

# Supervisory control

Decomposition: vertical and horizontal

TPM

Procedure

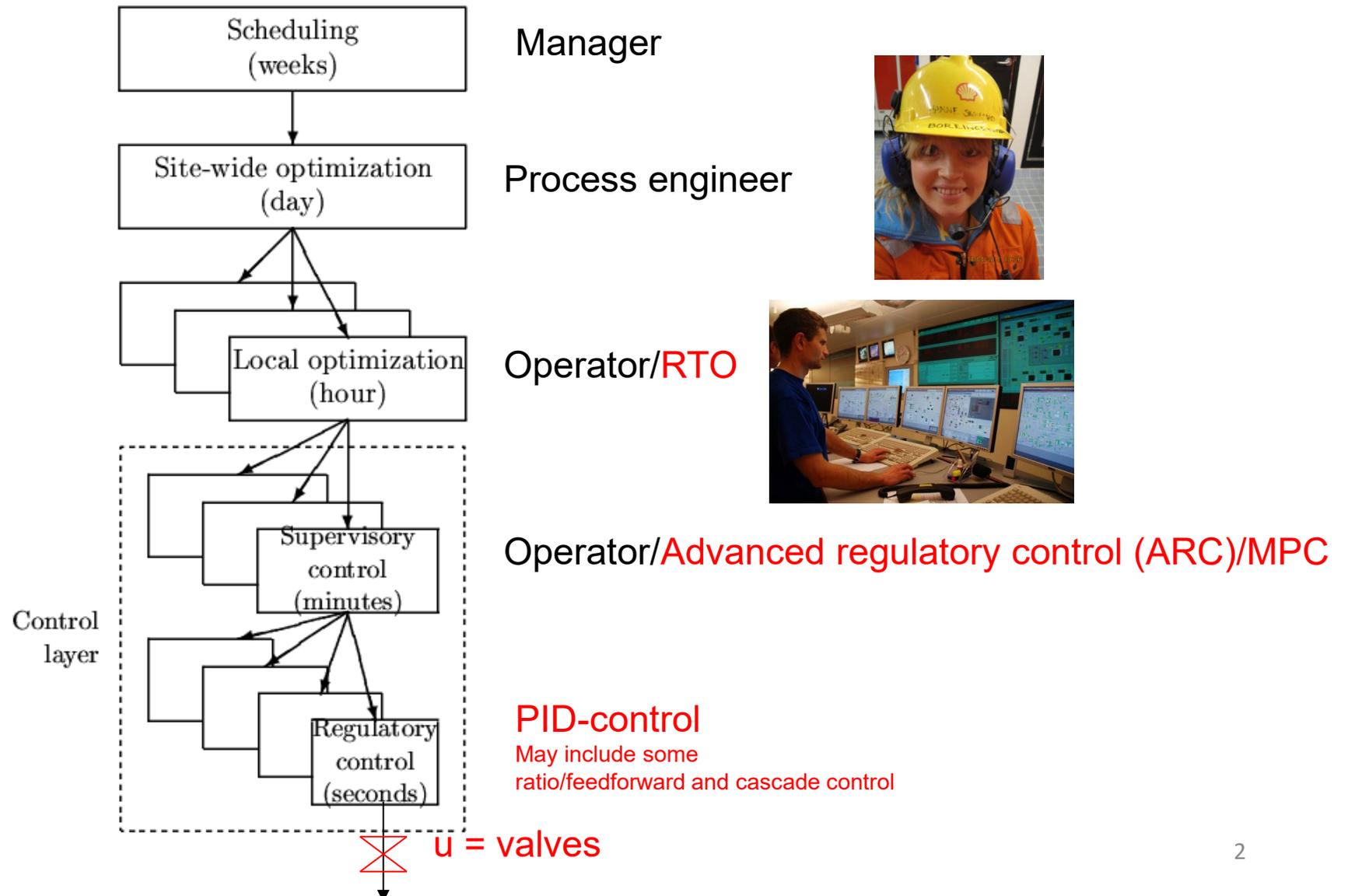
# Practical operation: Hierarchical (cascade) structure based on time scale separation

NOTE: Control system is decomposed both

- Hierarchically (in time)
- Horizontally (in space)

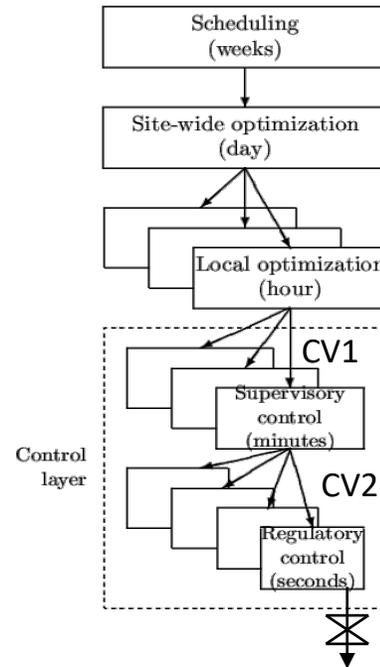
Status industry:

- **RTO** is rarely used.
- **MPC** is used in the petrochemical and refining industry, but in general it is much less common than was expected when MPC «took off» around 1990
- ARC is common
- Manual control still common...



# Two fundamental ways of decomposing the controller

- Vertical (hierarchical; cascade)
- Based on time scale separation
- Decision: Selection of CVs that connect layers



- Horizontal (decentralized)
- Usually based on distance
- Decision: Pairing of MVs and CVs within layers

CV = controlled variable  
MV = manipulated variable

Combine control and optimization into one layer?

## EMPC: Economic model predictive “control”

$$J_{EMPC} = J + J_{control}$$

Penalize input usage,  $J_{control} = \sum \Delta u_i^2$

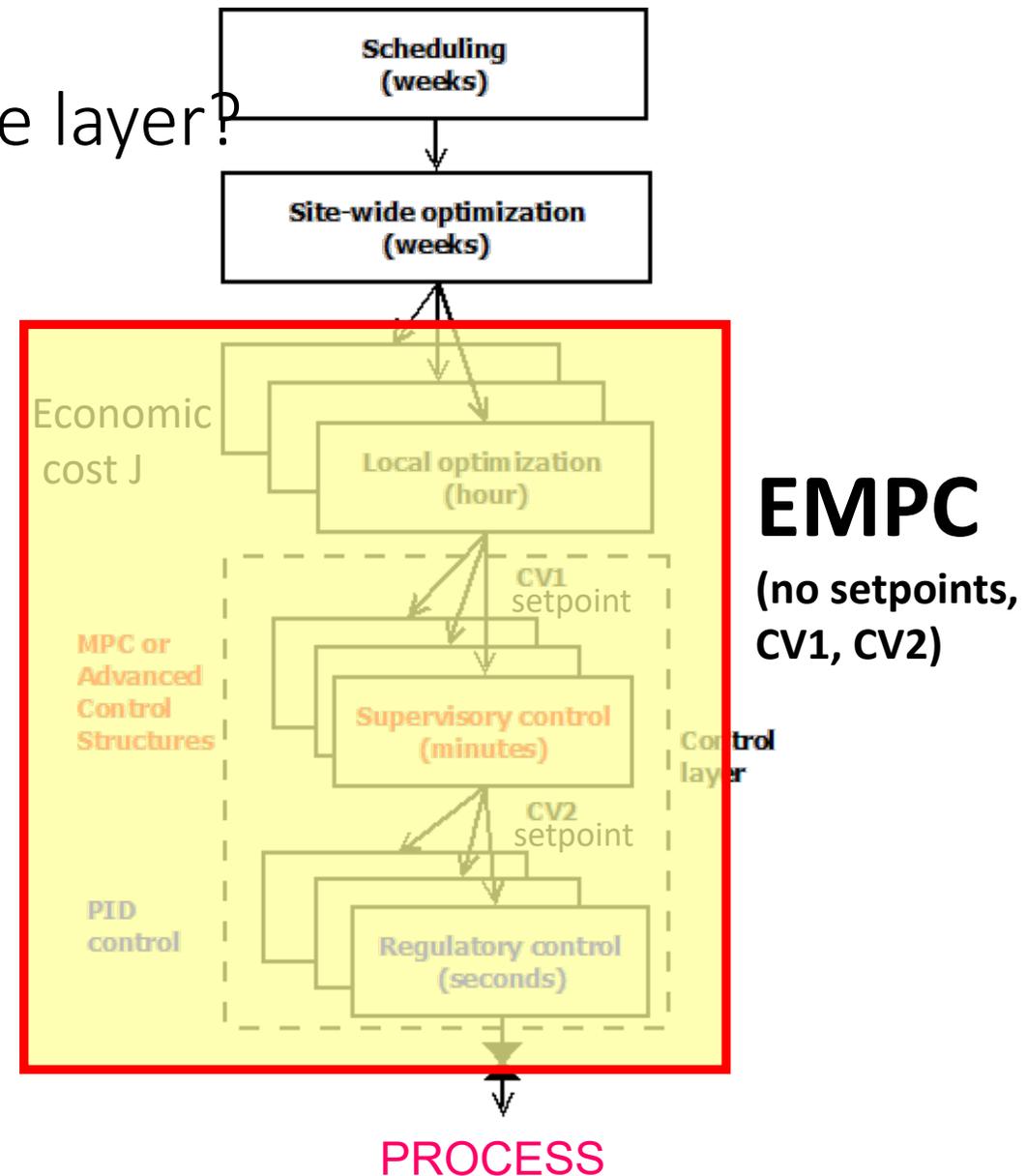
**NO, combining layers is generally not a good idea!  
(the good idea is to separate them!)**

