Part 5: inventory control +
Example: Level control

MV1 = z0 (inflow valve position)
MV2 = z1 (outflow valve position) (likely to saturate)
CV1 = F0 (inflow): Should be controlled at setpoint F0,s (if possible)
CV2 = level: must always be controlled (at some SP)

Nominal design (follow “pair-close” rule)

Problem: outflow-valve may saturate at fully open (z1=1) and then we lose level control

Note: We did not following the “input saturation rule” which says:
Pair MV that may saturate (z1) with CV that can be given up (F0)
Disturbance

Reverse pairing (follows “input saturation rule”):

This gives simple MV-CV switching (if z2 saturates at fully open)

BUT with Reverse pairing: Get “long loop” for F0
In addition: loose control of y2= level if z0 (F0-valve) saturates

«Long loop» = Works through other loops
Alternative solution: **Follow “Pair close”-rule** and use **Complex MV-CV switching**. When z1 saturates at max, use the other MV (z0) for level control and give up controlling F0. Get: “Bidirectional inventory control”

Three options for MV-MV switching:
1. SRC (problem since F0s varies)
2. Two controllers
3. VPC (“Long loop” for z1, backoff)

- Avoid long loop for control of F0
- Works both when F0-valve or F1-valve saturate at open
Overall: seems to be the best solution
Alt. 3. Valve position control on $z_1$

VPC: “reduce inflow ($F_0$) if outflow valve ($z_1$) approaches fully open”
Alt. 2: Two controllers (recommended)

In addition: Use of two setpoints is good for using buffer dynamically!!

SP-L = low level setpoint
SP-H = high level setpoint
Inventory control for units in series and TPM

• **TPM** ("gas pedal") = *Variable used for setting the throughput/production rate (for the entire process)*.

• Where is the TPM located for the process?
  • Usually at the feed, but not always!
  • Important for dynamics
  • Determines the inventory control structure

• **Rule (Price et al., 1994):** *Inventory control (Level and pressure) must be radiating around TPM:*
Inventory control for units in series

**Radiating rule:**
Inventory control should be “radiating” around a given flow (TPM).

Follows radiation rule

Does NOT follow radiation rule
Generalization of bidirectional inventory control

Reconfigures automatically with optimal buffer management!!

Maximize throughput:
\[ F_s = \infty \]

![Diagram of bidirectional inventory control](image)

F. G. Shinskey, «Controlling multivariable processes», ISA, 1981, Ch.3

Cristina Zotica, Krister Forsman, Sigurd Skogestad, «Bidirectional inventory control with optimal use of intermediate storage», Computers and chemical engineering, 2022
Fig. 13. Simulation of a temporary (19 min) bottleneck in flowrate $F_1$ for the proposed control structure in Fig. 10. The TPM is initially at the product ($F_5$).
Challenge: Can MPC be made to do his? Optimally reconfigure loops and find optimal buffer?

YES. Use «trick»/insight of unachievable high setpoints on all flows
Extension. **Bidirectional inventory control with minimum flow for** $F_2$

Max flow: $F_S = \infty$

$L = 10\%,$
$M_L = 40\%,$
$M_H = 60\%$
$H = 90\%.$

Fig. 37. Bidirectional inventory control scheme for maximizing throughput (dashed black lines) while attempting to satisfy minimum flow constraint on $F_2$ (red lines). $H$, $L$, $M_L$, and $M_H$ are inventory setpoints.

The control structure in Fig. 37 may easily be dismissed as being too complicated so MPC should be used instead. At first this seems reasonable, but a closer analysis shows that MPC may not be able to solve the problem (Bernardino & Skogestad, 2023). Besides, is the control structure in Fig. 37 really that complicated? Of course, it is a matter of how much time one is willing to put into understanding and studying such structures. Traditionally, people in academia have dismissed almost any industrial structure with selectors to be ad hoc and difficult to understand, but this view should be challenged.
Fig. 38. Bidirectional inventory control structure for industrial plant with on/off (1/0) control of filtration unit.

$H$, $L$, and $M$ are inventory setpoints with typical values 90\%, 10\% and 50\%.
If it is desirable to set a flowrate ($F_3$) somewhere in the system, then flow controllers must be added at this location.
Inventory control (level, pressure)

- All inventories (level, pressure) must be regulated by
  - Controller, or
  - “self-regulated” (e.g., overflow for level, open valve for pressure)
  - Exception closed system: Must leave one inventory (level) uncontrolled

- Usually only one TPM
  - To get consistent mass balance: Can only fix same flow once
  - But there are exceptions
    - Multiple feeds (they are then usually set in ratio to the “main” TPM)
    - Recycle systems often have a flow that can be set freely

- Rule for maximizing production for cases where we cannot rearrange inventory loops: Locate TPM at expected bottleneck
  - Otherwise you will need a “long loop” and you get loss in production because of backoff from constraint
Consider a gas pipeline with two valves. We have measurements of the inflow $F_1$ and the intermediate pressure $p$ and these should be controlled. The volume of the pipeline can be represented as a tank with volume $V$ as shown in the figure above.

Steady-state data: $F_1=1 \text{ kg/s}$, $z_1=z_2=0.5$, $p_1=2 \text{ bar}$, $p=1.88 \text{ bar}$, $p_2=1.8 \text{ bar}$, $V=130 \text{ m}^3$, $T=300 \text{ K}$, Parameters: $R=8.31 \text{ J/K.mol}$, $M_w=18\times 3 \text{ kg/mol}$ (so the gas is steam).

