5. Plantwide Control Design

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5.1 Overview

5.1.1 Control hierarchy

In practice, the control system of a (chemical) plant is divided into several layers. Typically, layers include:
- scheduling (weeks)
- site-wide optimization (day)
- local optimization (hour)
- supervisory/predictive control (minutes)
- regulatory control (seconds)

The optimization layer typically recomputes new setpoints only once an hour or so, whereas the feedback layer operates continuously. The layers are linked by the controlled variables, whereby the setpoints are computed by the upper layer and implemented by the lower layer.

[From Larsson (2000): Studies on Plantwide Control (Dr.Ing. Thesis)]
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5.1 Overview

5.1.2 Design objectives

The goals for an effective plantwide process control system include:

- safe and smooth process operation
- tight control of product quality in the face of disturbances
- a control system run in automatic, not manual, requiring minimal operator attention
- rapid rate and product quality transitions
- minimum environmental releases

By the term *plantwide control* it is *not* meant the tuning and behaviour of all control loops in the plant, but rather the:

- *control philosophy* of the overall plant with emphasis on
- *structural decisions*, which include the
  - selection/placement of manipulators and measurements as well as
  - *decomposition* of the overall problem into smaller subproblems (the control configuration).
5. Plantwide Control Design

5.1 Overview

5.1.3 Heuristic design methods

The plantwide control design problem is difficult to define and solve mathematically because of

- the size of the problem
- the large cost involved in making a precise problem definition, which would require, e.g., a detailed dynamic and steady state model

An alternative to this is to develop heuristic rules based on experience and process understanding.

Luyben et al. (1999) have proposed the following 9-step procedure:

1. Establish the control objectives.
2. Determine the control degree of freedoms.
3. Establish an energy management system.
4. Regulate the production rate.
5. Regulate the product quality.
6. Directly regulate the flow rate in every recycle loop and control inventory.
7. Perform a component balance for each chemical component.
8. Regulate individual unit operations.
9. Optimize the economics and/or improve dynamic controllability.

This procedure is based on heuristics developed in industry and academia.
Buckley basics (1964)

Page Buckley of DuPont was the first to suggest the idea of separating the plantwide control problem into two parts:

- material balance control
- product quality control

He suggested looking first at the flow of material through the system. A logical arrangement of

- level and pressure control loops is established

He then proposed establishing the product-quality control loops by

- choosing appropriate manipulated variables (control variables)

The time constants of the closed-loop product-quality loops are estimated. It is desired to make them as small as possible so that

- good, tight control is achieved

The time constants of liquid-level loops should be about 10 times larger than the product-quality time constants to enable

- independent tuning of material-balance and product-quality loops
Douglas doctrines (1988)

Douglas points out that in the typical chemical plant the costs of raw materials and the value of the products are usually much greater than the costs of capital and energy. This leads to the two Douglas doctrines:

- Minimize losses of reactants and products. This implies that tight control of stream compositions exiting the process are needed.
- Maximize flowrates through gas recycle systems to improve the yield. This rests on the principle that yield is worth more than energy.

The control structure implication is that

- it is not attempted to regulate the gas recycle flow
- the recycle flow rate is maximized

This

- removes one control-degree of freedom
- simplifies the control problem
Jim Downs of Eastman Chemical Company has pointed out the importance of looking at the chemical component balances around the entire plant.

