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5. Process Dynamics, Control and Monitoring and Identification - Plantwide Control

Sigurd Skogestad, Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), Trondheim, Norway

1. INTRODUCTION

A chemical plant may have thousands of measurements and control loops. By the term *plantwide control* it is not meant the tuning and behavior of each of these loops, but rather the *control philosophy* of the overall plant with emphasis on the *structural decisions.* In practice, the control system is usually divided into several layers, separated by time scale (see Figure 1).

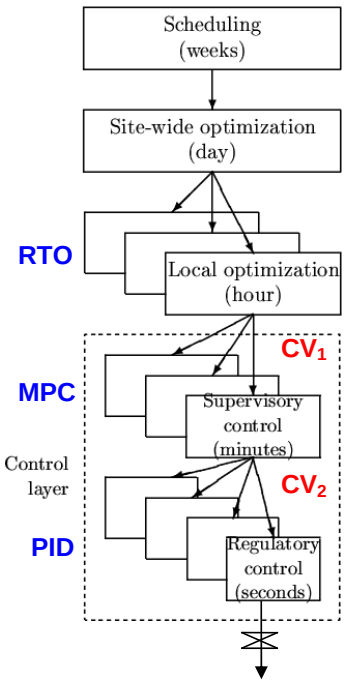

Figure 1: *Typical control hierarchy in a chemical plant.*

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The issue in plantwide control is *not* the tuning and behavior of each control loop, but rather the *control philosophy* of the overall plant with emphasis on the ***structural decisions****:*

- *Selection of controlled variables (CVs, "outputs")*
- *Selection of manipulated variables (MVs, "inputs")*
- *Selection of (extra) measurements*
- *Selection of control* ***configuration*** (structure of overall controller that interconnects the controlled, manipulated and measured variables)
- *Selection of controller type* (PID, decoupler, MPC, LQG, ratio, etc.).

Plantwide control thus involves all the decisions necessary to make a block diagram (used by control engineers) or a process & instrumentation diagram (used by process engineers) for the entire plant, but it does not involve the actual design of each controller.

In any mathematical sense, the plantwide control problem is a formidable and almost hopeless combinatorial problem involving a large number of discrete decision variables, and this is probably why the progress in the area has been relatively slow. In addition, the problem has been poorly defined in terms of its objective. Usually, in control, the objective is that the controlled variables

(CVs, outputs) should remain close to their setpoints. However, what should we control? Which CVs? The answer lies in considering the overall plant objective, which normally is to minimize the economic cost (=maximize profit) while satisfying operational constraints imposed by the equipment, market demands, product quality, safety, environment and so on. We will get back to this.

The truly optimal "plantwide controller" would be a single centralized controller which at each time step collects all the information and computes the optimal changes in the manipulated variables (MVs). Although such a single centralized solution is foreseeable on some simple processes, it seems to be safe to assume that it will never be applied to any normal-sized chemical plant. There are many reasons for this, but one important one is that in most cases one can obtain acceptable control performance with simple structures where each controller block only involves a few variables, and such control systems can be designed and tuned with much less effort, especially when it comes to the modelling and tuning effort. After all, most real plants operate well with simple control structures. So how are systems controlled in practise? The main simplification is to decompose the overall control problem into many simpler control problems. This decomposition involves two main principles

- *Decentralized (local) control.* This "horizontal decomposition" of the control layer is mainly based on separation in space, for example, by using local control of individual process units.
- *Hierarchical control.* This "vertical decomposition" is mainly based on time scale separation, and in a process one typically has the following layers (see Figure 1)
    - scheduling (weeks)
    - site-wide optimization (day)
    - local optimization (hour)

- supervisory (predictive, advanced) control (minutes)
- regulatory control (seconds)

The upper three layers in Figure 1 deal explicitly with economic optimization and are not considered in this chapter. We are concerned with the two lower *control layers* where the main objective is to track the setpoints specified by the layer above. A very important structural decision, probably more important than the controller design itself, is the choice of controlled variables (CVs) that interconnect the layers. More precisely, the decisions made by each layer (boxes in Figure 1) are sent as setpoints for the controlled variables (CVs) to the layer below. Thus, we do indirectly consider optimization because we want to select CVs that are favorable from an economic point of view.

Typically, PID controllers are used in the *regulatory control layer,* where "stabilization" of the plant is the main issue. In the *supervisory control layer*, one has traditionally used manual control or single-loop PID control, complemented by "advanced" elements such as static decouplers, feedforward elements, selectors, split-range controllers and various logic elements. However, over the last 25 years, model predictive control (MPC) has gradually taken over as a unifying tool to replace most of these elements. In the (local) *optimization layer*, the decisions are usually executed manually, although real-time optimization (RTO) is used for a few applications, especially in the refining industry.

No matter what procedure we choose to use, the following decisions must be made when designing a plantwide control strategy:

Decision 1. Select "economic" (primary) controlled variables (CV_1) for the supervisory control layer.

Decision 2. Select ”stabilizing” (secondary) controlled variables (CV_2) for the regulatory control layer.

Decision 3. Locate the throughput manipulator (TPM), that is, where to set the production rate..

Decision 4. Select pairings for the stabilizing layer, that is, pair inputs (valves) and controlled variables (CV_2). By “valves” is here meant the original dynamic manipulated variables.

Decisions 1 and 2 are illustrated in Figure 2, where the matrices H and H_2 represent a selection, or in some cases a combination, of the available measurements y.

