comparison of control structures where the dynamic performance depends on the type and the parameterization of the controllers that are used. The weights used in the cost functions of the NMPC-controllers are optimized to get an optimal economic performance. In this paper, realistic disturbance cases are considered by defining both the size and the frequency content of the disturbances. Thus the result is a good approximation of the real performance of each structure and the best structure is found by evaluating all promising candidates. The proposed control structure selection procedure consists of 6 steps:

2.1. Define the optimization problem:

The available degrees of freedom of the process are determined and the manipulated variables are chosen. A profit function J to be maximized and the constraints that need to be fulfilled during the operation are defined:

$$\max_{u} J(\underline{x},\underline{u},\underline{d}_i)$$

$$s.t \quad \underline{f}(\underline{x},\underline{u},\underline{d}) = 0 \text{ (plant model)}$$

$$\underline{h}(\underline{x},\underline{u}) \leq 0 \text{ (constraints).}$$

The output mapping is given by: $\underline{y} = \underline{m}(\underline{x})$.

2.2. Choose the disturbances:

Two types of disturbances are assumed: measurement errors and external disturbances. While the former ones can be found in the instrument data-sheets, the latter are caused by errors in the assumed model, disturbances, etc.

2.3. Pre-selection of the control structures

The number of possible control structures is a function of the number of available measurements and controlled variables which grows quickly with the number of measurements. Hence for large problems, unpromising structures should be pre-screened before. Many indices can be utilized, e.g. RHP zeros, generalized non negative RGA, etc.

2.4. Selection of the set-points for regulatory control

The optimal set-points are determined by solving:

$$\begin{aligned} &\max_{\underline{y}_{set}} \sum_{i=1}^{n} J(\underline{x}, \underline{u}_i, \underline{d}_i) \\ &s.t.: \forall \underline{d}_i : \\ &\quad \underline{\dot{x}} = f(\underline{x}, \underline{u}_i, \underline{d}_i) = 0 \\ &\quad \underline{h}(\underline{x}, \underline{u}) \le 0, \underline{y}_{set} = \underline{m}(\underline{x}). \end{aligned}$$

The idea is to find set-points that satisfy the constraints for all disturbances. The optimization problem can be infeasible, meaning that for the given constraints and disturbances there is no common set-point which can be attained.

2.5. Quantitative evaluation of the benefits of the control structures with constant disturbances

For all disturbances $\underline{d}_i$, the following optimization problem is solved to obtain the worst case control performance for regulation of the controlled variables to values in the range around the nominal set-point $\underline{y}_{set}$ defined by the measurement errors:

$$\begin{aligned} &\min J(\underline{x}, \underline{u}_i, \underline{d}_i), \\ &s.t.: \underline{\dot{x}} = f(\underline{x}, \underline{u}_i, \underline{d}_i) = 0, \\ &\quad \underline{h}(\underline{x}, \underline{u}) \le 0, \underline{y} = \underline{m}(\underline{x}), \\ &\quad \underline{y}_{set} - \underline{e}_{sensor} \le \underline{y} \le \underline{y}_{set} + \underline{e}_{sensor}. \end{aligned}$$

If the value of the maximum loss is large, it means that in the presence of the measurement errors, the corresponding control structure is not able to ensure a good stationary performance and should be excluded.

2.6. Quantitative evaluation of the benefits of the control structures with dynamic disturbances

In this step, the dynamic performance which can be achieved if disturbances occur and the controlled variables are kept close to the set-points as possible is computed. This is done by employing a simulation of nonlinear model predictive control for tracking the set-points. The objective function is defined by:

$$\begin{aligned} &P(t_k) = \min_{\mathbf{u}} \left(\int_{t=t_k}^{t_k+H_P} \left(\left\| y(t) - y_{set} \right\|_P + \left\| \Delta u \right\|_Q \right) dt \right) \\ &s.t.: \quad \dot{x}(t) = f(x(t), u(t), d(t)) \qquad (P1) \\ &\quad \underline{h}(\underline{x}, \underline{u}) \le 0 \\ &\quad y(t) = m(x(t)). \end{aligned}$$

where $\|.\|_X$ denotes the norm defined as: $\|u\|_X = u^T X u$, X is a positive semi-definite matrix. The controlled variables are steered toward the set-points while the change of the manipulated variables should be minimized which is a requirement in reality. *P* and *Q* are degrees of freedom and are chosen such that the economic profit function J is maximized by an upper layer optimization:

$$\max_{P,Q} \int_{t=0}^{t=t_{end}} \sum_{i=1}^{n} J(\underline{x}, \underline{u}_i, \underline{d}_i) dt$$

$$s.t.: P, Q \succ 0.$$

The results are compared and the structure which yields the best performance is chosen.