- It must be ensured that all components (reactants, products, and inerts) have a way to leave or be consumed within the process.
- Usually, inerts are not difficult to handle.
  - Heavy inerts can leave the system in the bottoms product from a distillation column.
  - Light inerts can be purged from a gas recycle stream or from a partial condenser on a column.
  - Intermediate inerts can be removed in side stream purges or in separate distillation columns.
- Most of the problems occur in the consideration of reactants.
  - All of the reactants fed into the system must either be consumed via reaction or leave the plant as impurities in the exiting streams.
  - To minimize raw material costs and maintain high-purity products, most of the reactants fed into the process must be consumed in the reactions.
  - The stoichiometry must be satisfied *down to the last molecule* because the plant essentially operates as an integrator in terms of reactants.
5.1 Overview

5.1.3 Heuristic design methods

**Luyben laws (1997)**

As a result of a number of case studies of many types of systems, Bill Luyben suggests the following

- A stream somewhere in all recycle loops should be flow controlled to prevent the snowball effect.
- For systems with reaction types such as \( A + B \rightarrow \text{products} \), a fresh reactant feed stream cannot be flow-controlled unless there is essentially complete one-pass conversion of one of the reactants.
- If the final product comes from a distillation column, the column feed should be
  - liquid, if the product is in the distillate flow
  - vapour, if the product is in the bottoms flow

By this arrangement, disturbances in the feed flow rate or the feed composition have a smaller dynamic effect on the product than otherwise.
5.1 Overview

5.1.3 Heuristic design methods

**Richardson rule**

Bob Richardson of Union Carbide suggested the heuristic that

- the largest stream should be selected to control the liquid level in a vessel

because it provides more “muscle” to achieve the desired control objective. The point is that the bigger the handle you have to affect a process, the better you can control it. This is why there are often fundamental conflicts between steady-state design and dynamic controllability.

**Shinskey schemes (1988)**

Greg Shinskey of Foxboro has proposed a number of "advanced control" structures that permit improvements in dynamic performance. These schemes are not only effective, but they are simple to implement in basic control instrumentation. Liberal use should be made of

- ratio control
- cascade control
- override control
- valve-position (optimizing) control
5. Plantwide Control Design

5.1 Overview

5.1.4 Mathematical design methods

In a mathematical sense, the plantwide control problem is
- a formidable (and almost hopeless) combinatorial problem involving a large number of discrete decision variables
- poorly defined in terms of its objective

Usually, in control, the objective is that the controlled variables should remain close to their setpoints. However, what should we control
- to minimize cost (=maximize profit)
- while satisfying operational constraints imposed by the
  - equipment
  - marked demands
  - product quality
  - safety
  - environment
  - and so on
5.1 Overview

5.1.4 Mathematical design methods

**Open-loop methods**

The assumption is that if a mathematical controllability analysis can determine the dynamic performance of the process at the designed conditions,

- the screening of alternate designs will be more efficient;
- only those designs that are fully controllable will survive.

Here, the definition of controllability is the

- ability of the process to achieve acceptable control performance by regulating the controlled variables within specified bounds using the available actuators.

Quantitative tools that are often used to establish controllability are the

- *relative gain array* (RGA)
- *Niederlinski Index* (NI)

For further details, see Chapter 3 of this course.

An advantage of open-loop methods is that they are easy to employ already when the process is designed as an aid in process design to guarantee that the process is controllable.
5.1 Overview

5.1.4 Mathematical design methods

**Closed-loop methods**

Closed-loop methods address the dynamic process and controller designs simultaneously. There are

- systematic design procedures that use hybrid mixed integer linear programming (MILP).

The practical problem is that this involves

- obtaining a detailed dynamic 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.

In principle, the optimal structure is a centralized controller which collects all available information and computes changes in all control variables.

However, in all likelihood a fully centralized controller will never be applied to any normal-size chemical plant because acceptable control can be achieved with much less effort using simple structures where each controller block only involves a few variables.
5. Plantwide Control Design

5.1 Overview

5.1.5 Combined design methods

It is clear that both
- engineering knowledge
- mathematical analysis
are useful in addressing a plantwide design and control problem.

The systematic design procedure presented next is such a method.
5. Plantwide Control Design

5.2 A systematic design procedure

The design procedure presented in this section is mainly based on
- Skogestad and coworkers (several articles)

The design procedure, which is used to develop multiloop SISO (and MIMO) control strategies for plantwide control, assists the engineer in determining
- how to choose the best controlled, manipulated, and measured variables in the plant,
- when to use advanced control techniques such as MPC, and
- how to select appropriate multiloop control structures with minimum interactions among the coupled processes in the plant.