One layer (EMPC) is optimal theoretically, but

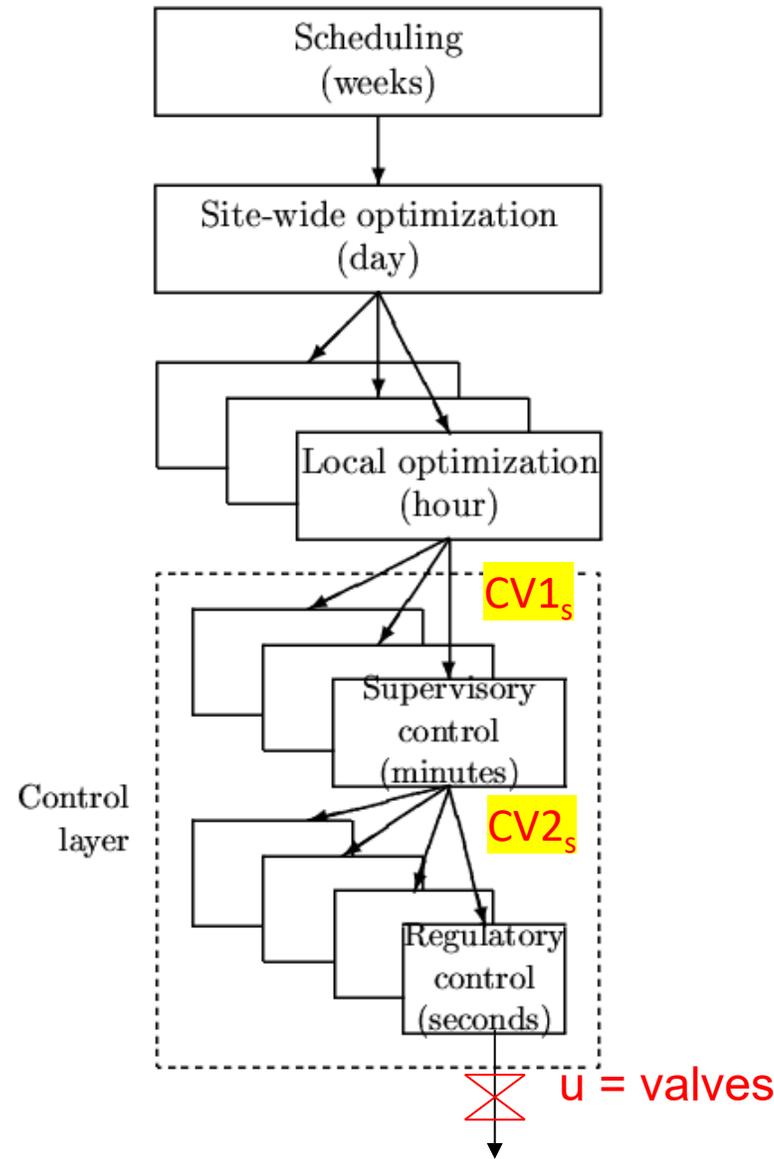
- Need detailed dynamic model of everything
- Tuning difficult and indirect
- Slow! (or at least difficult to speed up parts of the control)
- Robustness poor
- Implementation and maintenance costly and time consuming

Typical economic cost function:

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



# What is the difference between optimization and control?



My definition:

Optimization:

- Minimizes economic cost

Control:

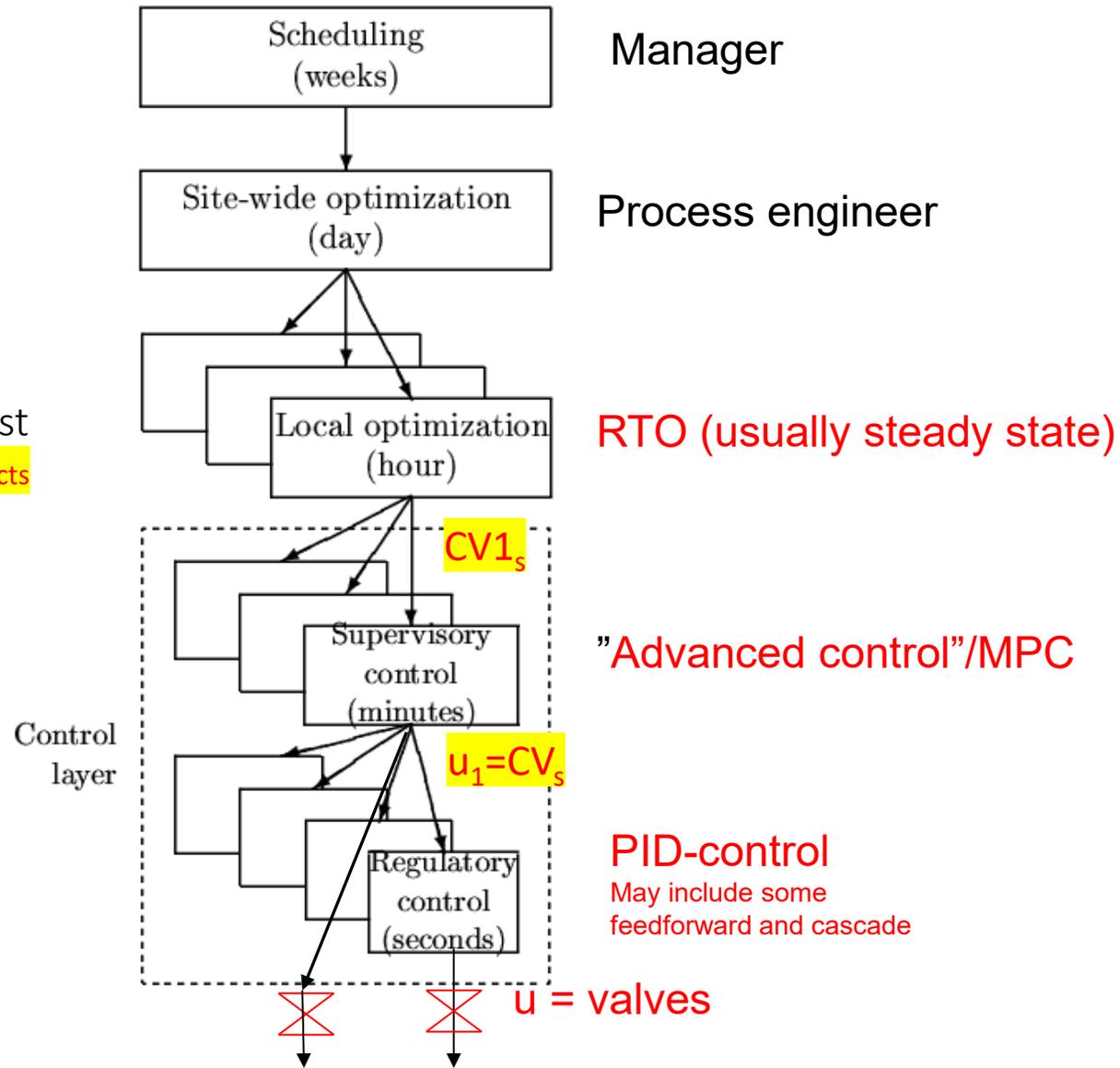
- Follow **setpoints  $y_s$**

# Cost functions in layers

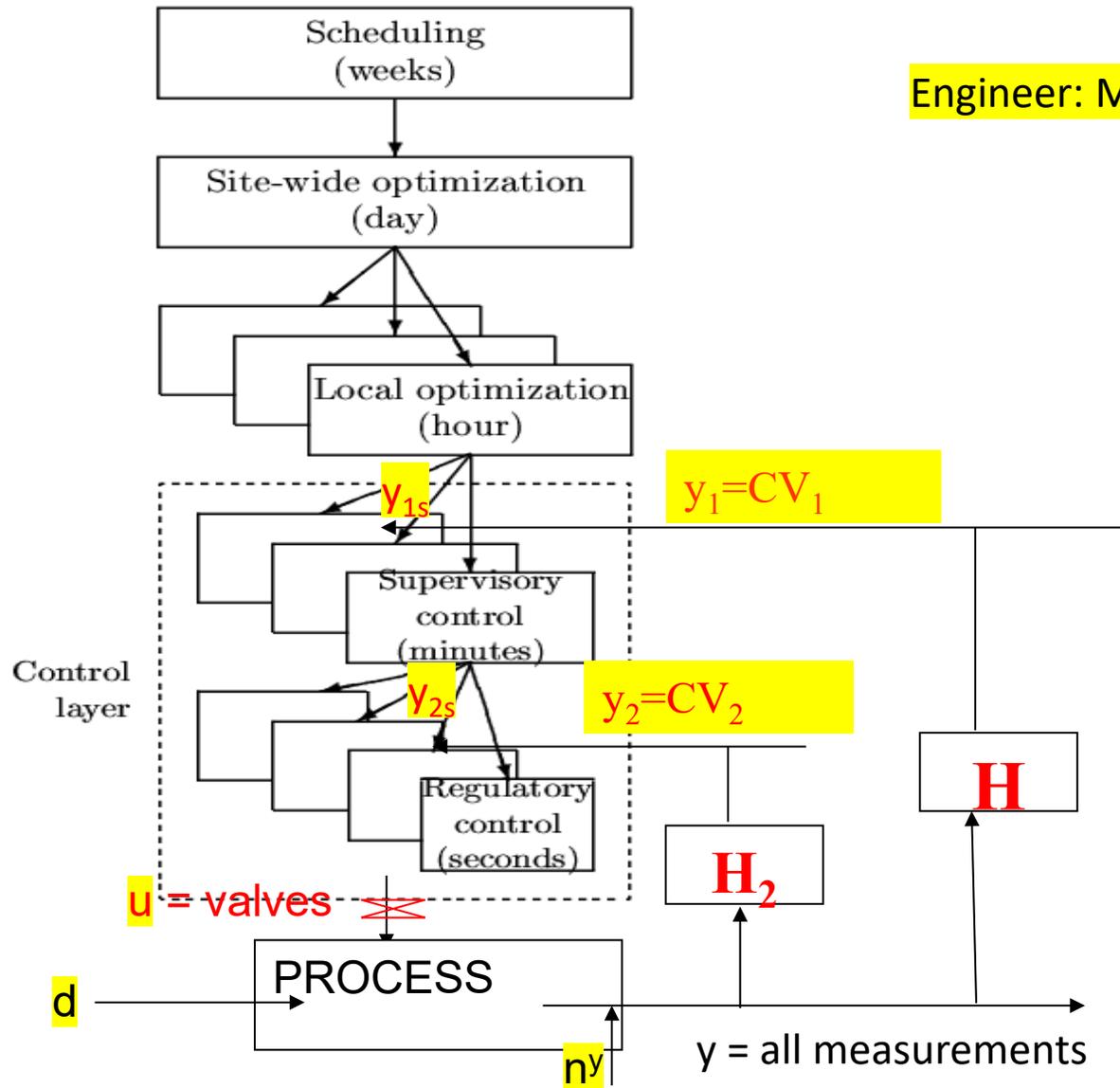
RTO: Minimize economic cost  
 $J_{\xi} = \text{cost feed} + \text{cost energy} - \text{value products}$

Setpoint control  
 $J_{c1} = Q(y_1 - y_{1s})^2 + R\Delta u_1^2$  (MPC)  
 (+ look after other variables,  
 Avoid constraints)

PID: Stabilize + avoid drift  
 $J_{c2} = Q(y_2 - y_{2s})^2 + R\Delta u^2$   
 + look at Gain margin...  
 (or just use SIMC-rules!)



# Use of data in layers (feedback)



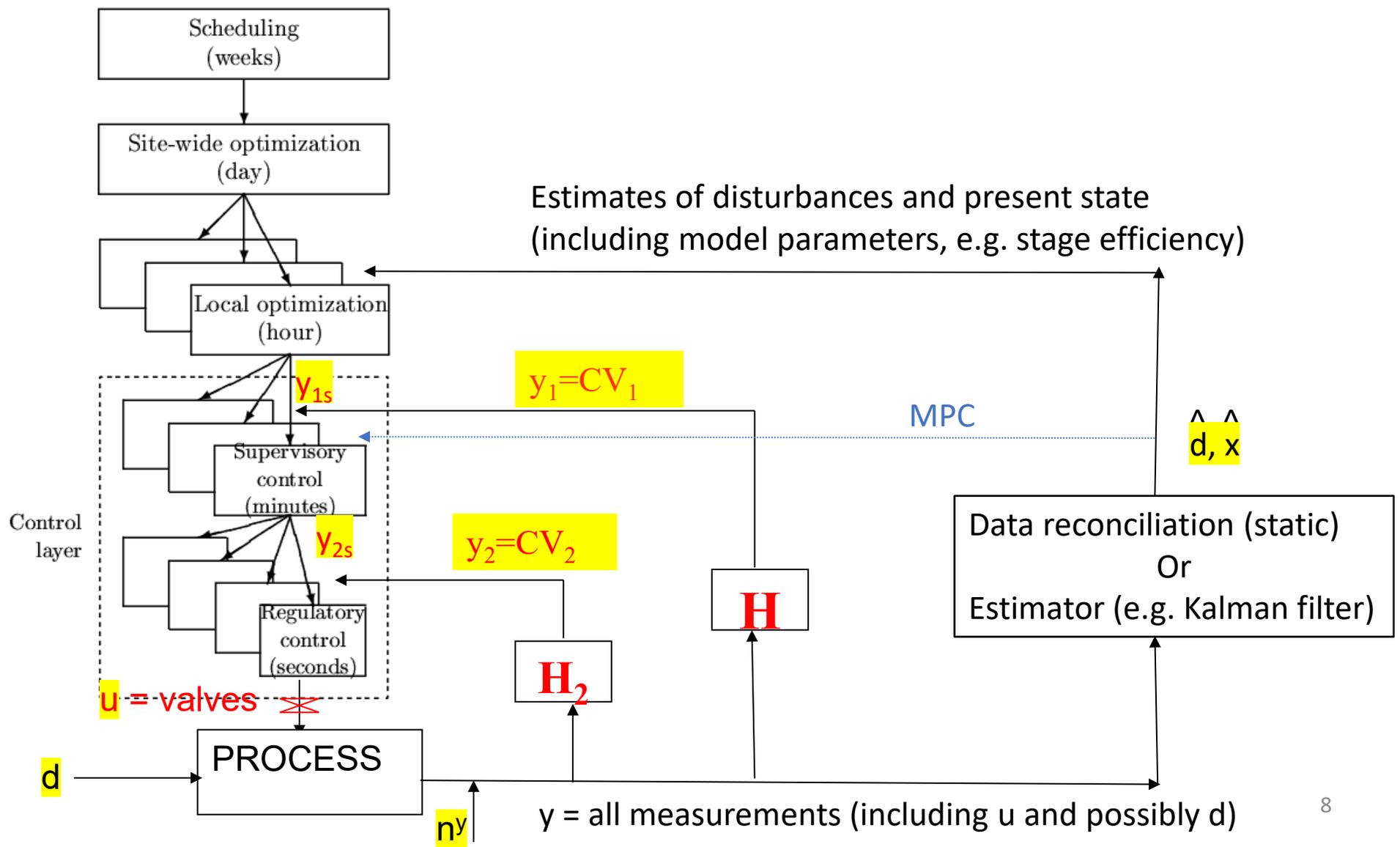
Engineer: Must choose what to control ( $H$  and  $H_2$ )

- $H$  and  $H_2$  are usually selection matrices

Typically:

- $y_1 = Hy = \text{active constraints} + \text{«self-optimizing» variables}$
- $y_2 = H_2y = \text{drifting variables (levels, pressures, temperatures)}$

# Optimization layer: Needs model parameters and disturbances

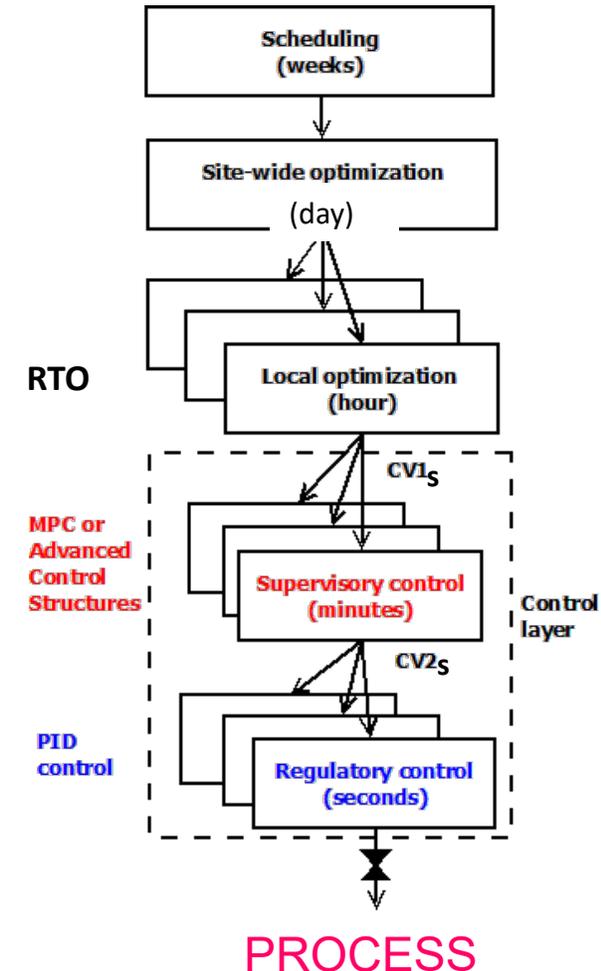




# Is there a problem with model consistency between layers?

Quote from a recent paper I reviewed

- “One of the difficulties in practical implementations of classic Real-Time Optimization (RTO) strategy is the integration between optimization (RTO) and control layers (MPC), mainly due to the differences between the models used in each layer, which may result in unreachable setpoints coming from optimization to the control layer. In this context, Economic Model Predictive Control (EMPC) is a strategy where optimization and control problems are solved simultaneously.”
- Is this likely to happen?
- **No, This is a myth and no reason for choosing EMPC**
- Truth: With integral action in the control layer (MPC), the process will go to the setpoints ( $y_{1s}=CV1_s$ ) desired by the RTO layer, irrespective of any model error in the MPC layer
  - $J_{MPC} = Q(y_1 - y_{1s})^2 + R\Delta u_1^2$
- Of course, the setpoints from the RTO layer must correspond to a feasible steady state, but the model in the MPC layer does not affect this
- Of course, there may be economic losses dynamically, for example, dynamic constraints may mean it takes some time to reach the setpoints



# Main objectives operation

**1. Economics:** Implementation of acceptable (near-optimal) operation

**2. Regulation:** Stable operation around given setpoint

ARE THESE OBJECTIVES CONFLICTING?

