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Simple Rules for Economic Plantwide Control

Vladimiros Minasidis[a], Sigurd Skogestad[a*] and Nitin Kaistha[b]

[a]*Department of Chemical Engineering, Norwegian University of Science and Technology, 7491 Trondheim, Norway*
[b]*Department of Chemical Engineering, Indian Institute of Technology, 208016 Kanpur, India*
**skoge@ntnu.no*

Abstract

In this work, we consider the systematic economic plantwide control design procedure proposed by Skogestad (2004) and from this we derive practical rules that can be used to devise close-to-optimal control structures based on engineering insight. We attempt to present these rules in an easy-to-understand fashion and exemplify them on a simple but quite famous reactor-separator-recycle plant case study. We successfully demonstrate that while Skogestad's procedure requires an optimization of the plant model under various disturbances in order to be fully utilized, using its practical rules, combined with a good engineering insight of the process, can facilitate the design of a close-to-optimal control structure by suggesting what should be controlled and how.

Keywords: Self-Optimizing Control, Economic Plantwide Control, Advanced Process Control

1. Introduction

Designing a control structure for an entire chemical plant, that ensures safe and stable economic operation, is known as economic plantwide control. Energy integration, smaller in-process inventories and material recycles, all contribute to an increase of operational complexities and thus a plantwide perspective becomes crucial when designing control systems for such plants. Many different plantwide control methodologies have been proposed during the last decades, nonetheless, they can all be classified based on their approach. Vasudevan and Rangaiah (2011) suggest the following classification: mathematical, optimization-based, heuristic and mixed approaches. The first two approaches lead in many cases to an optimal control structure design but often require an optimization of rigorous process models in order to be fully utilized. Often, the entire plant model is not available and it is too time consuming to develop one. More seriously, the resulting mathematical and optimization problems are often difficult to formulate and even more difficult to solve. The alternative in such cases, is to attempt to apply the practical rules derived from heuristic/mixed procedures, that require only a good engineering insight in order to design a close-to-optimal control structure. In addition, these practical guidelines are usually easier to understand, thus easier to implement and support without requiring high expertise (Konda, 2005).

Many such practical rules that either focus on a complete plantwide control structure design or its specific parts, have been proposed in scientific literature. Just to mention a few, starting with the pioneering work of Page S. Buckley (1964), are: (Price et al., 1994), (Luyben et al., 1997), (Skogestad, 2004), (Aske et al., 2007). It should be noted that, Skogestad (2004) procedure is inspired by Luyben et al. (1997) procedure, but it clearly distinguishes between economic control for economic performance and regulatory control for stability and robustness and in order to emphasize this distinction, it is referred to as Economic Plantwide Control. For a more detailed overview of plantwide control design methodologies we refer the reader to (Larsson and Skogestad, 2000) and (Vasudevan and Rangaiah, 2012).

When examining the suggested practical guidelines, we often find that the assumptions behind them are not stated clearly, leading to the likelihood of non-optimal control strategies. For example, the most famous of Luybens laws *“Fix a Flow in Every Recycle Loop”*, when applied to certain systems, can lead to poor economic performance and a small feasible range. The correct interpretation could be that the flow should only be fixed on a fast time scale. If it is indeed truly fixed, then it should involve moving the throughput manipulator to this location. This motivates this work’s attempt to present and justify the practical rules derived from Skogestad’s procedure in an easy-to-understand fashion .