It is important to realize that the design of plantwide control systems is an art as well as a science. Typically, more than one design will be satisfactory; thus, there is no single solution to the design problem.
5. Plantwide Control Design

5.2 A systematic design procedure

The design procedure generally involves iteration of individual steps until a satisfactory design is obtained. Thus, the design procedure produces preliminary designs that are subject to further exploration and refinement.

Simulation methods should be employed to examine alternative control configurations while exploring the effect of controller tuning on the response of key process variables.

The goal is a plantwide control system design that is no more complicated or expensive than necessary and that, when built, can be operated easily by typical plant operators.

Ultimately, the only definitive way of validating a selected plantwide control system design is by plant tests and by the operating plant’s performance.
5. Plantwide Control Design

5.2 A systematic design procedure

5.2.1 Combined top-down/bottom-up design

In a pure *top-down* design, optimization methods could be used to develop the plantwide control system based on

- a comprehensive dynamic model of the entire plant
- complete operating requirements for the plant

Unfortunately, such an approach is *impractical* because of the large number of process variables involved in modern processing plants.

On the other hand, some aspects of a top-down design approach may be quite *useful* as part of a realistic design procedure.

The traditional design procedure used for industrial control systems has been a *bottom-up* approach, where the control system is designed unit-by-unit (i.e., the entire plant is not considered at this point).

Even though it *incorporates systematic methods* to develop the control structure, this approach also relies on *heuristic design methods* and *rules of thumb* developed from previous designs and the experience of both the process and control system design groups.

In the following, a *combined top-down/bottom-up* method is presented.
5.2 A systematic design procedure  5.2.1 Top-down/bottom-up design

**Step I. Control system design objectives**

**A.** State the plant production, economic, and control objectives, including composition and production rate specifications of all products.

**B.** Identify process constraints that must be satisfied, including safety, environmental, and quality restrictions.

**Step II. Top-down analysis**

**A.** Identify the process variables, control degrees of freedom, control structure, and options for decomposition.

1. Identify the potential controlled variables (CVs).
2. Determine how the CVs can be measured or inferred, and identify other process variables to be measured.
3. Select the potential manipulated variables (MVs).
4. Perform a preliminary control degrees of freedom analysis (compare the numbers of potential manipulated and controlled variables).
5.2.1 Combined top-down/bottom-up design

Step II. Top-down analysis

5. Identify the source and nature of the significant disturbances that must be mitigated.

6. Perform a structural analysis based on a steady-state model, select the final controlled and manipulated variables, and evaluate the possibilities for decomposition of the control problem.

B. Establish the overall control structure (in conceptual form).

1. Identify where the production rate of each product will be measured and controlled.

2. Identify how quality will be measured for each product, and how product quality will be controlled.

3. Determine how each recycle loop throughput/composition will be controlled.

4. Specify how the constraints will be satisfied.

5. Determine how major disturbances will be handled.

6. Analyze the energy management scheme, and indicate conceptually how it will be controlled.
5.2 A systematic design procedure

### 5.2.1 Top-down/bottom-up design

#### Step III. Bottom-up design

**A.** Develop a strategy for regulatory control.

1. Specify how the control system will respond to unsafe or abnormal operating conditions and deal with constraints.
2. Identify control loops to regulate production rates and inventories.
3. Identify control loops that will mitigate major disturbances.

**B.** Examine the potential of applying advanced control strategies.

1. Evaluate the use of enhanced single-loop control strategies, including feedforward, ratio, cascade, and selective control schemes.
2. Employ MIMO control for highly interactive processes.