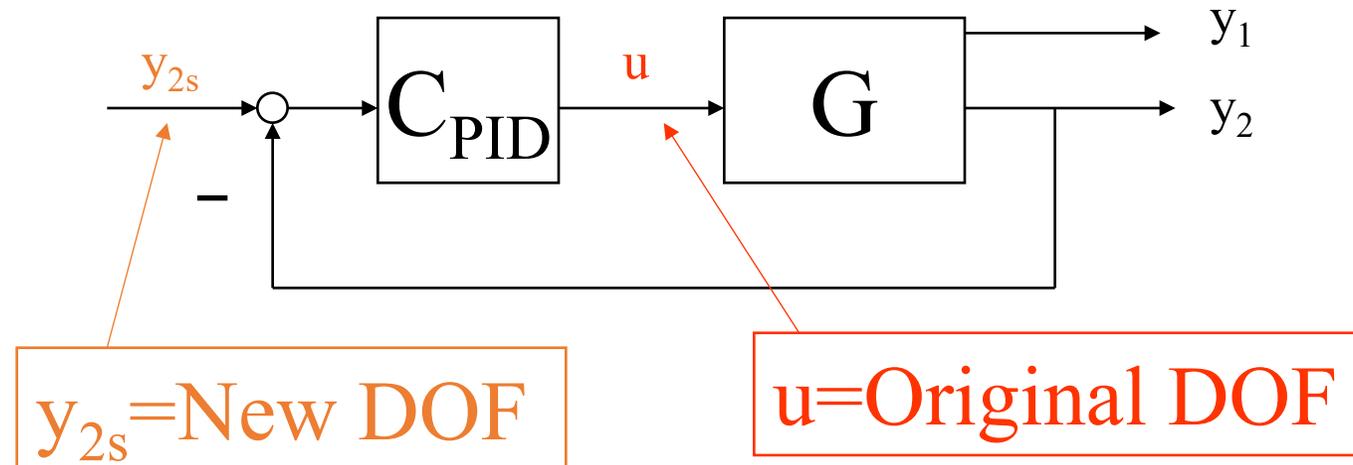
IS THERE ANY LOSS IN ECONOMICS?

- Usually NOT
  - Different time scales
    - Stabilization fast time scale
  - Stabilization doesn't "use up" any degrees of freedom
    - Reference value (setpoint) available for layer above
    - But it "uses up" part of the time window

# Hierarchical structure: Degrees of freedom unchanged

- No degrees of freedom lost as setpoints  $y_{2s}$  replace inputs  $u$  as new degrees of freedom for control of  $y_1$

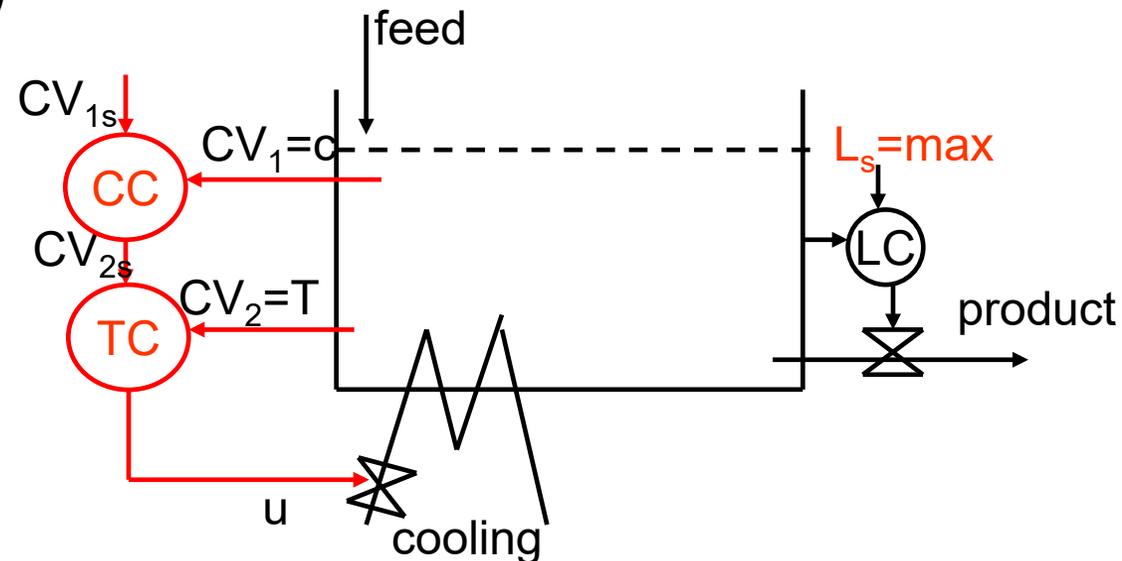
Cascade control:



# Example: Exothermic reactor (unstable)

Want to control  $y_1=CV1=\text{composition (+ level)}$

- $u = \text{cooling flow (F)}$
- $CV_1 = \text{composition (c)}$
- $CV_2 = \text{temperature (T)}$



# Systematic procedure for economic process control

## Start “top-down” with economics (steady state):

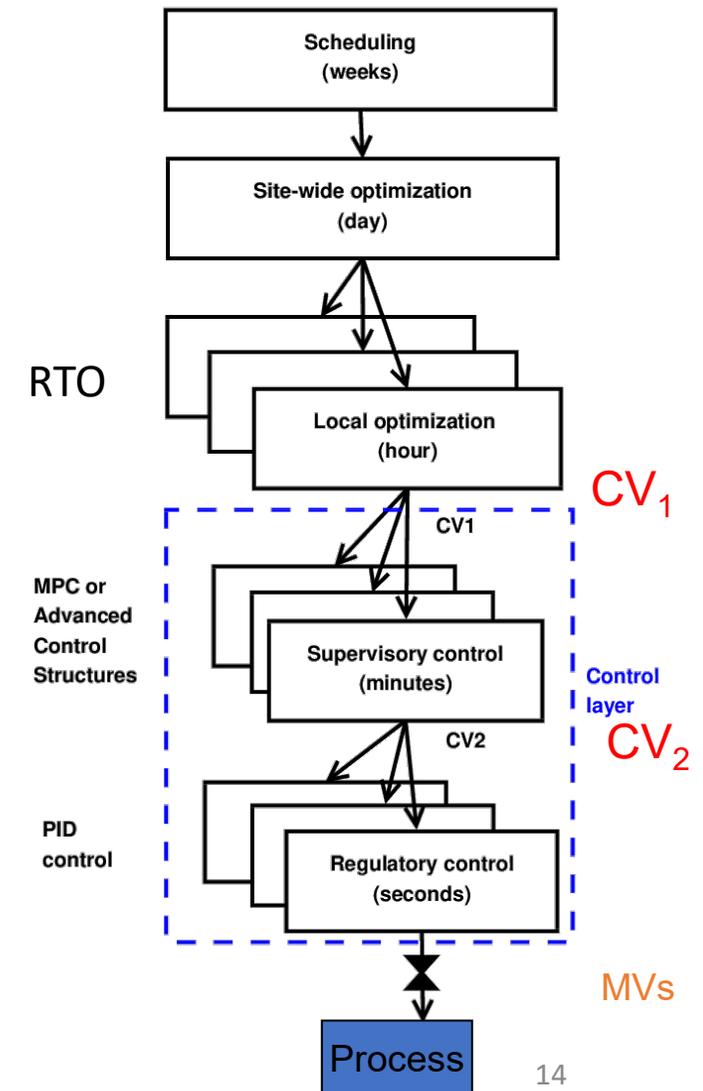
- Step 1: Define operational objectives (J) and constraints
- Step 2: Optimize steady-state operation
- Step 3: Decide what to control (CVs)
  - Step 3A: Identify active constraints = primary CV1.
  - Step 3B: Remaining unconstrained DOFs: Self-optimizing CV1 (find H)
- Step 4: Where do we set the throughput? TPM location

## Then bottom-up design of control system (dynamics):

- Step 5: Regulatory control
  - Control variables to stop “drift” (sensitive temperatures, pressures, ....)
  - Inventory control radiating around TPM