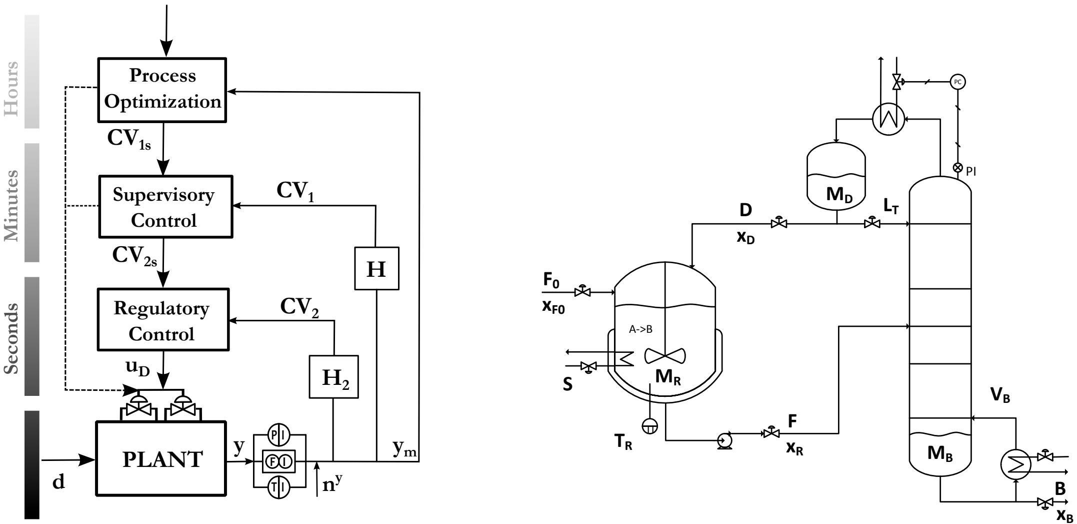
In this work, we briefly present the systematic plantwide control design procedure proposed by Skogestad (2004) in a simple Q&A form. We present its practical guidelines, with emphasis on those rules that can be used based on engineering insight and propose an additional rule, that is stated as follows: ”*Never try to control a cost function*”. We apply the presented rules and demonstrate their effectiveness on a simple but quite famous reactor-separator-recycle case study (T. Larsson et al., 2003). We conclude that, while Skogestad’s procedure requires an optimization of the plant model under various disturbances in order to be fully utilized, using its practical rules, combined with a good engineering insight of the process, can facilitate the design of a close-to-optimal control structure by suggesting what should be controlled and how.

2. Economic Plantwide Control

2.1. Preliminaries

The definitions and terms presented here are based on terminology used in (Larsson and Skogestad, 2000) and (Skogestad, 2012), unless otherwise noted. A typical hierarchical decomposition of the plantwide control structure is depicted in Figure 1a. The lowest layer, the regulatory (or stabilizing) layer, consists usually of single loop (decentralized) PID controllers and operates in a time scale of seconds. Next, the supervisory (or advanced, or economic) layer, which provides setpoints to the regulatory layer, may consist of a multivariable controller (MPC) or a mix of various decentralized ”advanced” controllers, including PID controllers selectors, feedforward controllers and so on. We define some of the terms extensively used in this work, as follows:
Controlled variable (CV) is a variable with a given setpoint.
Manupulated variable (MV) is a degree of freedom used by the controller to control a CV.
Pairing is a selection of a MV to be used to control a specific CV.
Control structure design deals with the structural decisions of identification/selection of the following sets: controlled variables $\{\boldsymbol{CV}_1\boldsymbol{CV}_2\}$, physical manipulated variables $\{\boldsymbol{u_D}\}$, measurements $\{\boldsymbol{y_m}\}$, control pairings $\{\boldsymbol{H},\boldsymbol{H_2}\}$ and controller algorithms for the layers depicted in Figure 1a.
Plantwide control is a control philosophy of the overall plant with emphasis on the structural decisions and is used interchangeably with the term “control structure design”.
Self-optimizing variables are controlled variables for which the optimal setpoints $\boldsymbol{CV}_{1s}$ can be kept relatively unchanged, i.e. their optimal values are insensitive to disturbances, at least on a fast time scale.
Optimal control structure is a control structure that achieves economically optimal -in addition to safe and stable- plant operation.
Rule in this work means a practical guideline, and both terms are used interchangeably.