**C.** Evaluate the economic benefits of real-time optimization.
5.2 A systematic design procedure

5.2.1 Top-down/bottom-up design

**Step IV. Validate the proposed control structure**

**A.** Perform a final degrees of freedom analysis. Check the allocation of the $N_m$ degrees of freedom.

**B.** Check control of individual process units.

**C.** Check the effect of constraints and disturbances on manipulated and controlled variables.

**D.** Simulate control system performance for a wide range of conditions.
5.2.2 Guidelines for selection of variables

Some general aspects

In general, it is desirable to have at least as many manipulated variables (MVs) as controlled variables (CVs).

It may be infeasible to control some output variables, e.g.,
- it may not be possible or economical to measure some outputs, especially chemical composition;
- there may not be enough MVs;
- potential control loops may be impractical because of slow dynamics, low sensitivity to MVs, or interactions with other control loops.

In general, CVs are measured on-line, and the measurements are used for feedback control.

Sometimes it is possible to control an unmeasured variable by using a process model (a soft sensor) to estimate it from measurements of other variables. This strategy is called inferential control.

In the following some general guidelines for the selection of controlled, manipulated, and measured variables are given.
5.2 A systematic design procedure

5.2.2 Selection of variables

**Controlled variables**

1. All variables that are not self-regulating must be controlled.
   - A non-self-regulating variable is an output variable that has an unbounded step response, e.g., a liquid level in a tank behaving as an integrator.

2. Choose output variables that must be kept within equipment and operating constraints (e.g., temperatures, pressures, compositions).

3. Select output variables that are a direct measure of product quality (e.g., composition, refractive index) or that strongly affect it (e.g., temperature or pressure).

4. Choose output variables that seriously interact with other controlled variables.
   - The pressure in a steam header that supplies steam to downstream units is a good example; if the supply pressure is not regulated, it will act as a significant disturbance to downstream units.

5. Choose output variables that have favourable dynamic and static characteristics.
   - Output variables that have large measurement time delays, large time constants, or are insensitive to MVs, are poor choices.
5.2 A systematic design procedure

5.2.2 Selection of variables

**Manipulated variables**

6. Select inputs that have large effects on controlled variables.
   - Ideally, an input should have a large effect on one output and a small effect on variables controlled by other inputs.
   - It is more effective to control an inventory (i.e., a level) by a large flow than by a small flow.

7. Choose inputs that rapidly affect the controlled variables.

8. The manipulated variables should affect the controlled variables directly, rather than indirectly.

9. Avoid recycling of disturbances.
   - It is preferable not to manipulate an inlet stream or a recycle stream; otherwise, disturbances tend to be propagated forward, or recycled back, to the process.

Note that the above guidelines may be in conflict. Then a trade-off has to be made in selecting a manipulated variable from a number of candidates.
5.2 A systematic design procedure

5.2.2 Selection of variables

Measured variables

10. Reliable, accurate measurements are essential for good control.
   – Inadequate measurements are a key factor in poor control performance.

11. Select measurement points that have an adequate degree of sensitivity.
   – If a product composition in a distillation column cannot be measured on-line, it is often controlled indirectly by regulating a tray temperature near the product end of the column. However, for high-purity separations, tray temperatures very close to the top of the column tend to be quite insensitive to composition changes making them bad choices for control.

12. Select measurement points that minimize time delays and time constants.
   – Smaller time delays and time constants improve closed-loop stability and response characteristics.
5.2.3 Degrees-of-freedom analysis

A necessary (but not sufficient) requirement for a model to have a unique solution is that
- the number of unknown variables, and
- the number of independent model equations, including control laws, are equal.

An equivalent statement is that all available degrees-of-freedom (DOF) must be utilized (by introducing controllers).