## Finally: Make link between “top-down” and “bottom up”

- **Step 6: “Advanced/supervisory control”**
  - Control economic CVs: Active constraints and self-optimizing variables
  - Look after variables in regulatory layer below (e.g., avoid saturation)
- **Step 7: Real-time optimization (Do we need it?)**



# Step 3: Sigurd's rules for CV selection

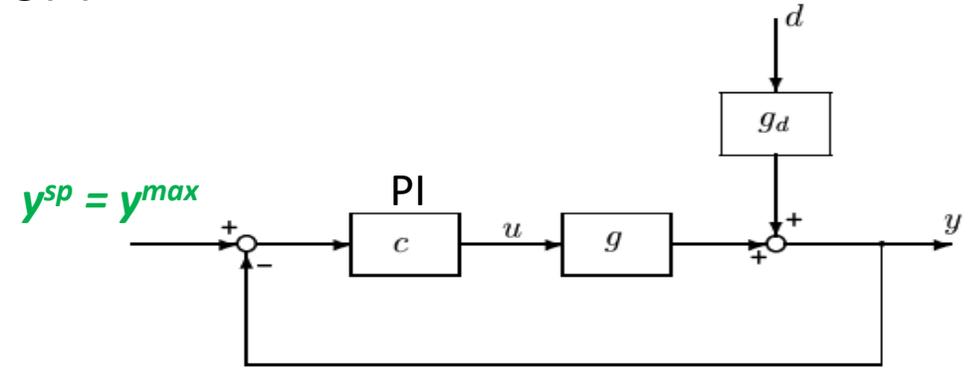
1. Always control active constraints! (almost always)
2. Purity constraint on expensive product always active (no overpurification):
  - (a) "Avoid product give away" (e.g., sell water as expensive product)
  - (b) Save energy (costs energy to overpurify)

## Unconstrained optimum:

3. Look for "self-optimizing" variables. They should
  - Be sensitive to the MV
  - have close-to-constant optimal value
4. **NEVER try to control a variable that reaches max or min at the optimum**
  - In particular, never try to control directly the cost  $J$
  - Assume we want to minimize  $J$  (e.g.,  $J = V = \text{energy}$ ) - and we make the stupid choice of selecting  $CV = V = J$ 
    - Then setting  $J < J_{\min}$ : Gives infeasible operation (cannot meet constraints)
    - and setting  $J > J_{\min}$ : Forces us to be nonoptimal (which may require strange operation)

# Cruise control: Optimization with PI-controller

$$\begin{aligned} &\max y \\ &\text{s.t. } y \leq y^{max} \\ &\quad u \leq u^{max} \end{aligned}$$



**Example: Drive as fast as possible to airport ( $u$ =power,  $y$ =speed,  $y^{max} = 110$  km/h)**

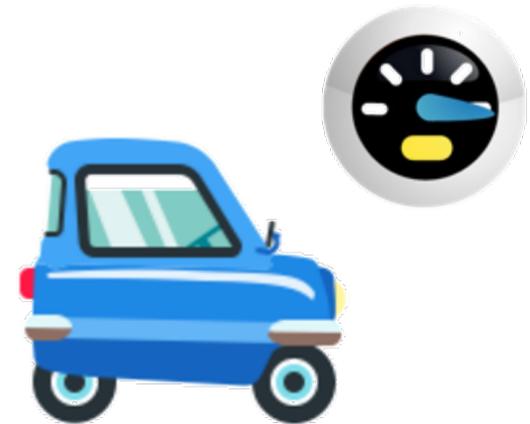
• **Optimal solution has two active constraint regions:**

1.  $y = y^{max}$  → speed limit
2.  $u = u^{max}$  → max power

• Note: Positive gain from MV ( $u$ ) to CV ( $y$ )

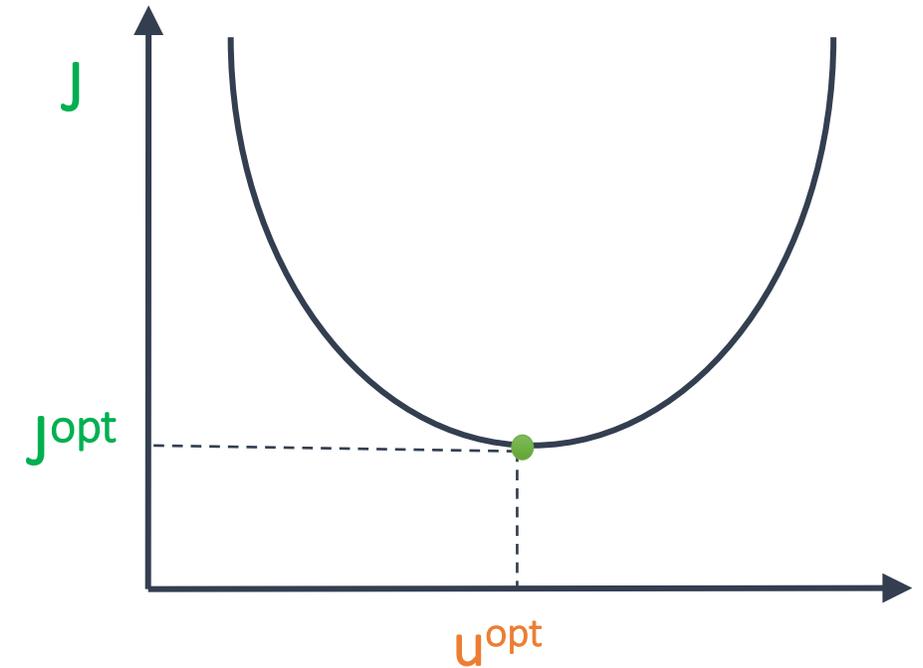
• **Solved with PI-controller**

- $y^{sp} = y^{max}$
- Anti-windup: I-action is off when  $u = u^{max}$



# The less obvious case: Unconstrained optimum

- $u$  = unconstrained MV
- What to control?  $y=CV=?$



Analytical solution:

$$H = Gy^T (YY^T)^{-1} \text{ where } Y = [FW_d \quad W_{ny}]$$

# Step 4: Inventory control and TPM

## Economic Plantwide Control of the Ethyl Benzene Process

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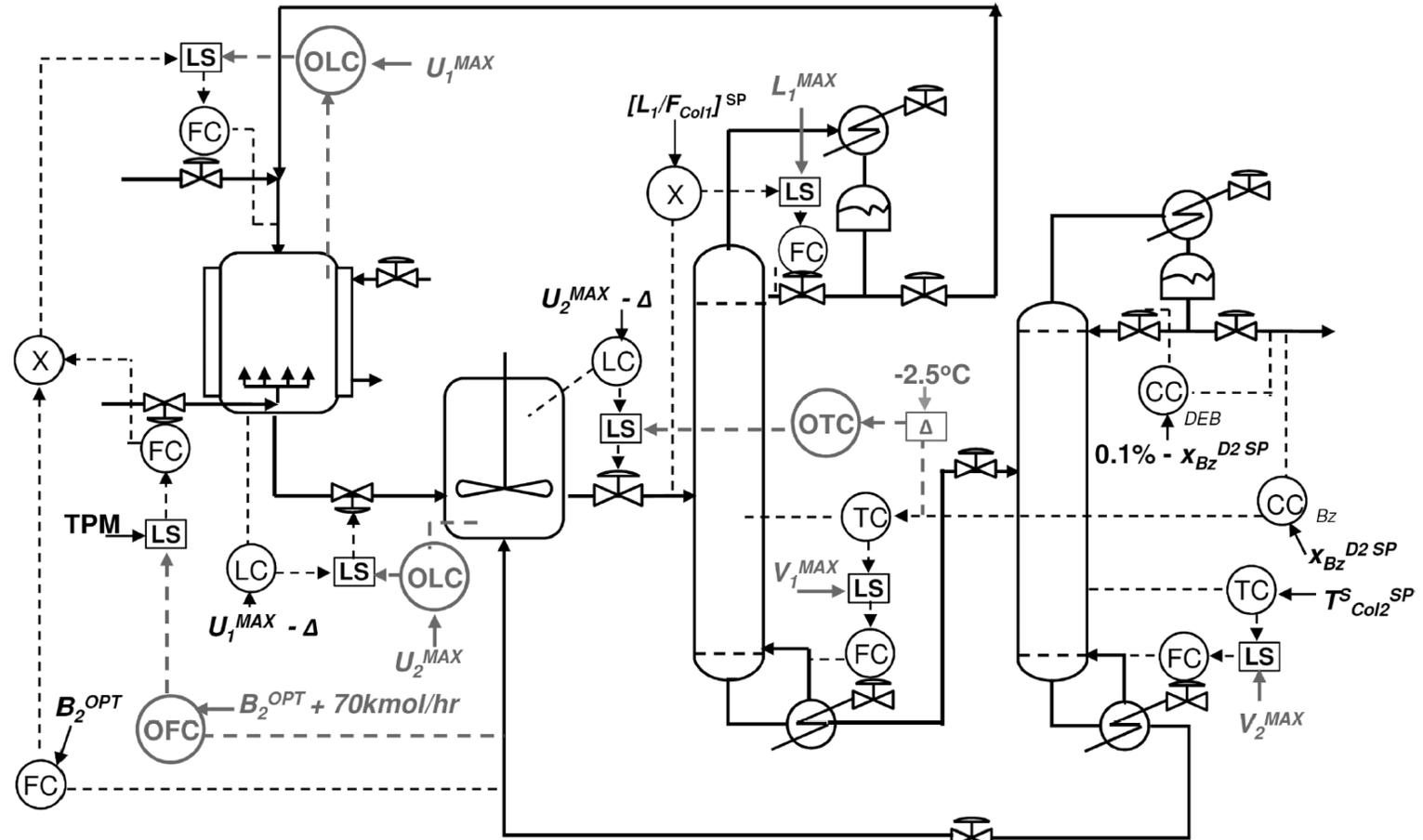
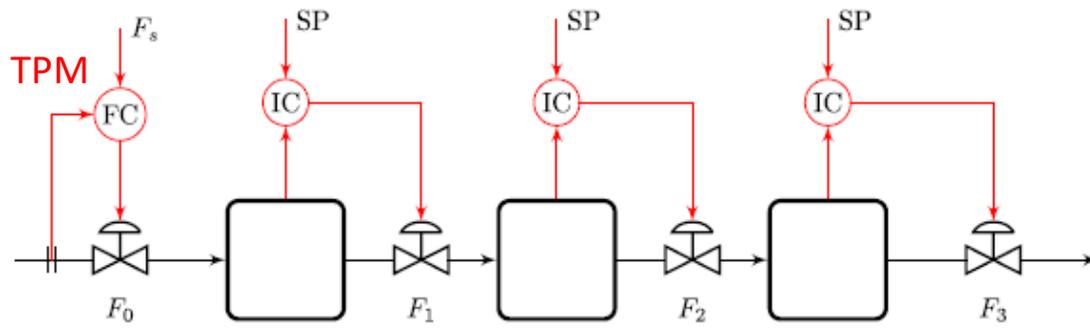
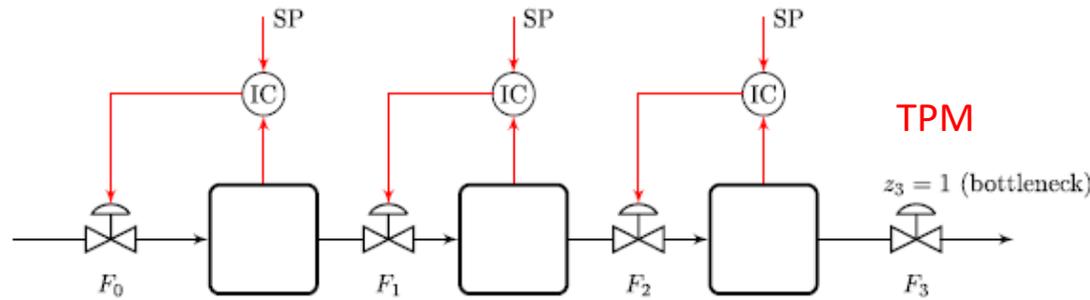


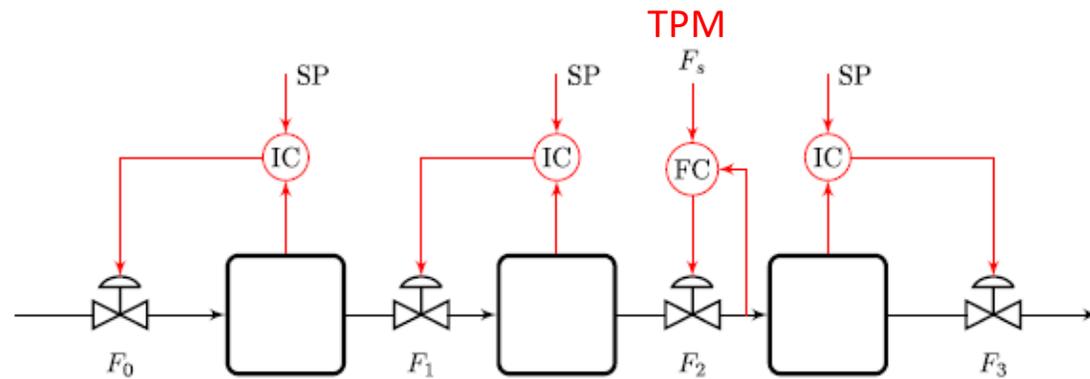
Figure 7. CS2 with overrides for handling equipment capacity constraints.



(a) Inventory control in direction of flow (for given feed flow,  $TPM = F_0$ )



(b) Inventory control in opposite direction of flow (for given product flow,  $TPM = F_3$ )



(c) Radiating inventory control for TPM in the middle of the process (shown for  $TPM = F_2$ )

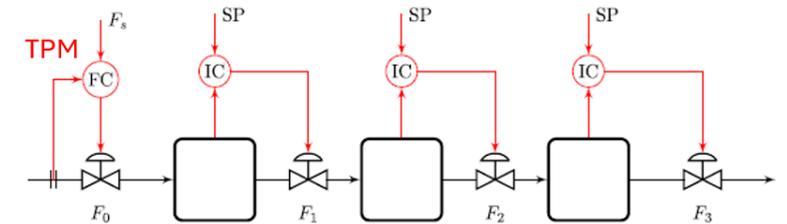
**Radiation rule** Inventory control should be “radiating” around a given flow (TPM),

*TPM = Gas Pedal = Variable used for setting the throughput/production rate (for the entire process).*

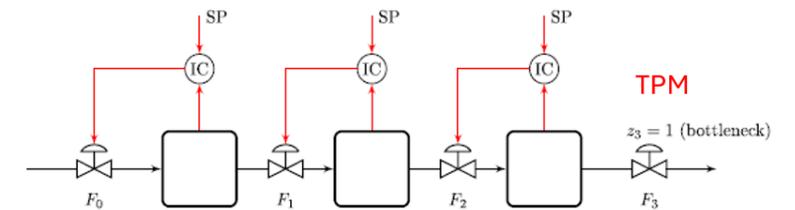
# Rules for inventory control

## Rules for inventory control

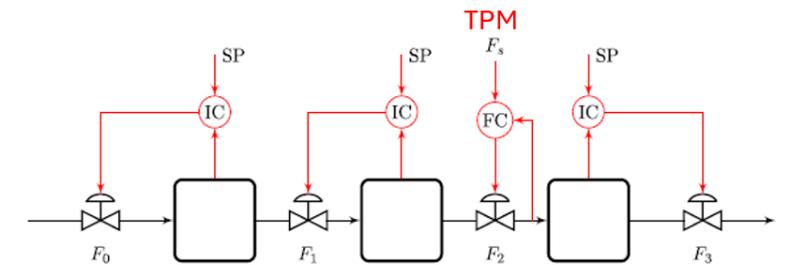
- **Rule 1.** Cannot control (set the flowrate) the same flow twice
- **Rule 2.** Controlling inlet or outlet pressure indirectly sets the flow (indirectly makes it a TPM)
- **Rule 3.** Follow the radiation rule whenever possible
- **Radiation rule** (actually more a strong recommendation): Inventory control should be “radiating” around a given flow (TPM), that is, it should be in the direction of flow downstream the TPM and it should oppose the direction of flow upstream the TPM.
  - Ref: (Aske & Skogestad, 2009; Buckley, 1964; Price et al., 1994)
- Breaking the radiation rule results in a “long loop”, that is, a control loop that only works when other loops are closed
- **Rule 4** (which should never be broken): No inventory loop should cross the location of the TPM
  - Ref: Not sure, but I have seen it stated



(a) Inventory control in direction of flow (for given feed flow, TPM =  $F_0$ )



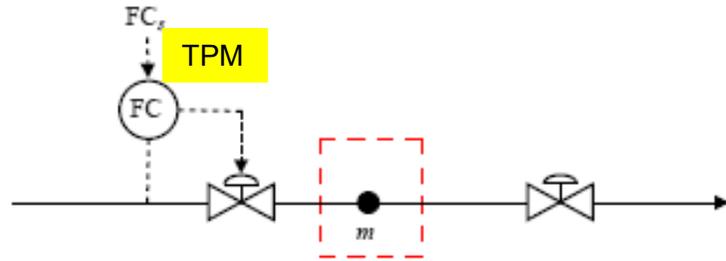
(b) Inventory control in opposite direction of flow (for given product flow, TPM =  $F_3$ )



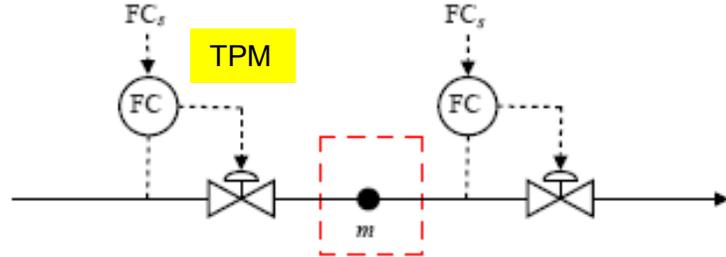
(c) Radiating inventory control for TPM in the middle of the process (shown for TPM =  $F_2$ )

# QUIZ. Are these structures workable (consistent)? Yes or No?

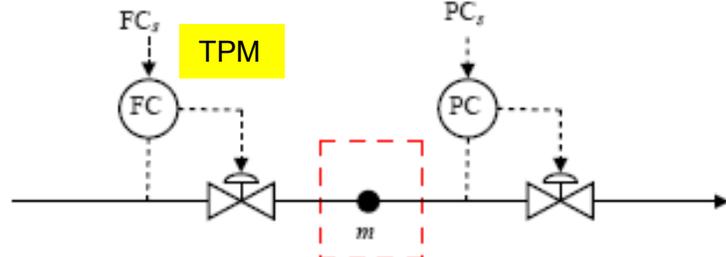
Hint: What happens to the mass holdup inside the red box? Is it self-regulated?