Before describing the procedure for economic plantwide control it should be noted that, in general, it is simple to define the economic cost function J for a steady-state optimal operation of chemical processes. Since the plant has already been built and the operators are already at the site, there are no costs related to capital costs or other fixed costs. The cost function can be expressed as



(a) Hierarchical time-scale decomposition of the control structure.

(b) Reactor-separator-recycle process flow diagram

$J = p_F F + p_Q Q - p_P P$, where: p is the corresponding price, $p_F F$ is the sum of the costs of all the feed streams, $p_Q Q$ is the cost of all utilities (including energy) and $p_P P$ is the sum of the values of all products. In this context, we can define two main operational modes: **Mode 1** (nominal operation) where the feeds are given, so the term $p_F F$ is fixed. Furthermore, because of given product specifications, this may imply that the product rates P are also fixed. So, to minimize plant operation cost, we want to minimize the utility costs $p_Q Q$. This mode often has an unconstrained optimum, since there is a tradeoff between using too much or too little energy. This is the case for our simple example in Mode 1, where the optimal operation is the same as minimizing $J = V_B$, since the energy (boilup) to the column is the only utility; **Mode 2** (maximum throughput) that has high product prices and low energy prices, so the optimal operation corresponds to maximizing the product rate P. In general, we have more active constraints in Mode 2 than in Mode 1.

2.2. Economic Plantwide Control description

Any methodology that aims to facilitate the design of an optimal control structure, based on the hierarchical decomposition depicted on Figure 1a, should, independent of the approach, at least consider the following structural decisions:

1. **Decision 1:** Select primary controlled variables $\boldsymbol{CV}_1$ for the supervisory control layer or select $\boldsymbol{H}$. The setpoints $\boldsymbol{CV}_{1s}$ link the process optimization with supervisory control layer.
2. **Decision 2:** Select secondary controlled variables $\boldsymbol{CV}_2$ for the regulatory control layer or select $\boldsymbol{H_2}$. The setpoints $\boldsymbol{CV}_{2s}$ link the supervisory and regulatory control layers.
3. **Decision 3:** Locate the throughput manipulator (TPM) location. This is an important step, since it links the top-down and the bottom-up parts of the economic plantwide control.
4. **Decision 4:** Select pairings for the stabilizing layer controlled variables $[\boldsymbol{CV}_2 \leftrightarrow \boldsymbol{u}_D]$

Furthermore, the operational goals should be defined clearly and if possible separated into: i) economic objective ii) stabilization/regulation objectives. One reason for the separation is, that it is very difficult to measure them in the same units, for example, how much is a gain margin increase from 2 to 3 worth in dollars?

Skogestad's procedure clearly distinguishes between the economic control and regulatory control and decomposes the structural decisions, into two parts: the ***Top-down*** part, which attempts to find a slow-time-scale supervisory control structure that achieves a close-to-optimal economic operation and the ***Bottom - up*** part that aims to design a fast-time-scale regulatory control structure that is stable and robust and works under all the conditions imposed by the economic supervisory layer. A brief description of the procedure is presented in Table 1 in the form of typical structural decisions (Q) and how it attempts to address those in a systematic manner (A).

Note that, the structural decisions related to the parings (steps *S5, S5, S6*) could in theory be avoided, if we used a single multivariable controller (e.g. MPC). Furthermore, the choice of economic controlled variables $\boldsymbol{CV}_1$ (related to steps S3, S7) could also, in theory, be avoided, if we used a single dynamic real-time optimizer, which combined optimization and control. However, designing, maintaining and tuning such a controller can be too difficult and too expensive for a large plant.