3. Case study

The methodology described above is applied to a ternary distillation column (Skogestad, 2010) shown in Fig. 1. The column consists of 41 stages including the condenser and reboiler and is used to separate the mixture of methanol, ethanol and 1-propanol. The non-ideal vapor liquid equilibrium (VLE) is modeled by Wilson equation. The liquid hold up is modeled by Francis Weir formula. The system is described by a large nonlinear DAE system with 160 dynamic state variables. The reboiler and the condenser level are assumed to be perfectly controlled using the distillate and bottom flow rate. The reflux ratio and the boiler heat duty are left as the degrees of freedom which are chosen as the manipulated variables. We assume that methanol is the desired product in the distillate. The profit function is chosen as:

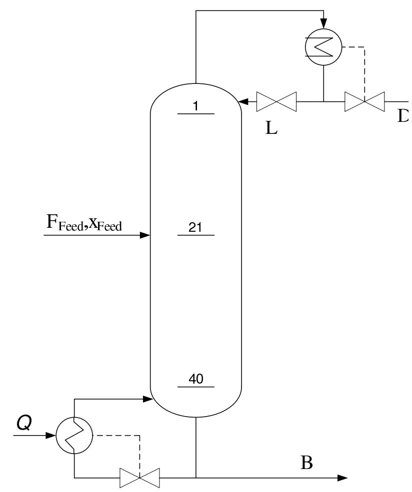

Fig. 1. Ternary distillation column

$$J = c_{Methanol} H(x_{Methanol} - 0.99)\dot{n}_{Methanol} - c_{heat}\, heatinput - c_{Feed} F.$$

H is the Heaviside step function which implies that the purity of distillate product should satisfy the requirement. To avoid the numerical problem when working with discontinuous function for optimization solver, H is approximated by a logistic function: $H(x) = 0.5 + 0.5 tanh(kx)$ with k chosen to be 100.

The possible controlled variables are the temperatures on 41 trays as well as *LT*, *VB*, *LT/D*, *LT/F*, *D/F*, *VB/B*, *B/F*, *VB/F* and the concentration of methanol in the distillate, where *LT* denotes reflux flow, *VB* heat duty, *D* and *B* distillate and bottom flow rate, *F*: feed flow rate. In total there are 50 possible controlled variables.

Structure	Profit
(T_{33}, *VB/B*)	40.1357
(*LT*, x_D)	39.8565
(T_{34}, *VB/B*)	40.1426
(*LT/F*, x_D)	40.0467
(T_{34}, *VB/F*)	40.2241

Tab. 1. Performance indices of structures resulting from a simulation of optimized NMPC controllers

For the conventional square feedback control system, with 2 manipulated variables and 50 possible controlled variables, there are: $C_{50}^{2} = 1225$ possible control structures. The disturbances are chosen as a change in the feed flow rate of ±15%, a change in the feed concentration of ±0.1 for both steady state and dynamic cases. 40% of the time the disturbances are assumed to be at the nominal value, 40% at the value of 1/3 of the worst value and 20% at the worst case value. In the dynamic case, they are simulated as step changes. Additionally uniformly distributed random noise with a magnitude of 1/6 of the maximum value is added to the disturbances. The sensor error is 0.5 K for the temperature sensor, 0.001 for the concentration measurement and 5% for others.

The nominal profit is 40.30. A pre-screening was performed based upon generalized non-square RGA and RHP zeros removing 675 unpromising structures leaving 550 structures. In step 4, common setpoints could be found for 478 structures, 72 structures are excluded from further consideration. In step 5, where the worst case analysis is performed, 11 structures that yield the smallest losses are selected for the final step. In this step, first a linear MPC simulation was employed with the nonlinear rigorous process model. From the result of the linear MPC simulation, 5 structures with best performance were chosen. For the optimization of the weights of the MPC controllers, a gradient-free optimization procedure was used. The performance indices of several structures are given in Tab. 1. It can be observed that structure (T_{34},*VB/F*) which controls the temperature of tray 34 and the ratio *VB/F* yields the best performance. The result of this structure with disturbances in the feed flow rate and the feed composition are given in Fig. 2. It is interesting that a measurement of the top composition gives no advantage here.

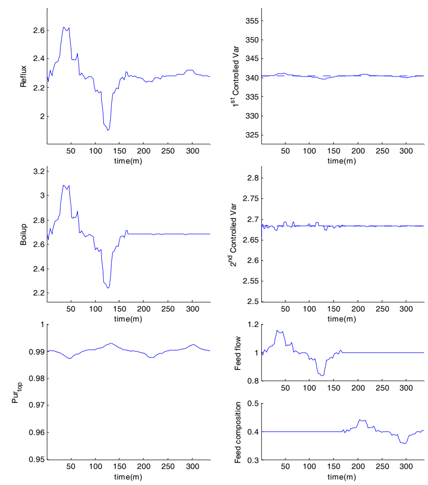

Fig. 2. NMPC simulation of (T_{34},*VB/F*) control structure with disturbance in the feed flow rate and composition

4. Conclusion

A methodology for control structure selection was presented with the aim of optimizing the plant performance taking into account the presence of both steady-state and dynamic disturbances. The method was applied successfully to the example of a ternary distillation column.

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