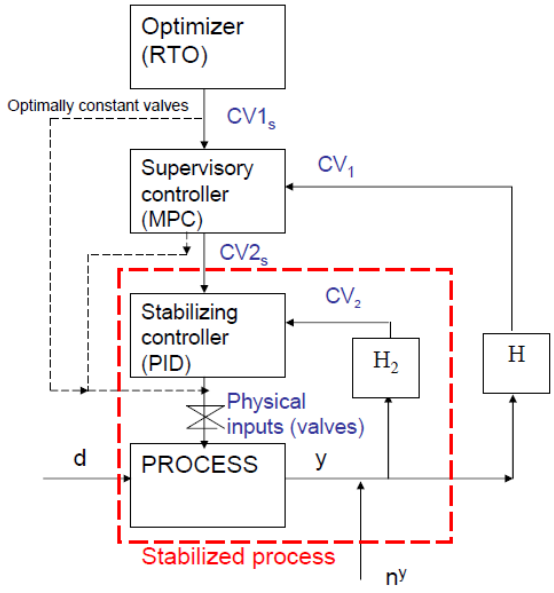

Figure 2: *Block diagram of control hierarchy illustrating the selection of controlled variables (H and H_2) for optimal operation (CV_1) and stabilization (CV_2).*

This paper deals with continuous operation of chemical processes, although many of the arguments hold also for batch processes.

2. PREVIOUS WORK

Over the years, going back to the early work of Buckley (1964) from the DuPont company, several approaches have been proposed for dealing with plantwide control issues. Nevertheless, taking into account the practical importance of the problem, the literature is relatively scarce. Larsson and Skogestad (2000) provide a good review and divide into two main approaches. First, there are the *process-oriented* (engineering or simulation-based) approaches of Buckley (1964)), Shinskey (1984), Douglas (1988), Downs, (1992), Luyben et al. (1997, 1998) and Konda et al. (2005). One problem here is the lack of a really systematic procedure and that there is little consideration of economics. Second, there is the optimization or *mathematically oriented* (academic) approaches of Narraway and Perkins (1993), Hansen et al. (1998), Kookos and Perkins (2002), Chen and McAvoy (2003) and Engell (2007). The problem here is that the resulting optimization problems are intractable for a plantwide application. Therefore, a hybrid between the two approaches is more promising; Larsson and Skogestad (2000), Zheng et al. (1999), Vasbinder and Ho (2003), Skogestad (2004), Ward et al. (2006).

The first really systematic plantwide control procedure was that of Luyben et al. (1997, 1998) which has been applied in a number of simulation studies. **Luyben's procedure** consists of the following nine steps

Step L1. *Establish control objectives*

Step L2. *Determine control degrees of freedom*

Step L3. *Establish energy management system*

Step L4. *Set the production rate* ***(Decision 3)***

Step L5. *Control product quality and handle safety, environmental and operational constraints*

Step L6. *Fix a flow in every recycle loop and control inventories*

Step L7. *Check component balances*

Step L8. *Control individual unit operations*

Step L9. *Optimize economics and improve dynamic controllability*

Note that "establish control objectives" in step L1 does not lead directly to the choice of controlled variables **(Decisions 1 and 2**). Thus, in Luyben's procedure, **Decisions 1, 2 and 4** are not explicit, but are included implicitly in most of the steps. Even though the procedure is systematic, it is still heuristic and *ad hoc* in the sense that it is not clear how the authors arrived at the steps or their order. A major weakness is that the procedure does not include economics, except as an afterthought in step L9.

In this chapter, we mainly discuss the seven-step plantwide control procedure of Skogestad (Larsson and Skogestad, 2000; Skogestad 2004). It was inspired by the Luyben procedure, but it is clearly divided into a top-down part, mainly concerned with steady-state economics, and a bottom-up part, mainly concerned with stabilization and pairing of loops. **Skogestad's procedure** consists of the following steps:

I. Top-down part (focus on steady-state optimal operation)

Step S1. *Define operational objectives (economic cost function J and constraints)*

Step S2. *Determine the optimal steady-state operation conditions*

Step S3. *Select "economic" (primary) controlled variables, CV_1* (**Decision 1**)

Step S4. *Select the location of the throughput manipulator (TPM)* (**Decision 3)**

II. Bottom-up part (focus on the control layer structure)

Step S5. *Select the structure of the regulatory (stabilizing) control layer)* **(Decisions 2 and 4)**

Step S6. *Select the structure of the supervisory control layer*

Step S7 *Select structure of (or need for) the optimization layer (RTO)*

The top-down part (steps 1-4) is mainly concerned with economics, and steady-state considerations are often sufficient. Dynamic considerations are more important for steps 4 to 6, although steady-state considerations are important also here. This means that it is important in plantwide control to involve engineers with a good steady-state understanding of the plant. A detailed analysis in step S2 and step S3 requires that one has a steady-state model available and that one performs optimizations for the given plant design ("rating mode") for various disturbances.

In the rest of this chapter, we want to consider in more detail the steps in the procedure of Skogestad and then discuss how it relates to the Luyben procedure. However, before doing this, we consider the issue of degrees of freedom, which is common to both procedures.

3. DEGREES OF FREEDOM FOR OPERATION

The issue of degrees of freedom for operation, or control degrees of freedom, is often confusing and not as simple as one would expect. First, note that we are talking about operation, so the equipment design is fixed. Second, note that the degrees of freedom change depending on where we are in the control hierarchy. This is illustrated in Figures 1 and 2, where we see that the degrees of freedom in the optimization and supervisory control layers are not the physical degrees of freedom (valves), but rather setpoints for the controlled variables in the layer below. The control degrees of freedom are often referred to as manipulated variables (MVs) or inputs. We call the physical degrees of freedom (dynamic process inputs) for "valves", because this is usually what they are in process control.

Steady-state DOFs (u). A simple approach is to first identify all the physical (*dynamic*) degrees of freedom (valves). However, because the economics usually depend mainly on the steady-state, we should subtract variables that have no or negligible effect on the economics (steady-state), such as inputs with only a dynamic effect or controlled variables (e.g. liquid levels) with no steady-state effect.

steady-state degrees of freedom (u) = #valves - #variables with no steady-state effect

For example, even though a heat exchanger may have a valve on the cooling water and in addition have bypass valves on both the hot and cold side, it usually has only one degree of freedom at steady-state, namely the amount of heat transferred, so two of these three valves only have a dynamic effect from a control point of view.

In addition, we need to exclude valves that are used to control variables with no steady-state effect (usually, liquid levels). This is illustrated in the following example.

