Symposium on Recent and Emerging Advances in Chemical Engineering (REACH), Indian Institute of Technology Madras, Chennai, India. December 2-4, 2010.

A Systematic Approach To Plantwide Control

Sigurd Skogestad
Department of Chemical Engineering
Norwegian University of Science and Technology (NTNU)
N7491 Trondheim, Norway
skoge@ntnu.no

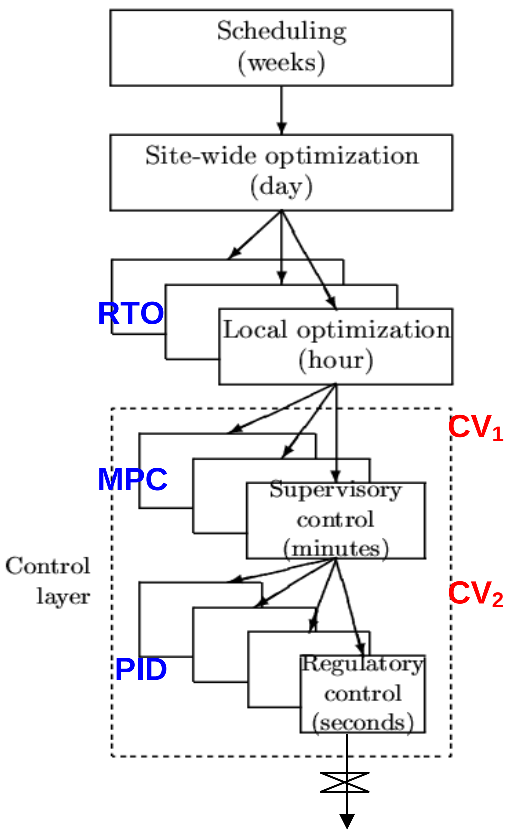
ABSTRACT

This paper summarizes my struggles in the plantwide control field.

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.*



My interest in this field of plantwide control dates back to 1983 when I started my PhD work at Caltech. As an application, I worked on distillation column control, which is excellent example of a plantwide control problem. I was inspired by Greg Shinskey's book on Distillation Control, which came out with a second edition in 1984 (Shinskey, 1984). In particular, I liked his systematic procedure, which involved computing the steady-state relative gain array (RGA) for 12 different control structures ("configurations"); the DV-configuration, LV-configuration, ratio configuration, and so on. However, when I looked in more detail on the procedure I discovered that its theoretical basis was weak. First, it did not actually include all structures, and it even eliminated the DB-configuration as "impossible" even through it is workable in practise (Luyben, 1989). Second, controllability theory tells that the steady-state RGA by itself is actually not useful, except that one should avoid pairing on negative gains. Third, the procedure focused on dual composition control, while one in practise uses only single end control, for example, because it may be optimal economically to use maximum heating to maximize the recovery of the valuable product.

Furthermore, when I studied the distillation column control problem in more detail, I discovered that there were several control objectives, which often were conflicting. First, there was the issue of "stabilizing control" which involved closing the level and pressure loops, and maybe also a temperature loop, so that the column did not drift and could be controlled manually which too much effort. Second, there was the issue of "economic control" (advanced and supervisory control) which involves keeping the column close to its economically optimal operation. In many cases, "economic control" was the same as "dual composition control" but not always.

In fact, depending on marked conditions and disturbances, the best economic mode of operation changes. For a distillation column, it is always optimal to control the valuable product at its spec. to avoid product "give-away" (however, if the valuable product is recycled then there may be no spec). For the "low-value" product it is often optimal to overpurify in order to minimize the loss of valuable product and if product prices are sufficiently high (compared to energy prices) then it is optimal to use maximum energy (boilup) to get "maximum overpurification". The important conclusion from this is that the optimal configuration will change depending on marked conditions, so there is no single "best" control configuration, even for a given column.

Enough about distillation. Another heavy influence on my work was the famous critique article from Alan Foss ("Critique of chemical process control theory", AIChE Journal,1973). He writes:

- *The central issue to be resolved ... is the determination of control system structure.* ***Which variables should be measured, which inputs should be manipulated and which links should be made between the two sets?*** *There is more than a suspicion that the work of a genius is needed here, for without it the control configuration problem will likely remain in a primitive, hazily stated and wholly unmanageable form. The gap is present indeed, but contrary to the views of many, it is the theoretician who must close it.*

Here, he states that "determination of control system structure", which for process control is the same as what I call "plantwide control", is the central issue to be resolved in control. This statement should inspire people to work on the plantwide control. Then, he adds that this most likely will require the work of a genius. I am not sure if this addition is correct, and I am also unsure if it was so smart to make this addition if he wanted to inspire people to work on plantwide control. Nevertheless, it did inspire me and I have worked on the problem since then. Now, after 25 years, I am finally approaching a situation where I have a reasonably clear picture on how to approach the problem. This paper provides the main ideas.