**DOF definitions**

The number of DOF is defined

$$N_f = N_v - N_e$$

- $N_v =$ number of variables in the model (time $t$ is not counted!)
- $N_e =$ number of equations in the model
5.2.3 Degrees-of-freedom analysis

DOF definitions

The number of control DOF is defined

\[ N_m = N_f - N_s = N_v - N_s - N_e \]

- \( N_m \) = number of variables that need to be manipulated
- \( N_s \) = number of externally specified variables, which include
  - disturbance variables
  - variables with fixed values (e.g., variables of feed streams)

If \( N_m' \) manipulated variables have been assigned for control, the remaining DOF is

\[ N_c = N_f - N_s - N_m' \]

Obviously, \( N_c = 0 \) when \( N_m' = N_m \).

The steady-state DOF is defined

\[ N_{ss} = N_m - N_{v0} \]

- \( N_{v0} \) = number of variables with no steady-state effect (i.e., they do not affect the steady-state solution); typical variables are
  - liquid levels in holdup tanks

Usually, the operational cost depends on the steady-state only. This means that the steady-state DOF indicates the degrees of freedom for process optimization.
5.2 A systematic design procedure  

5.2.3 Degrees-of-freedom analysis

**Calculation of \( N_f \)**

Calculation of the \( N_f \) DOF is illustrated by a flash drum example [from Steph]. The flash drum model is defined by the following equations. Certain assumptions are made.

**Total mass balance**

\[
A \rho \frac{d\dot{h}}{dt} = F_f - (F_v + F_L)
\]

**Component balances**

\[
A \rho \frac{d(hx_i)}{dt} = F_f z_i - (F_v y_i + F_L x_i) \quad i = 1, 2, \ldots, N - 1
\]

**Heat balance**

\[
c_{p,L} A \frac{d(hT)}{dt} = c_p F_f T_f - (c_p V F_v T + c_{p,L} F_L T) + U A_S(T_S - T)
\]

**Vapour-liquid equilibrium (VLE) relationships**

\[y_i = K_i(T, p) x_i \quad i = 1, 2, \ldots, N\]

**Consistency constraints**

\[
\sum_{i=1}^{N} x_i = 1 \quad \text{and} \quad \sum_{i=1}^{N} y_i = 1
\]

\[
N_v = [4] + [3(N - 1)] + [3] + [1] + [0] = 3N + 7
\]

\[
N_e = [1] + [N - 1] + [1] + [N] + [2] = 2N + 3 \quad \Rightarrow \quad N_f = N + 4
\]
5.2 A systematic design procedure

### Calculation of $N_m$ and $N_{ss}$

Calculation of the control DOF $N_m$ and the steady-state DOF $N_{ss}$ is illustrated by continuation of the flash drum example [from Steph].

Assume that the feed properties $T_f$ and $z_i, i = 1, ..., N - 1$ are given ($p_f$ is not consider because it was not counted as a variable). Then,

\[
N_s = 1 + N - 1 = N
\]

\[
N_m = N_f - N_s = N + 4 - N = 4
\]

This means that 4 variables can be controlled by 4 manipulated variables.

- Variables to be controlled: $F_f, T, p, h$
- Manipulated variables (control valves): $F_f, W_S, F_V, F_L$

The variable $h$ does not have a steady-state effect. Thus,

\[
N_{v0} = 1
\]

\[
N_{ss} = N_m - N_{v0} = 4 - 1 = 3
\]

This means that optimization can be done with respect 3 variables, e.g., $F_f, T$ and $p$. $T$ and $p$ will affect the VLE and thus the product compositions.
5.2 A systematic design procedure

5.2.3 Degrees-of-freedom analysis

**Easier calculation of** $N_m$ **and** $N_{ss}$

The calculation of $N_m$ according to

\[ N_m = N_f - N_s = N_v - N_s - N_e \]

is rather complicated and can easily lead to errors (because of mistakes!).

For most practical control problems, the following **general rule** applies:

- The control DOF $N_m$ is equal to the number of independent input variables that can be controlled.

**Example** (Skogestad, 2004)

Assuming no externally specified variables, i.e.,

$N_s = 0$, we have

$N_m = 11$ (control valves)

$N_{v0} = 4$ (liquid levels)

$N_{ss} = 7$