Example. DOFs for distillation. *A simple distillation column has 6 dynamic degrees of freedom (valves): feed F, bottom product B, distillate product D, cooling, reflux L and heat input. However, two degrees of freedom (e.g., B and D) must be used to control the condenser and*

reboiler levels (M_B and M_D) which have no steady-state effect. This leaves 4 degrees of freedom at steady-state. For the common case with a given feed flow and a given column pressure, only 2 steady-state degrees of freedom remain. Thus, based on an economic analysis in step S3 we need to select 2 controlled variables (CV_1) associated with these. Typically, these will be the top and bottom composition, but not always.

4. SKOGESTAD'S PLANTWIDE CONTROL PROCEDURE

We here go through the Skogestad (2004) procedure in more detail. We consider an existing plant and assume that we have a steady-state model of the process available.

I. TOP-DOWN PART

The top-down part is mainly concerned with the plant economics, which are usually determined primarily by the steady-state behavior. Therefore, although we are concerned about control, steady-state models are usually sufficient for the top-down part.

Step S1. *Define operational objectives (cost J and constraints).*

A systematic approach to plantwide control requires that we first quantify the operational objectives in terms of a scalar cost function J [$/s] that should be minimized (or equivalently, a scalar profit function, P = -J, that should be maximized). This is usually not very difficult, and typically we have

$$J = \text{cost feed} + \text{cost utilities (energy)} - \text{value products} \quad [\$/s]$$

Note that fixed costs and capital costs are not included, because they are not affected by plant operation on the time scale we consider (around 1 hour). The goal of operation (and of control) is to minimize the cost J, subject to satisfying the operational constraints ($g \leq 0$), including safety and

environmental constraints. Typical operational constraints are minimum and maximum values on flows, pressures, temperatures and compositions. For example, all flows, pressures and compositions must be non-negative.

Step S2. ***Determine the steady-state optimal operation***

What is the optimal way of operating the process? We should answer this question before designing the control system. For example, we may find that a valve (for example a bypass) should always be closed. This valve should then not be used for (stabilizing) control unless we are willing to accept the loss implied by "backing off" from the optimal operating conditions.

To determine the steady-state optimal operation, we first need to obtain a steady-state model. Then we need to identify degrees of freedom and expected disturbances, and perform optimizations for the expected disturbances:

(a) **Identify steady-state degrees of freedom (u).** To optimize the process, we first need to identify the steady-state degrees of freedom (u) as has already been discussed. Actually, it is the number of u's which is important, because it is does not really matter which variables we include in u, as long as they make up an independent set.

(b) **Identify important disturbances (d) and their expected range.** Next, we must identify the expected range of disturbances (d) for the expected future operation. The most important disturbances are usually related to the feed rate (throughput) and feed composition, and in other external variables such as temperature and pressure of the surroundings. We should also include as disturbances changes in specifications and constraints (such as purity

specifications or capacity constraints) and changes in parameters (such as equilibrium constants, rate constants and efficiencies). Finally, but not least, we need to include as "disturbances", the expected changes in prices of products, feeds and energy.

(c) **Optimize the operation for the expected disturbances.** Here we specify the disturbances (d) and vary the degrees of freedom ($u_{opt}(d)$) in order to minimize the cost (J), while satisfying the constraints. The main objective is to find the *constraints regions* (sets of active constraints) and the optimal nominal setpoints in each region.

Mathematically, the steady-state optimization problem can be formulated as

$$\min_u \ J(u,x,d)$$

subject to:

Model equations: $f(u,x,d) = 0$

Operational constraints: $g(u,x,d) \leq 0$

Here u are the steady-state degrees of freedom, d are the disturbances, x are the internal states, f=0 represents the mathematical model equations and possible equality constraints (like a given feed flow), and $g \leq 0$ represents the operational constraints (like a maximum or nonnegative flow, or a product composition constraint). The process model, f=0, is often represented indirectly in terms of a commercial software package (process simulator), such as Aspen or Hysis/Unisim. This usually results in a large, nonlinear equation set which often has poor numerical properties for optimization.

Together with obtaining the model, the optimization step S2 is often the most time consuming step in the entire plantwide control procedure. In many cases, the model may not be available or one

does not have time to perform the optimization. In such cases a good engineer can often perform a simplified version of step S1-S3 by using process insight to identify the expected active constraints and possible "self-optimizing" controlled variables (CV_1) for the remaining unconstrained degrees of freedom.

A major objective of the optimization is to find the expected regions of active constraints. An important point is that *one cannot expect to find a single control structure that is optimal because the set of active constraints will change depending on disturbances and economic conditions (prices).* Thus, one should prepare the control system for the future, by using offline analysis and optimization to identify regions of active constraints. The optimally active constraints will vary depending on disturbances (feed composition, outdoor temperature, product specifications) and market conditions (prices).

Note here that there generally are two main *modes of operation* depending on market conditions (Rijnsdorp, 1991):

Mode I. Given throughput (buyers market). This is usually the "nominal" mode for which the control system is originally set up. Usually, it corresponds to a "maximize efficiency" situation where there is some "trade-off" between utility (energy) consumption and recovery of valuable product, corresponding to an unconstrained optimum.

Mode II. Maximum throughput (sellers market). When the product prices are sufficiently high compared to the prices of raw materials (feeds) and utilities (energy), it is optimal to increase the throughput as much as possible. However, as one increases the feed rate, one will usually encounter constraints in various units, until eventually we reach the bottleneck where a further increase is infeasible.

Step S3. ***Select "economic" (primary) controlled variables, CV_1*** **(Decision 1):**

We are here concerned with implementing the optimal operation points found in Step S2 in a robust and simple manner. To make use of all the economic degrees of freedom (inputs u), we need to identify as many economic controlled variables (CV_1) as we have inputs (u). In short, the issue is: What should we control?

(a) Identify candidate measurements (y) and their expected measurement error (n^y).
We must first identify all the candidate measurements (y) together with their expected static measurement error (n^y) . In general, we should in the set y include all inputs (valves) to allow, for example, for the possibility of keeping an input constant.