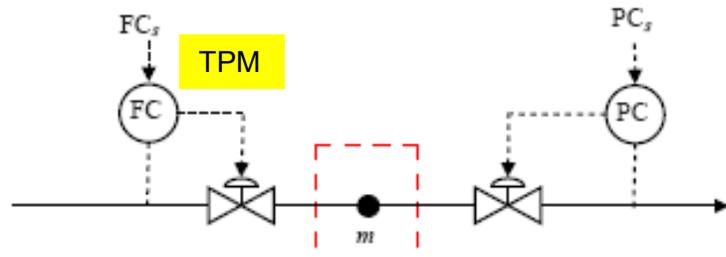
(a) [REDACTED]



(b) [REDACTED]



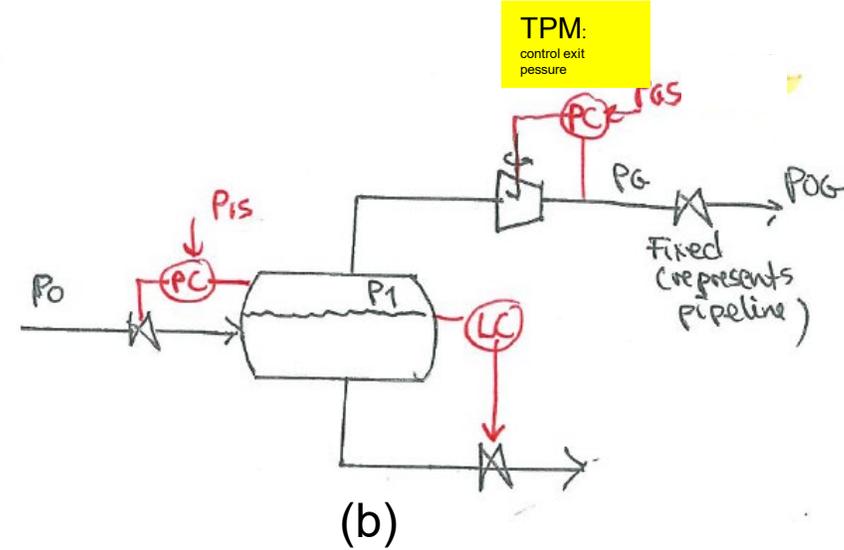
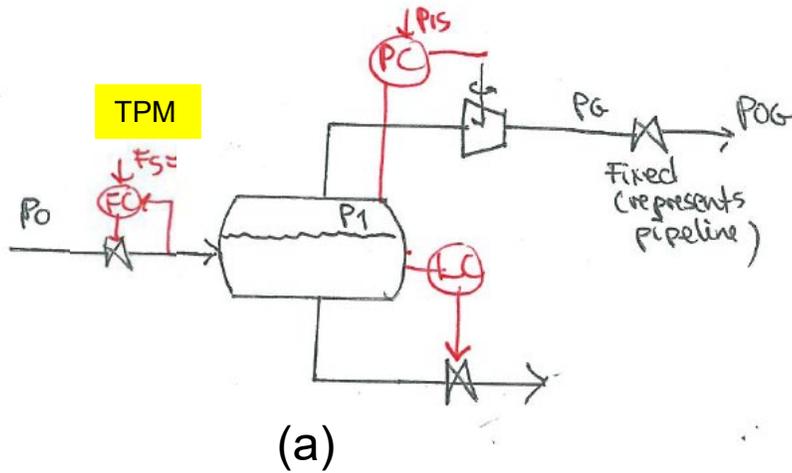
(c) [REDACTED]



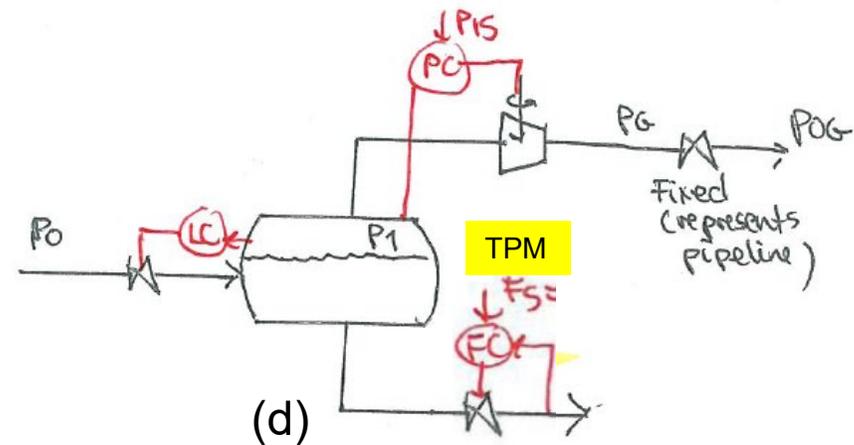
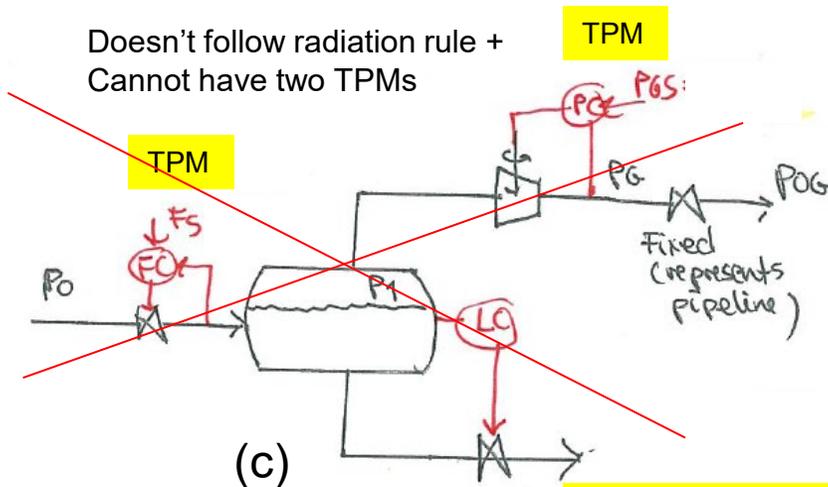
(d) [REDACTED] on inventory  $m$

# Quiz 2. Gas-liquid separator.

## Where is TPM? Consistent (One is not)?

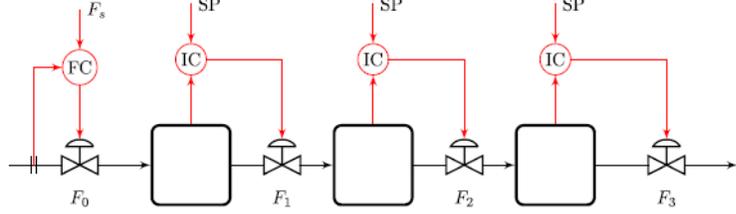


Doesn't follow radiation rule +  
Cannot have two TPMs

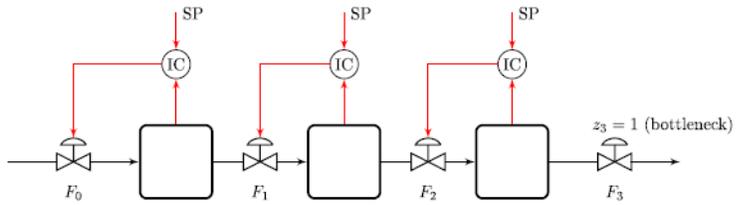


Case (a): Given feedrate. Could alternatively set  $p_0$   
Cases (b) and (c): Gas production limiting  
Case (d): Liquid production limiting

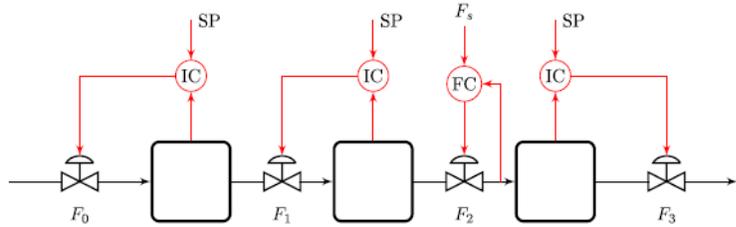
**Rule:** Setting in-pressure  $p_0$  sets inflow = TPM at inlet or inlet direction (no cases above)  
Setting out-pressure  $p_G$  sets outflow = TPM at outlet or outlet direction (offdiagonal two cases)