Top-down part	*(mainly steady-state)*
Step S1:	**Define the operational objectives (economics) and constraints.**
Q:	*What are the overall economic objectives for operation?*
A:	*Identify a scalar cost function J and the operational constraints. Identify the dynamic degrees of freedom (DOFs)* $\boldsymbol{u_D}$*. Determine how many steady state DOFs* n_{SS} *are available. Identify the expected economic disturbances* $\boldsymbol{d}$ *and and their range.*
Step S2:	**Determine the steady-state optimal operation**
Q:	*What are the operational conditions for an optimal steady state operation?*
A:	*Determine the optimal operation and, in particular, the optimal constraints regions (regions in the disturbance space with the same active constraints) for the expected disturbance* $\boldsymbol{d}$ *ranges*
Step S3:	**Select primary (economic) controlled variables** (Decision 1)
Q:	*What should be controlled to maintain a close-to-optimal economic operation in-spite of disturbances ?*
A:	*Identify the candidate measurements* $\boldsymbol{y_m}$ *and select the primary controlled variables such as* $\boldsymbol{y_m} \supset \boldsymbol{CV}_1 = \boldsymbol{H}\boldsymbol{y_m}$*, where* $\boldsymbol{H}$ *is a selection matrix. The set* $\boldsymbol{CV_1}$ *should include active constraints* $\boldsymbol{A_C}$ *and if* $\#\boldsymbol{A_C} < n_{SS}$*, a* n_{SS} *number of "self-optimizing variables"* $\boldsymbol{CV}_{soc}$ *should be identified and included in* $\boldsymbol{CV}_1$*.*
Step S4:	**Select the location of throughput manipulator (TPM)** (Decision 3)
Q:	*Where should the plant's "gas pedal" be located ?*
A:	*The location of the TPM is a dynamic issue but it may have significant economic implications and, therefore, is located in the Top-down part. For TPM location see Rules 4 and 5.*
Bottom-up part	*(mainly dynamic)*
Step S5:	**Select the control structure for the Regulatory Control layer.** (Decisions 2 and 4)
Q:	*What variables should be controlled to stabilize the operation and how ?*
A:	*Select all the "drifting" process variables* $\boldsymbol{CV}_2 = \boldsymbol{H_2 y}$ *that need to be controlled to ensure safe and stable plant operation e.g. inventories, pressures, temperatures. The inventory control structure should be designed starting from the TPM location and following the radiating rule to achieve local consistency. Usually single loop (decentralized) PID controllers are used in this layer.*
Step S6:	**Select the control structure for the Supervisory Control layer.**
Q:	*How should the economic controlled variables* $\boldsymbol{CV}_1$ *be controlled ?*
A:	*The primary economic variables* $\boldsymbol{CV}_1$ *should be controlled at this layer, using the setpoints to the regulatory layer in addition to any unused valves. Two types of controllers are usually selected for this layer: multivariable controller (e.g. MPC) or a mix of various decentralized "advanced" controllers, including PID controllers selectors, feedforward controllers and so on.*
Step S7:	**Select the control structure for the Real-Time Optimization layer.**
Q:	*How often should the optimal setpoints for the economic controlled variables* $\boldsymbol{CV}_1$ *and their pairings be updated ?*
A:	*This layer should re-optimize the setpoints for* $\boldsymbol{CV}_1$ *and track any changes in the set of active constraints, in order to update the economic control structure, if necessary. A limiting factor for the updating frequency is the requirement that the plant has to settle, before the re-optimization problem can be solved. A successful mapping of the active constraints regions for the entire range of expected disturbances, combined with set of good self-optimizing variables, may remove the need for having this layer.*

Table 1: Economic Plantwide Control design procedure. The nomenclature used here is depicted in Figure 1a. For more details we refer the reader to (Skogestad, 2012)

3. Practical rules

Here we present a set of simple rules for economic plantwide control to facilitate a close-to-optimal control structure design in cases where the optimization of the plant model is not possible. Due to space limitations, only short justifications for the rules are given and, in addition, we attempt to exemplify the rules, when possible, using the reactor-separator-recycle process (Figure 1b). Note that, the rules may be conflicting in some cases and in such cases, human reasoning is strongly advised.