(b) Select primary (economic) controlled variables $CV_1 = Hy$ (Decision 1). Next, the controlled variables are selected among the candidate measurements (see Figure 2), usually by selecting individual measurements. One needs to find one CV_1 for each steady-state degree of freedom (u).

For economic optimal operation, the rules for CV_1 selection are

1. Control active constraints

2. For the remaining unconstrained degrees of freedom: Control "self-optimizing" variables with the objective of minimizing the economic loss with respect to disturbances.

The two rules are discussed in detail below. In general, step S3(b) must be repeated for each constraint region. To reduce the need for switching between regions one may consider using the same CV_1's in several regions, but this is nonoptimal and may even lead to infeasibility.

1. Control active constraints.

In general, the obvious controlled variables to keep constant are the active constraints. The active constraints come out of the analysis in step S2 or may in some cases be identified based on physical insight The active constraints are obvious "self-optimizing" variables and could be input constraints (in the set u) or output constraints

Input constraints are trivial to implement; we just set the input at its optimal minimum or maximum, so no control system is needed. For example, if we are operating a very old car then optimal operation (defined as minimum driving time, J=T) may be achieved with the gas pedal at its maximum position.

For *output constraints*, we need a controller, and a simple single-loop feedback controller is often sufficient. For example, if we have a better car then the maximum speed limit (say 80 km/h) is likely an active constraint and should be selected as the controlled variable (CV_1). To control this, we may use a "cruise controller" (automatic control) which adjusts the engine power to keep the car close a given setpoint. In this case, the speed limit is a *hard constraint* and we need to *back off* from the speed limit (say to a setpoint of 75 km/h) to guarantee feasibility if there is a steady-state measurement error (n^y) or a dynamic control error. In general, we want to minimize the backoff because any backoff results in a loss (i.e., a larger J=T) which can never be recovered.

The backoff is the "safety margin" from the active constraint and is defined as the difference between the constraint value and the chosen setpoint

$$Backoff = |Constraint - Setpoint|$$

In the car driving example, backoff = 5 km/h.

The active constraints should be selected as CVs because the optimum is not "flat" with respect to these variables. Thus, there is often a significant economic penalty if we "back off" from an active constraint, so tight control of the active constraints is usually desired. If a constrained optimization method is used for the optimization, then we can quantify the loss by using the Lagrange multiplier λ associated with the constraint. We have that

$$\text{Loss} = \lambda \cdot \text{backoff}$$

For input (valve) constraints, we usually need no backoff, unless we choose to use the input for stabilization in the lower regulatory (stabilizing) layer because we need some range to use it for control. For output constraints we have two cases

- Soft output constraints (only average value matters): Backoff = measurement error (bias n^y)
- Hard output constraints (must be satisfied at all times): Backoff = measurement error (bias n^y) + control error (dynamic)

To reduce the backoff, we need accurate measurements of the constraint outputs, and for *hard output constraints* we also need tight control with a small dynamic control error. We have the *"squeeze and shift rule"* for hard output constraints: By squeezing the output variation, we can shift the setpoint closer to its limit (i.e., reduce the backoff). For soft output constraints, only the steady-state control error matters, which will be zero if the controller has integral action.

2. **Control "self-optimizing" variable which when held constant keeps the operation close to the optimum in spite of disturbances**.

It is usually simple to identify and control the active constraints. The more difficult question is: What should we use the remaining unconstrained degrees of freedom for? Does it even make a difference what we control? The answer is "yes"!

As an example, consider optimal operation of a marathon runner where the objective is to adjust the power (u) to minimize the time (J=T). This is an unconstrained problem; one cannot simply run at maximum speed ($u=u_{max}$) as for a sprinter (100 m runner). A simple policy is constant speed (c_1=speed), but it is not optimal if there are "disturbances" (d) caused by wind or hilly terrain. A better choice is to run with constant heart rate (c_2=pulse), which is easy to measure with a pulse clock. With a constant heart rate (c_2=constant), the speed (c_1) will increase when we run downhill as one would expect for optimal operation, so heart rate (c_2) is clearly a better "self-optimizing" variable than speed (c1). "Self-optimizing" means that when the selected variables are kept constant at their setpoints, then the operation remains close to its economic optimal in spite of the presence of disturbances (Skogestad, 2000). One problem with the feedback is that it also introduces a measurement error (noise) n^y which may also contribute to the loss, see Figure 2.

In the following let CV_1= c. There are two main possibilities for selecting self-optimizing c = Hy:

1. Select single measurements as CV_1's (H is a selection matrix with a single 1 in each row/column and the rest of the elements 0).

2. Use measurement combinations as CV_1's. Here, methods exist to find optimal linear combinations c =H y, where H is a "full" combination matrix.

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In summary, the problem at hand is to choose the matrix H such that keeping the controlled variables c=Hy constant (at a given setpoint c_s) gives close-to-optimal operation in spite of

the presence of disturbances d (which shift the optimum), and measurement errors n^y (which give an offset from the optimum).

Quantitative approaches. Are there any systematic methods for finding the matrix H, that is, to identify self-optimizing c's associated with the unconstrained degrees of freedom? Yes, and there are two main approaches

1. "**Brute force" approach.** Given a set of controlled variables c=Hy, one computes the cost J(c,d) when we keep c constant ($c = c_s + Hn^y$) for various disturbances (d) and measurement errors (n^y). In practise, this is done by running a large number of steady-state simulations to try to cover the expected future operation. Typically, expected extreme values in the parameter space (for d and n^y) are used to compute the cost for alternative choice of the controlled variables (matrix H). The advantage with this method is that it is simple to understand and apply and it works also for nonlinear plants and even for changes in the active constraint. Only one nominal optimization is required to find the setpoints. The main disadvantage with the method is that the analysis for each H is generally time consuming and one cannot guarantee that all important cases are covered. In addition, there exist an infinite number of choices for H so one can never guarantee that the best c's are found.