The following model equations are suggested to describe the system.

1. $\frac{dm}{dt} = F_1 - F_2$
2. $m = k_p \cdot p$ where $k_p=VM_w/(RT)$
3. $F_1 = C_1z_1\sqrt{p_1 - p}$
4. $F_2 = C_2z_2\sqrt{p - p_2}$

(a) Suggest a control structure
(b) What if we want to control $p_2$ instead of $p$?
(a) The «obvious» pair-close pairing os OK. However, interactions between loops may be severe. Suggest tuning the FC first, and the PC about 5 times slower.
(b) \textbf{NO!} 

Not consistent

Disturbances in $F_1$, $p_2$, pend

Or: $z_2$ to fully open (lose control of $p_2$)

$F_3 = c_3(p_2 - p_{end})^{1/2}$

$z_1$ to fully open (lose control of $F_1$)

Opposite Disturbances

Not consistent
Distillation example.

Can be given up

CV1=x_B (cannot be given up)
MV1=V (can reach max)
MV2=L (normally used for x_D)
MV3=F (has setpoint)

Spec. is max 5% heavy in top, but heavy in top gives economic loss. Economic optimal over-purification of top product (x_Dopt) depends on prices p

Bottom product is the most valuable and should always be kept at its purity constraint (x_Bmin).

Split range control*:
1. Normally we control x_B with V
2. If V reaches V_max then we instead use reflux L (and give up x_Dopt)
3. If L is not available (because it’s used to control x_Dmin) then we instead use the feedrate F to control x_B

*Split range control can be replaced by three controllers with different setpoints for x_B

This is an example where MPC may be preferred
Alternative solution with different setpoints

This solution looks simpler, but it is not as good dynamically in cases where we need to limit feed F to the column. We then use F to control top composition, and L to control bottom composition. The reverse pairing is better (which is what we get with the other solution)
Simulation of alternative solution. The problematic pairing is used toward the end (t>80), but it’s not really tested because there are no disturbances
Important insight

• Many problems: Optimal steady-state solution always at constraints

• In this case optimization layer may not be needed
  • if we can identify the active constraints and control them using selectors
Here is a summary of some additional insights from this paper:

- If the industrial solution has a selector (sometimes realized using a saturation element, especially for the cascade implementation) then generally there is a CV constraint involved. Most likely, the selector is performing a steady-state CV-CV switch (E4), although there may be exceptions as seen in the cross-limiting example below.

  - A CV-CV switch can be realized in two ways, either with two (or more) independent controllers with a selector on the MV (Fig. 17), or as a cascade implementation with a selector on the CV setpoint (Fig. 19).

  - If there are several selectors (max and min) in series then we know that the constraints are potentially conflicting and that the highest priority constraint should be at the end (Fig. 18).

- For MV-MV switching there are three alternatives.

  1. A common solution is split range control (E5; Fig. 21) which is usually easy to identify.
  2. Another common solution is multiple controllers with different setpoints (E6; Fig. 23). It may be a bit more difficult to identify.
  3. Finally, there is VPC (E7), as just discussed, which is probably the least common solution for MV-MV switching.

One should have all these three alternatives in mind when choosing the best solution for MV-MV switching, as there is not one alternative which is best for all problems (see Section 5.1 for details).

- If the industrial solution has a valve position controller (VPC) then there may be two quite different problems that it is addressing (see E3 and E7 in Table 1), and it may not be immediately clear which.

  1. If we have an extra MV for dynamic reasons (E3; Fig. 12) then the two controllers (and MVs) are used all the time. The MV manipulated by the VPC (MV$_1$ in Fig. 12) is then used on the longer time scale, whereas the MV linked to the CV (MV$_2$ in Fig. 12) is used for dynamic reasons (fast control). Here, an alternative is to use parallel control (Fig. 13).
  2. There is also another possibility, namely, when the VPC makes use of an extra MV to avoid that the primary MV saturates at steady-state (E7; Fig. 24). This is then a case where the VPC is used for MV-MV switching and the VPC is only active part of the time.
Challenges selector design

• Standard approach requires pairing of each active constraint with a single input
  • May not be possible in complex cases

• Stability analysis of switched systems is still an open problem
  • Undesired switching may be avoided in many ways:
    • Filtering of measurement
    • Tuning of anti-windup scheme
    • Minimum time between switching
    • Minimum input change
Implementing optimal operation

**Summary**

- **Most people think**
  - You need a detailed nonlinear model and an on-line optimizer (RTO) if you want to optimize the process
  - You need a dynamic model and model predictive control (MPC) if you want to handle constraints
  - The alternative is Machine Learning

- **No! In many cases you just need to measure the constraints and use PID control**
  - «Conventional advanced regulatory control (ARC)»

- **How can this be possible?**
  - Because optimal operation is usually at constraints
  - Feedback with PID-controllers can be used to identify and control the active constraints
  - For unconstrained degrees of freedom, one often have «self-optimizing» variables

- **This fact is not well known, even to control professors**
  - Because most ARC-applications are *ad hoc*
  - Few systematic design methods exists

- **Today ARC and MPC are in parallel universes**
  - Both are needed in the control engineer's toolbox