2 CONTROL LAYERS AND TIME SCALE SEPARATION

I define the term plantwide control as "control structure design applied to chemical plants". Here, "control structure design" 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.).

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

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 is poorly defined in terms of its objective. Usually, in control, the objective is that the CV (output) should remain close to its setpoint. However, what should we control? The answer lies in considering the overall plant objective, which is to minimize cost (=maximize profit) while satisfying operational constraints imposed by the equipment, marked demands, product quality, safety, environment and so on. We will get back to this.

Actually, the "original" mathematical problem is not so difficult to formulate and with today's computing power it may even be solvable. It would involve obtaining a detailed dynamic and steady-state model of the complete plant, defining all the operational constraints, defining all available measurements and manipulations, defining all expected disturbances, defining expected, allowed or desirable ranges for all variables, and then designing a nonlinear controller that satisfies all the constraints and objectives while using the possible remaining degrees of freedom to minimize the cost. This would involve 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 and on some very simple processes, it seems to be safe to assume that it will never be applied to any normal-size chemical plant. There are many reasons for this, but one important one is that in most cases one can acceptable control 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 plants operate well with simple control structures. A related example is control of biological systems. These are extremely complex and there is no computing power available (the brain has many good features but exact computations is not one of its strong sides) to do the optimal centralized control task, so one has to rely on very simple, but still effective, control strategies.

So how are real systems controlled in practise? The main simplification is to decompose the overall control problem in to many simple 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 if 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 also Figure 1)
    - scheduling (weeks)
    - site-wide optimization (day)
    - local optimization (hour)

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

We generally have more centralization as we move upwards in the hierarchy. Such a hierarchical (cascade) decomposition with layers operating on different time scale is used in the control of all real (complex) systems including biological systems and airplanes, so the issues raised in this paper (section) are of general interest and not limited to process control.

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 upper layers. 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 send as setpoints for the controlled variables (CVs) to the layer below.

3 A PLANTWIDE CONTROL PROCEDURE

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 (the setpoints CV_{1s} link the optimization layer with the control layers).

Decision 2. Select ”stabilizing” (secondary) controlled variables (CV_2) for the regulatory control layer (the setpoints CV_{2s} link the two control layers).

Decision 3. Locate the throughput manipulator (TPM).

Decision 4. Select pairings for the stabilizing layer, that is, pair inputs (valves) and controlled variables (CV2). By “valves” is here meant the original dynamic manipulated variables.

The remaining part of this paper gives a summary of my proposed systematic procedure for plantwide control. The key idea is to start “top-down” with the economics and identify the various operating regions (Figure 1). In each of these regions we need to identify the controlled variables that lonk optimization and control. After an intermediate step where we consider the location of the throughput manipulator, we proceed with the “bottom-up” design of the stabilizing layer. The goal os to find a simple control structures that combines all the possibly conflicting objectives with minimal need for switching and other kinds of supervisory control.

Overview of plantwide control procedure

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

Step 1. Define operational objectives (economic cost J to be minimzed) and constraints.

Step 2. Identify degrees of freedom (MVs) and optimize the operation for important disturbances (offline analysis) to identify regions of active constraints.

Step 3.Each region of active constraints: Select primary (economic) controlled variables CV1 (**Decision 1**)

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

II. Bottom-up part(focus on dynamics)

Step 5. Choose structure of regulatory control layer (including inventory control)

- Select “stabilizing” controlled variables CV2 (**Decision 2**)
- Select inputs (valves) and “pairings” for controlling CV2 (**Decision 4**)

Step 6. Select structure of supervisory control layer.

Step 7. Select structure of (or need for) optimization layer (RTO)

Step 1. Define operational objectives (economic cost) and constraints.

A systematic approach to plantwide control requires that we first quantify the operational objectives with a scalar cost function J (or equivalently, a scalar profit function, P = -J). This is usually not very difficult and typically we have

J = cost feed + cost utilities (energy) – value products

The goal of operation (and of control) is to minimize the cost J (or equivalently, to maximize the profit P = - J). The cost J should be minimized subject to satisfying the operational constraints (g <= 0), including safety and environmental constraints. Typical operational constraints are minimum and maximum values on flows, pressures, temperatures and compositions. For example, we always have that all flows, pressures and compositions must be non-negative.