2. **"Local" approaches** based on a quadratic approximation of the cost. This is discussed in more detail in Alstad et al. (2009) and references therein. The main local approaches are:

1. **Maximum gain rule.** In words, the maximum gain rule says that one should control "sensitive" variables, with a large gain from the inputs (u) to c=Hy. This rule is good for pre-screening and also yields good insight

2. **Nullspace method.** This method yields optimal measurement combinations for the case with no noise, $n^y=0$. By simulations one must first obtain the optimal measurement sensitivity, **F** =

dy^{opt}/dd. Then, assuming that the number of (independent) measurements y is the sum of the number of inputs (u) and disturbances (d), the optimal is to select H such that HF=0. Note that H is a nonsquare matrix, so HF=0 does not require that H=0 (which is a trivial uninteresting solution), but rather that H is in the nullspace of F^T.

3. **Exact local method** (loss method). This extends the nullspace method to the case with noise and to any number of measurements. For details see Alstad et al. (2009).

The use of these methods is discussed in many papers by Skogestad and coworkers, for example, see Downs and Skogestad (2011) for some practical applications of the nullspace method.

Regions and switching. Note that new self-optimizing variables must be identified (offline) for each region, and that switching of controlled variables is required as one encounters a new region (online). In practise, it is easy to identify when to switch when one encounters a constraint. It seems less obvious when to switch out of a constraint, but actually one simply has to monitor the value of the unconstrained CVs from the neighbouring regions and switch out of the constraint region when the unconstrained CV reaches its setpoint.

As an example, consider a recycle process where it is optimal to keep the inert fraction in the purge at 5% using the purge flow as a degree of freedom (unconstrained optimum). However, during operation there may be a disturbance (e.g. increase in feed rate) so that the recycle compressor reaches its maximum load (e.g. because of constraint on maximum speed). The recycle compressor was used to control pressure, and since it is still optimal to control pressure, the purge flow has to take over this task, This means that one has to give up controlling the inert fraction, which will drop below 5%. In summary, we have gone from an unconstrained operating region (I) where we control the inert fraction to a constrained region (II) where the compressor is at maximum load. In region II, we keep the recycle flow

at its maximum. How do we know when to switch back from region II to region I? We do this by monitoring the inert fraction, and when it reaches 5% we switch back to controlling it (region I).

In general, one would like to simplify the control structure and reduce the need for switching. This may require using a suboptimal CV_1 in some regions of active constraints. In this case the setpoint for CV_1 may not be its nominally optimal value (which is the normal choice), but rather a "robust setpoint" which reduces the loss when we are outside the nominal constraint region.

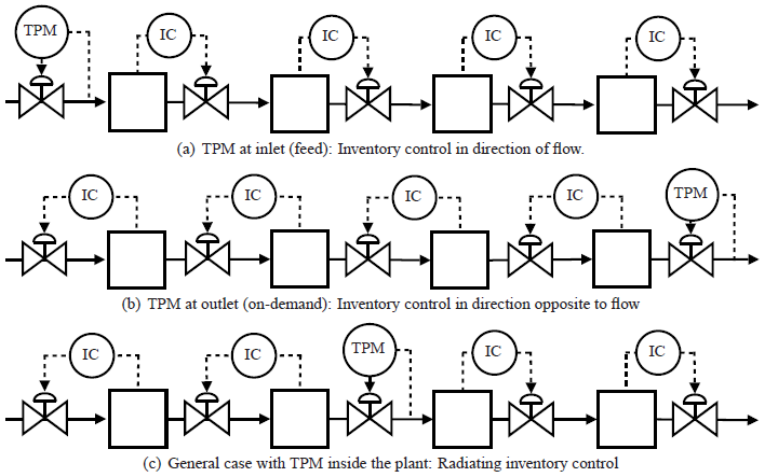

Figure 3. Radiation rule: Local-consistency requires a radiating inventory control around a fixed flow (TPM). (Price and Georgakis, 1993; Aske and Skogestad, 2009).

Step 4. ***Select the location of throughput manipulator*** **(TPM) (Decision 3)**

The main purpose of a process plant is to transform feedstocks into more valuable products and this involves moving mass through the plant. The amount of mass moved through the plant, as expressed by the feed rate or product rate, is determined by specifying one degree of freedom, which we call the throughput manipulator (TPM). The TPM or "gas pedal" is usually a flow but not always, and it is usually set by the operator (manual control). Some plants, e.g., with parallel units, may have more than one TPM. The TPM is usually at a fixed location, but to get better control (with less backoff) one may consider moving the TPM depending on the constraint region.

- Definition (Aske and Skogestad, 2009). *A TPM is a degree of freedom that affects the network flow and which is not directly or indirectly determined by the control of the individual units, including their inventory control.*

The TPM has traditionally been placed at the feed to the plant. One important reason is that most of the control structure decisions are done at the design stage (before the plant is build) where the feed rate is considered fixed, and there is little thought about the future operation of the plant where it is likely that one wants to maximize the feed (throughput). However, as discussed in the following, the location of the TPM is an important decision that links the **top-down and bottom-up part of the procedure.**

Where should we locate the TPM ("gas pedal") for the process?

In principle, the TPM may be located anywhere in the plant, although the operators often prefer to have it at the feed, so this will be the default choice. Note that from a purely steady-state point of view, the location of the TPM does not matter, but it is important dynamically. First, it may affect

the control performance (backoff from active constraints), and second, as soon as the TPM has been placed, the radiation rule (Figure 3) determines the structure of the regulatory layer.

There are two main concerns when placing the throughput manipulator (TPM):

1. **Economics.** The location has an important effect on economics because of the possible backoff if active constraints are not tightly controlled, in particular, for the maximum throughput case where tight control of the bottleneck is desired. More generally, the TPM should then be located close to the bottleneck to reduce the backoff from the active constraint that has the largest effect on the production rate.
2. **Structure of regulatory control system**. Because of the radiation rule (Price and Georgakis, 1993), the location of the throughput manipulator has a profound influence on the structure of the regulatory control structure of the entire plant, see Figure 3.