Rules for Step S3: Selection of primary (economic) controlled variables, $\boldsymbol{CV}_1$

Rule 1: *Control the active constraints.*
In general, process optimization is required to determine the active constraints, but in many cases these can be identified based on a good process knowledge and engineering insight. Here is one useful rule:

Rule 1A: The purity constraint of the valuable product is always active and should be controlled.
This follows, because we want to maximize the amount of valuable product and avoid product "give away" (Jacobsen and Skogestad, 2011). Thus, we should always control the purity of the valuable product at its specification. For "cheap" products we may want to overpurify (purity constraint may not be active) because this may reduce the loss of a more valuable component.
In other cases, we must rely on our process knowledge and engineering insight. For reactors with simple kinetics, we usually find that, the reaction and conversion rates are maximized by operating at maximum temperature and maximum volume (liquid phase reactor). For gas phase reactor, high pressure may increase the reaction rate, but this must be balanced against the compression costs.

Rule 2: *(for remaining unconstrained steady-state degrees of freedom, if any): Control the "self-optimizing" variables.*
This choice is usually not obvious, as there may be several alternatives, so this rule is in itself not very helpful. The ideal self-optimizing variable, at least, if it can be measured accurately, is the gradient of the cost function J_u, which should be zero for any disturbance. Unfortunately, it is rarely possible to measure this variable directly and the "self-optimizing" variable may be viewed as an estimate of the gradient J_u. The two main properties of a good "self-optimizing" variable are: (1) its optimal value is insensitive to disturbances (such that $\boldsymbol{F} = \frac{\Delta \boldsymbol{CV}_{1,opt}}{\Delta \boldsymbol{d}}$ is small) and (2) it is sensitive to the plant inputs (so the process scaled gain $\boldsymbol{G} = \frac{\Delta \boldsymbol{CV}_1}{\Delta \boldsymbol{u}}$ is large). The following rule shows how to combine the two desired properties:
Rule 2A: Select the set $\boldsymbol{CV}_1$ such that the ratio $\boldsymbol{G}^{-1}\boldsymbol{F}$ is minimized.
This rule is often called the "maximum scaled gain rule". For proof, see (Halvorsen et al., 2003).

Rule 3: *(for remaining unconstrained steady-state degrees of freedom, if any): Never try to control the cost function J (or any other variable that reaches a maximum or minimum at the optimal operating point).*
First, the cost function J has no sensitivity to the plant inputs at the optimal point and so $\boldsymbol{G} = 0$, which violates Rule 2A. Second, if we specify J lower than its optimal value, then clearly, the operation will be infeasible, see Figure 2a. From Figure 2a, we can also see that specifying J higher than its optimal value is problematic, as we have multiplicity of solutions. As mentioned above, rather controlling the cost J, we should control its gradient, J_u.

Rules for Step S4: Location of throughput manipulator (TPM)

Rule 4: *Locate the TPM close to the process bottleneck.*

The justification for this rule is to take advantage of the large economic benefits of maximizing production in times when product prices are high relative to feed and energy costs (**Mode 2**). To maximize the production rate, one needs to achieve tight control of the active constraints, in particular, of the bottleneck, which is defined as the last constraint to become active when increasing the throughput rate. For more details, we refer the reader to (Jagtap et al., 2013).

Rule 5: *(for processes with recycle) Locate the TPM inside the recycle loop.*
The point is to avoid "overfeeding" the recycle loop which may easily occur if we operate close to the throughput where "snowballing" in the recycle loop occurs. This is a restatement of Luyben's rule *"Fix a Flow in Every Recycle Loop"* (Luyben et al., 1997). From this perspective, snowballing can be thought of as the dynamic consequence of operating close to a bottleneck which is within a recycle system. In many cases, the process bottleneck is located inside the recycle loop and Rules 4 and 5 give the same result.

Rules for Step S5: Structure of regulatory control layer.