Step 2. Identify degrees of freedom (MVs) and optimize operation for important disturbances (offline analysis)

Identifying the economic (usually steady state) degrees of freedom (u) is not usually as simple as one would expect. One approach is to first identify all the (*dynamic* and steady state) control degrees of freedom (inputs, "valves") and then subtract the ones with no steady state effect. The optimization of the plant for expected operation is usually the most time consuming step. Note that there are two main modes of operation

Mode I. Given throughput: Maximize efficiency.

Mode II. Throughput is a degree of freedom: Find optimal throughput (often = maximum throughput) Maximize throughput (with production rate as a degree of freedom).

Altthough the plant and its control system is usually designed for mode 1, the most profit usally made when product process are high and optimal operation is tha same as maximum throughput (mode II). Plantwide control should generally be focused much more on mode II.

Step 3. Selection of primary controlled variables (CV1)

"What should we control"? This has been the main topic for my research in the plantwide control area over the last 25 years, and I think I finally have found the solution,

1. Control active constrains
2. Control "self-optimizing" CV1s for the remaining unconstrained degrees of freedom

In particular, the research has been focused towards the latter unconstrained CV1s. "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.

In general, the controlled variables (CV1) will be individual measurements or combinations of measurements. Thus we can write, Cv1 = H y. Here, y contains all the available measurements (including MVs) and H is the selection or combination matrix to be found. Unfortunately, I do not have space to go into this favorite topic of mine, but I refer the interested reader to paper by Alstad et al. (2009)

Step 4. Select location of throughput manipulator-(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 feedrate or product rate, is determined by specifying one degree of freedom, which we call the throughput manipulator (TPM). Most plants have one TPM, but complex plants with

parallel units and multiple or alternative products may have more than one TPM. From a steady-state point of view the location of the TPM does not matter, but it is important dynamically.
There are two main concerns when placing the throughput manipulator (TPM):
1. Economics. This is relevant when the plant is operated at maximum capacity and the TPM should then be located close to the bottleneck to reduce the back-off 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 X.

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 almost never done for obvious operational reasons, including risk of emptying or overfilling tanks, startup and tuning and increased complexity.

Step 5. Choose structure of regulatory (stabilizing) layer

The 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" away from acceptable operating conditions when there are disturbances. In addition, the regulatory layer should follow the setpoints given by the "supervisory layer".

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 marked conditions. The main decisions are:

(a) Identify "stabilizing" CV2s. These are typically "drifting" variables such as levels, pressures, reactor temperature, temperature profile in distillation column. In addition, active constraints (CV1) that require tight control (small backoff) may be assigned to the regulatory layer. (Comment: it is usually not necessary with tight control of unconstrained CVs because optimum is usually relatively flat). . Note that we do not "use up" any degrees of freedom in the regulatorey control layer because the setpoints CV_{2s} are left as manipulated variables (MVs) for the supervisory layer. To some extent the choices for CV_1 and CV_2 (Decision 1) are independent of each other.

(b) Identify pairings (MVs to be used to control CV2 ***(Decision 4)***), taking into account
- Want "local consistency" for the inventory control (Aske and Skogestad, 2009). This implies that the inventory control system is radiating around the given flow.
- Want tight control of important active constraints (to avoid back-off). The main rule is to "pair close".
- 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
- back-off for the MV variable (economic penalty)

Step 6. Select structure of supervisory control layer.

Objectives of supervisory layer:
1. Switch control structures (CV1) depending on operating region
2. Perform "advanced" economic/coordination control tasks:

- Control primary variables CV1 at setpoint using as degrees of freedom (MV):
- Setpoints to the regulatory layer (CV2s)
- ”unused” degrees of freedom (valves)
- Keep an eye on stabilizing layer
- Avoid saturation in stabilizing layer (which usually requires back-off and thus economic penalty)
- Feedforward from disturbances (If helpful)
- Make use of extra inputs
- Make use of extra measurements

Implementation

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.

Alternative 2. Multivariable control (usually MPC)

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

- Do we need it?

4 CONCLUSION

This paper presents the current version of the systematic plantwide control procedure. It is still being updated and tested out on applications, but after having worked on this issue of about 25 years, I have good hopes of converging at final procedure by about the year 2025.

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