An underlying assumption for the radiation rule, is that we want "local consistency" of the inventory control system **(**Aske and Skogestad, 2009). This means that the inventory in each unit is controlled locally, that is, by its own in- or outflows. In theory, one may not require local consistency and allow for "long" inventory loops, but this is not common for obvious operational reasons, including risk of emptying or overfilling tanks, startup and tuning and increased complexity.

Most plants have one "gas pedal" (TPM), but there may be more than one TPM for plants with parallel units, splits and multiple alternative feeds or products. Note that the feeds usually need to be set in a fixed ratio, so adding a feed usually does not give an additional TPM. For example, for

the reaction A+B→ C, we need to have the molar ratio F_A/F_B close to 1 to have good operation with small loss of reactants, so there is only one TPM even if there are two feeds, F_A and F_B.

If we only consider a part of the process, then this part may have no TPM. Instead, there will be a given flow, typically a feed or product, that acts as a disturbance on this part process, and the control system must be set up to handle this disturbance. One may also view this as having the TPM at a fixed location. For example, for a utility plant the product rate may be given and in an effluent treatment plant the feed rate may be given. On the other hand, note that a closed recycle system, like the amine recycle in a CO2 gas treatment plant, introduces an extra TPM.

Moving the TPM during operation. Preferably, the TPM fixed should be in a fixed location. First, it makes it simpler for the operators, who usually are the ones who set the TPM, and, second, it avoids switching of the inventory structure, which should be "radiating" around the TPM (Figure 3). However, since the TPM in principle may be located anywhere, it is tempting to use its location as a degree of freedom and move it to improve control performance and reduce backoff. The following rule is proposed:

- *To get tight control of the new active constraint and achieve simple switching, locate the TPM "close" to the next active constraint* (such that the TPM can be used to achieve tight control of the constraint when it becomes active).

The rule is based on economic considerations with the aim of simplifying the required switching when the next capacity constraint becomes active. However, moving the TPM may require switching regulatory loops, which is usually not desirable.

Step 5. ***Select the structure of regulatory (stabilizing) control layer***

The main purpose of the regulatory layer is to "stabilize" the plant, preferably using a simple control structure with single-loop PID controllers. "Stabilize" here means that the process does not "drift" too far away from acceptable operation when there are disturbances. The regulatory layer is the fastest control layer, and is therefore also used to control variables that require tight control, like economically important active constraints (recall the "squeeze and shift" rule. In addition, the regulatory layer should follow the setpoints given by the "supervisory layer".

As discussed in more detail below, the main decisions is Step 5 are to (a) select controlled variables (CV_2) **(Decision 2**) and (b) to *select inputs (valves) and "pairings" for controlling CV_2 (***Decision 4***)**. Interestingly, decision (a) on selecting CV_2 can often be based mostly on steady-state arguments, whereas dynamic issues are the primary concern when selecting inputs (valves) and pairings.

Note that we do not "use up" any degrees of freedom in the regulatory control layer because the setpoints CV_2's are left as manipulated variables (MVs) for the supervisory layer, see Figure 2. However, we do "use up" some of the time window as given by the close-loop response time (bandwidth) of the stabilizing layer

Step 5(a) Select "stabilizing" controlled variables CV_2 (Decision 2). These are typically "drifting" variables such as inventories (level and pressure), reactor temperature, and temperature profile in distillation column. In addition, active constraints (CV_1) that require tight control (small backoff) may be assigned to the regulatory layer. On the other hand, it is usually not necessary with tight control of unconstrained CV_1's because the optimum is usually relatively flat.

To select systematically the stabilizing $CV_2 = H_2$ y, one should consider the behavior of the "stabilized" or "partially controlled" plant with the variables CV_2 being controlled (see Figure 2), taking into account the two main objectives of the regulatory layer:

- **Local disturbance rejection (Indirect control** of primary variables CV_1): With the variables CV_2 controlled, the effect of the disturbances on the primary variables CV_1 should be small. This is to get "fast" control of the variables CV_1, which may be important to reduce the control error (and thus the backoff) for some variables, like active output constraints.

- **Stabilization (Minimize state drift**): More generally, the objective is to minimize the effect of the disturbances on the (weighted) states x. This is to keep the process in the "linear region" close to the nominal steady-state and avoid that the process drifts into a region of operation where it is difficult to recover. The advantage of considering some measure of all the states x is that the regulatory control system is then not tied to a particular control objective (CV_1) which may change with time, depending on disturbances and prices.

When considering disturbance rejection and stabilization, it is the behavior at the closed-loop time constant of the above supervisory layer which is of main interest. Since the supervisory layer is usually relatively slow, it is again (as with the selection of CV_1) usually sufficient to consider the steady-state behavior when selecting CV_2 (however, when selecting the corresponding valves/pairings in Step 5b dynamics are the key issue).

Step 5(b) Select inputs (valve) for controlling CV_2 ***(Decision 4)***. Next, we need to find the inputs (valves) that can be used to control CV_2. Normally, single-loop (decentralized) controllers are used in the regulatory layer, so the objective is to **identify pairings.** The main rule is to "pair close" so that the dynamic controllability is good with a small effective delay and so that the interactions between the loops are small. *In addition, the following should be taken into account:*

- We want “local consistency” for the inventory control (Aske and Skogestad, 2009). This implies that the inventory control system is radiating around the given flow.
- We want tight control of important active constraints (to avoid backoff).
- We should avoid selecting as MVs in the regulatory layer, variables that may optimally saturate (steady state), because this would require either reassignment of regulatory loop (complication penalty), or backoff for the MV variable (economic penalty)

Reassignments (logic) in the regulatory layer should be avoided. Preferably, the regulatory layer should be independent of the economic control objectives (regions of steady-state active constraints), which may change depending on disturbances, prices and market conditions. Thus, it is desirable that the choices for CV_1 (Decision 1) and CV_2 (Decision 2) are independent of each other.

In practise, in order to make the task more manageable, the choice of the regulatory layer structure, may be divided into Step S5.1: Structure of inventory control layer (closely related to Step S4) and Step S5.2: Structure of remaining regulatory control system, but we here consider them combined.