Rule 6: *Arrange the inventory control loops (for level, pressures, etc.) around the TPM location according to the radiation rule.*
The radiation rule (Price et al., 1994), says that, the inventory loops upstream of the TPM location must be arranged opposite of flow direction. For flow downstream of TPM location it must be arranged in the same direction. This ensures "local consistency" i.e. all inventories are controlled by their local in or outflows.

Rule 7: *Select "sensitive/drifting" variables as controlled variables $\boldsymbol{CV}_2$ for regulatory control.*
This will generally include inventories (levels and pressures), plus certain other drifting (integrating) variables, for example, a reactor temperature or a sensitive temperature in a distillation column. This ensures "stable operation, as seen from an operator's point of view.

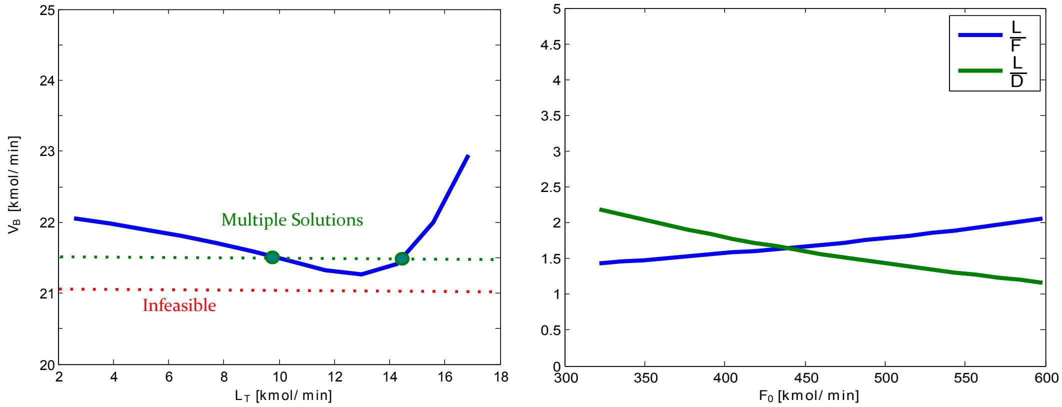
Some component inventories may also need to be controlled, especially for recycle systems. For example, according to *"Down's drill"* one must make sure that all component inventories are "self-regulated" by flows out of the system or by removal by reactions, otherwise their composition may need to be controlled (Luyben, 1999).

Rule 8: *Economically important active constraints, a subset of $\boldsymbol{A_C}$, should be selected as controlled variables $\boldsymbol{CV}_2$ in the regulatory layer.*
Economic variables $\boldsymbol{CV}_1$ are generally controlled in the supervisory layer. Moving them to the faster regulatory layer may ensure tighter control with a smaller backoff. The backoff is the difference between the actual average value (setpoint) and the optimal value (constraint).

Rule 9: *"Pair-close" rule: The pairings should be selected such that, effective delays and loop interactions are minimal.*

Rule 10: *: Avoid using MVs that may optimally saturate (at steady state) to control CVs in $\boldsymbol{CV}_2$.*
The reason is that we want to avoid re-configuring the regulatory control layer. To follow this rule, one needs to consider also other regions of operation than the nominal, for example, operating at maximum capacity (**Mode 2**) where we usually have more active constraints.

Rules for Step S6: Structure of supervisory control layer.



(a) Steady state boilup flow rates V_B w.r.t. reflux flow rate L_T changes while all the active constraints are kept constant

(b) Optimal steady state of suggested $\boldsymbol{CV}_{soc}$ candidates $\{\frac{L_T}{F}, \frac{L_T}{D}\}$ w.r.t. feed flow F_0 changes ± 30 percent while all the active constraints are kept constant

Rule 11: *MVs that may optimally saturate (at steady state) should be paired with the subset of* $\boldsymbol{CV}_1$ *that may be given up.*
This rule applies for cases when we use decentralized control in the supervisory layer and we want to avoid reconfiguration of loops. The rule follows because when a MV optimally saturates, then, there will be one less degree of freedom, so there will be a CV_1 which may be given up without any economic loss. The rule should be considered together with rule 10.