Step 6. Select structure of supervisory control layer.

The supervisory or ”advanced control” layer has three main tasks:

Task 1. Control the primary (economic) controlled variables (CV_1) using as MVs the setpoints to the regulatory layer plus any remaining (“unused”) valves (see Figure 2).

- The supervisory layer may use “dynamic” degrees of freedom, including level setpoints, to improve the dynamic response (at steady state these extra variables may be “reset” to their ideal resting values).

- The supervisory layer may also make use of measured disturbances (feedforward control).
- Estimators: If the primary controlled variables (CV_1) are not measured, typically compositions or other quality variables, then "soft sensors" based on other available measurements may be used for their estimation. The "soft sensors" are usually static, although dynamic state estimators (Kalman Filter, Moving horizon estimation) may be used to improve the performance. However, these are not common in process control, because the supervisory layer is usually rather slow.

Task 2. Supervise the performance of the regulatory layer. The supervisory layer should take action to avoid saturation of MVs used for regulatory control, which otherwise would result in loss of control of some "drifting" variable (CV_2).

Task 3. Switch controlled variables and control strategies when disturbances or price changes cause the process to enter a new region of active constraints.

Implementation. There are two main alternatives in terms of the controller used in the supervisory layer:

Alternative 1. "Advanced single loop control" = PID control with possible "fixes" such as feedforward (ratio), decouplers, logic, selectors and split range control (in many cases some of these tasks are moved down to the regulatory layer). With single-loop control an important decision is to select pairings. Note that the issue of finding the right pairings is more difficult for the supervisory layer because the interactions are usually much stronger at slower time scales, so measures such as the relative gain array (RGA) may be helpful.

Alternative 2. **Multivariable control (usually MPC).** Although switching and logic can be reduced when using MPC, it cannot generally be completely avoided. In general, it may be necessary to change the performance objective of the MPC controllers as we switch regions.

Step 7. Structure of (and need for) optimization layer (RTO) (Related to Decision 1):

The task of the RTO layer is to update the setpoints for CV_1, and to detect changes in the active constraint regions that require switching the set of controlled variables (CV_1).

In most cases, with a "self-optimizing" choice for the primary controlled variables, the benefits of the RTO layer are too low to justify the costs of creating and sustaining the detailed steady-state model which is usually required for RTO. In addition, the numerical issues related to optimization are very hard, and even offline optimization is difficult.

5. DISCUSSION: COMPARISON OF THE PROCEDURES OF LUYBEN AND SKOGESTAD.

The most striking difference between the two procedures is that whereas the Skogestad procedure starts with economics (part I), the Luyben procedure does not explicitly include economics, except at the very last stage, where all the decisions have been made already.

In the following, the steps of the Luyben procedure (Luyben et al, 1997, 1998) are discussed and compared with the Skogestad procedure.

Step L1. Establish control objectives. By "control objectives", Luyben means the primary CVs but the Luyben procedure is unclear about how these should be selected. It is stated that *"this is probably the most important aspect of the problem because different control objectives lead to different control structures"*, but the only guideline is that *"these objectives include reactor and separation yields, product quality specifications, product grades and demand determination, environmental restrictions, and the range of safe operating conditions. "*

In the Skogestad procedure, the first step is to define the cost function and the process constraints (step S1) and optimize the operation (step S2). The selection of CVs follows from this (step S3). The first thing is to control the active constraints. This will generally include product quality specifications on valuable products (cheap products should often be overpurified to avoid losses of more valuable components), minimum product rates (demands), environmental and safety constraints, pressure and temperature constraints and so on. For output constraints one may have to introduce a safety factor ("backoff") which will imply an economic loss. To reduce the backoff of hard output constraints one wants tight control, which may imply that some of these variables are controlled in the regulatory layer.

Step L2 (and Step S2a). Determine control degrees of freedom. This is an important step in both procedures, but in the Skogestad procedure it comes before the selection of CVs, which is reasonable because we need to identify one CV for each degree of freedom. In addition, in Skogestad's procedure one distinguishes clearly between the steady-state degrees of freedom (step S2a) and the physical degrees of freedom (valves, step S5b).

Luyben states that most of the control degrees of freedom (valves) are used to achieve basic regulatory control of the process: "*(1) set production rate, (2) maintain gas and liquid inventories, (3) control product qualities and (4) avoid safety and environmental constraints"*. He adds that "*any valves that remain after these vital tasks can be utilized to enhance steady-state economic objectives or controllability*".

This is in agreement with the Skogestad procedure. Note that many of these variables are related to optimal active constraints. Control of gas inventories (pressures) is usually required to stabilize the plant (avoid drift), but note that one does not really "consume" any degrees of freedom here because the pressure setpoint can be used as a degree of freedom for effecting the economic (steady-state) operation. With liquid inventories (levels) the situation is a bit different because many liquid levels do not have a steady-state effect.

Step L3. Establish energy management system. It seems a bit unclear why this issue is so high up on the list in the Luyben procedure and what is so special about control of the energy system. Of course, an unstable exothermic reactor needs to be stabilized and selecting an appropriate sensitive variable (typically, a temperature) and pairing it with an input (MV) will be one of the first issues when designing the regulatory control system (Step S5). However, since stabilizing control does not "use up" any degrees if freedom at steady-state, this may not be in conflict with the objectives of optimal economic operation, which is the third step (or actually Step S2b) in Skogestad's procedure.

Step L4 (= Step S4). Set the production rate. Note that in this work the terms "production rate" and "throughput" mean the same. As discussed in detail above, the location of the throughput manipulator (TPM) is very important, both for economic reasons (steady-state) and for dynamic

reasons. For economic reasons, it should be close to the bottleneck in order to reduce the backoff when it is optimal to maximize production (sellers market) (Skogestad, 2004). Dynamically, it determines the structure of the inventory control system, which to have "radiating" around the TPM (Price and Georgakis, 1993).