4. Case study

4.1. Process description

In this section we consider a quite famous example of a chemical plant that consists of a reactor, a separator and a recycle stream. For this case study, all the operational parameters have been taken from (Larsson et al., 2003). The process flow diagram of the process and all the required nomenclature are depicted in Figure 1b. An irreversible first order elementary reaction $A \rightarrow B$ takes place in continuous stirred tank reactor (CSTR). The effluent of the reactor, which consist of a solution of A and B is sent to fractional distillation column, where the products are separated and the distillate stream (mostly component A), is recycled back to the CSTR.

4.2. Applying the practical rules

S1: Assuming that the plant, depending on market conditions, will be switching between Mode 1 and 2, the objective for Mode 1 is to minimize energy consumption $J = V_B$ and for Mode 2, to maximize the production $J = -B$. The operational constraints are: $M_R \leq 2800\,kmol$, $x_B \leq 0.0105$, $V_B \leq 1500\,mol/min$, $T_R \leq 600K$. Assuming that the pressure in the column is kept constant by the condenser flow, the process has 7 DOFs: $\boldsymbol{u_D} = [L_T, V_B, D, B, F, F_0, S]$. Only two DOFs from $[L_T, V_B, D, B]$ can be set independently at a steady state, which means $n_{SS} = 5$ steady state DOFs.
S2: We assume, based on our engineering insight and *Rule 1A*, that the following constraints will be most probably be active $\boldsymbol{A_C} = [M_R, x_B, T_R]$.
S3: In addition to a DOF consumed by the TPM, all $\boldsymbol{A_C}$ need to be controlled, thus 1 DOF is left that can be used to improve the operational economics. Some candidates for $\boldsymbol{CV}_{soc}$ are $\{\frac{L_T}{F}, \frac{L_T}{D}\}$.
S4: The last constraint that becomes active in Mode 2 is V_B. In addition, there are two recycle loops in the process (one is the column itself) and V_B is inside both of the loops. According to the *Rule 4 and 5* the best candidate for TPM location is V_B.

S5: The selection of V_B as the TPM leads to the following pairings $(B \rightarrow M_B), (D \rightarrow M_D), (F_0 \rightarrow M_R), (S \rightarrow T_R)$ according to Rules 6 or 9. In order to get a tighter control of x_B and to stabilize the composition profile in the column, we can cascade the x_B control with some sensitive column temperature T_C such that $(F \rightarrow T_C \rightarrow x_B)$ according to Rule 7 and 8. The suggested pairings may remove the need to re-configure the structure when switching from one mode to another.
S6: The only available DOF for this layer is L_T, so it needs to be used to control one of the suggested $\boldsymbol{CV}_{soc} = \{\frac{L_T}{F}, \frac{L_T}{D}\}$. A preferable controller in this case would be a single loop PID.
S7: Assuming that the assumptions for the active constraints set and the "self-optimizing" controlled variable are correct, the additional benefits from having this layer would be quite limited.

5. Results and Conclusions

The validity of our practical analysis is checked using the optimization results presented in (Larsson et al., 2003). The practical choice of active constraints is in agreement with Larsson's findings. The optimal sensitivity of the $\boldsymbol{CV}_{soc}$ w.r.t to disturbances is depicted in Figure 2b, which shows that the optimal values of the suggested CVs change very little with quite large disturbances thus qualifying them as good "self-optimizing" variables. The dynamic performance of the plant control structure is omitted due to space limitations.

In conclusion, despite the fact that to fully utilize the true potential of the economic plantwide control procedure a series of plant model optimizations is required, the derived practical guidelines can be used to decide on "what to control and how" in order to improve the economics of the plant operation, even if the model is not available or optimization of the plant model is difficult.

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