Traditionally, the main process feed has been selected as the "gas pedal" (TPM). However, Luyben et al. (1998) recommend to locate it close to the reactor: "*Establish the variables that dominate the productivity of the reactor and determine the most appropriate TPM*". Again, the reasoning for focusing on the reactor is a bit unclear, and it is worth mentioning that the location of the TPM is also an issue in plants with no reactor, like a gas processing plant (Aske et al., 2008). Nevertheless, the reactor is obviously an important unit and will often be the bottleneck for the process. In addition, there is usually recycle around the reactor, and by locating the TPM in this recycle loop one can avoid the "snowball effect" and satisfy Luyben's rule L6 of "fixing a flow in every recycle loop".

Step L5. Control product quality and handle safety, environmental and operational constraints. Luyben says that we should *"select the best variables to control each of the product-quality, safety and environmental variables"*, but he does not state what "best" is. He adds that *"we want tight control of these important quantities for economic and operational reasons"* which makes sense if these variables are optimally active constraints. Having performed an economic optimization (Step S2b), it is then also easy to determine what these "best" variables are: They are active constraints when we operate the plant such that cost is minimized.

Luyben adds that *"it should be noted that establishing the product-quality loops first, before the material balance control structure, is a fundamental difference between our plantwide control procedure and Buckley's (1964) procedure"*.

In this respect, the Skogestad procedure is something in between the Luyben and Buckley procedures. Similar to Luyben, it starts by identifying which variables should be controlled, which typically includes some active product-quality constraints. However, similar to Buckley, in the Skogestad procedure the design of the actual control system, including choice of pairings, starts with the "stabilizing" loops, including the material balance (inventory) control, although it is recommended that control of active constraints that require tight control for economic reasons should be assigned to the regulatory layer.

Step L6. "Fix a flow in every recycle loop and control inventories". The recycle split adds a degree of freedom to the process, so it is possible to fix a flow in every recycle loop. This may be a good strategy from a regulatory and dynamic point of view, but not generally from an economic point of view. For example, if the throughput in the plant is increased then it will generally be economically optimal to increase all flows, including the recycles.

Luyben argues that fixing a flow in the recycle loop avoids the "snowball effect" where the recycle flow grows out of bound (Luyben, 1994). Note that the "snowball effect" is caused by having a unit, typically a reactor, which is too small compared to the desired throughput. This means that we are operating at high throughputs where the unit indirectly is the bottleneck of the process.

The systematic approach would be to perform an economic optimization with the throughput as a degree of freedom (Step S2b), and from this optimal control policy will follow (Step S3), which will give the optimal way of handling the "snowballing". In some cases, maximum production may correspond to maximal recycle, which means that "snowballing" is optimal and the recycle flow should simply be fixed at its maximum (e.g., maximum gas recycle, Luyben et al., 1998, Araujo and Skogestad, 2008). In other cases, maximum recycle is not optimal because other constraints are reached, and one needs to use a flow in the recycle to control an optimally active constraint (e.g., use the column feed flow to control product composition in the simple recycle system studied by Luyben (1994) and later by Larsson et al. (2003). In yet other cases, there may be an "optimal maximum throughput" and one needs to identify a self-optimizing variable associated with the feed being an unconstrained degree of freedom.

Nevertheless, Luyben is obviously right that with the TPM located outside the recycle loop, and with all the flows inside the recycle loop on inventory (level or pressure) control, one may get "snowballing" inside the recycle loop if we feed more into the loop than its units can handle. "Snowballing" is clearly a "drifting mode" and it is a task of the regulatory control system to avoid drift (step S5). Snowballing is caused by the positive feedback in the recycle loop, and one way to break this loop is to follow Luyben and fix a flow in the recycle loop (including selecting it as an TPM). This forces the excess feed to exit the recycle loop. Another option, which is likely to be better economically, is to use one of the flows in the recycle loop to control some other variable, like a sensitive temperature or composition.

In summary, the importance of the "snowball" effect has probably been overemphasized in some of the literature on plantwide control. If it is actually a "problem", then it cannot be economically

optimal, so it will automatically be avoided by following the procedure of Skogestad. Nevertheless, one should be aware of the "snowballing" that may occur if all the flows inside the recycle loop are on level or pressure control.

L7. Check component balances. "*Identify how chemical components enter, leave, and are generated or consumed in the process (Downs drill)".* This is a very important issue, even for processes without reactions, and is included in step S5 (regulator control strategy) in the Skogestad procedure.

L8. Control individual unit operations. This step seems a bit arbitrary, as application of the previous steps will "automatically" lead to control of the individual units. Of course, it can be useful to compare the resulting control structure with common rules of thumb for individual units and consider changes if it seems unreasonable. The Skogestad procedure contains many steps where choices are made, so some iteration may be needed.

L9. Optimize economics and improve dynamic controllability. Luyben writes that *"after satisfying all of the basic regulatory requirements, we usually have additional degrees of freedom involving control valves that have not been used and setpoints in some controllers that can be adjusted. These can be utilized wither to optimize steady-state economic process performance or to improve dynamic response."* This statement is true, but it is better to consider the economics much earlier. First of all, an economic analysis is generally needed in order to identify the optimally active constraints (in Step L1), so one may as well identify good self-optimizing CVs for the remaining unconstrained degrees of freedom. Second, if one knows the self-optimizing variables, then one can take this into account when designing the regulatory control system.

6. CONCLUSION

Control structure design deals with the structural decisions of the control system, including what to control and how to pair the variables to form control loops. Although these are very important issues, these decisions are in most cases made in an ad hoc fashion, based on experience and engineering insight, without considering the details of each problem. In the chapter, a systematic procedure for control structure design for complete chemical plants (plantwide control) is presented. It starts with carefully defining the operational and economic objectives, and the degrees of freedom available to fulfil them. Then the operation is optimized for expected future disturbances to identify constraint regions. In each region, one should control the active constraints and identify "self-optimizing" variables for the remaining unconstrained degrees of freedom. Following the decision on where to locate the throughput manipulator (TPM), one needs to perform a bottom-up analysis to determine secondary controlled variables and structure of the control system (pairing).

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