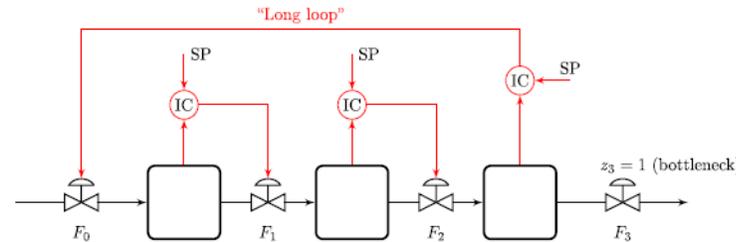
(a) Inventory control in direction of flow (for given feed flow,  $TPM = F_0$ )



(b) Inventory control in opposite direction of flow (for given product flow,  $TPM = F_3$ )

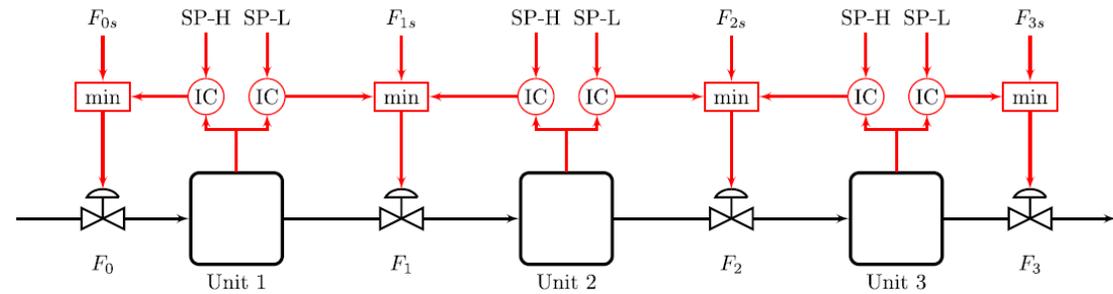


(c) Radiating inventory control for TPM in the middle of the process (shown for  $TPM = F_2$ )



(d) Inventory control with undesired "long loop", not in accordance with the "radiation rule" (for given product flow,  $TPM = F_3$ )

## Bidirectional inventory control



**Fig. 36.** Bidirectional inventory control scheme for automatic reconfiguration of loops (in accordance with the radiation rule) and maximizing throughput (Shinsky, 1981) (Zotică et al., 2022).

SP-H and SP-L are high and low inventory setpoints, with typical values 90% and 10%. Strictly speaking, since there are setpoints on the (maximum) flows ( $F_{i,s}$ ), the four valves should have slave flow controllers (not shown). However, one may instead have setpoints on valve positions (replace  $F_{i,s}$  by  $z_{i,s}$ ), and then flow controllers are not needed.

**Fig. 35.** Inventory control for units in series. Cases (a), (b) and (c) are in accordance with the "radiation rule".

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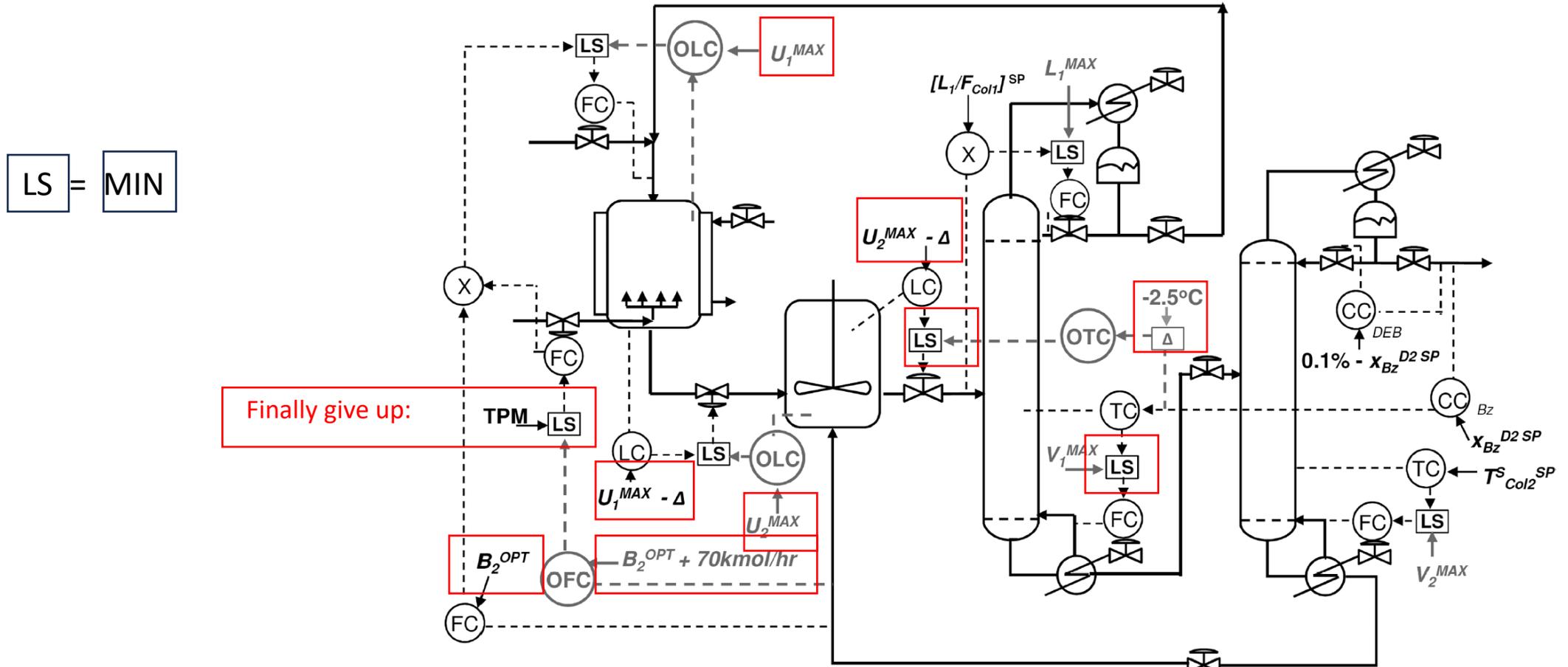


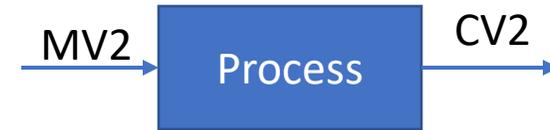
Figure 7. CS2 with overrides for handling equipment capacity constraints.

# Step 5: Design of regulatory control layer

Usually single-loop PID controllers

Choice of CVs (CV2):

- **CV2 = «drifting variables»**
  - Levels, pressures
  - Some temperatures
- CV2 may also include economic variables (CV1) that we want to control on a fast time scale
  - Hard constraints



Choice of MV2s and pairings (MV2-CV2):

**1. Main rule: “Pair close”. Want:**

- Large gain
- Small delay
- Small time constant

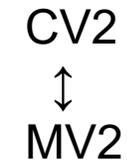
**2. Avoid MVs that may saturate in regulatory layer**

- Otherwise, will need logic for re-pairing (MV-CV switching)
- The exception is if you follow the **Input saturation rule: “Pair a MV that may saturate with a CV that can be given up (when the MV saturates)”**

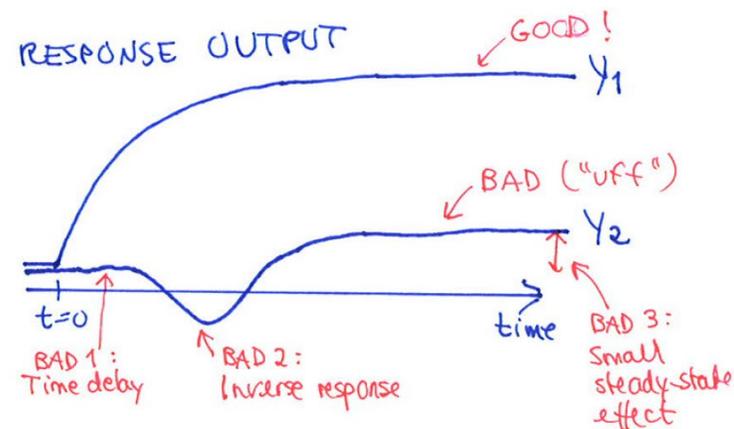
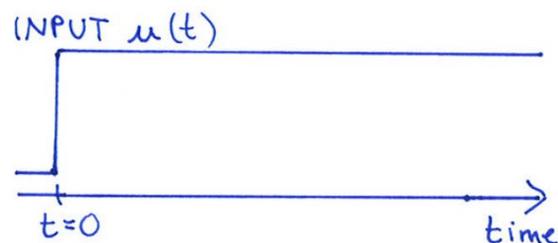
**3. Avoid pairing on negative steady-state RGA-elements**

- It’s possible, but then you must be sure that the loops are always working (no manual control or MV-saturation)

- May include cascade loops (flow control!) and some feedforward, decoupling, linearization



# Main rule: "Pair close"



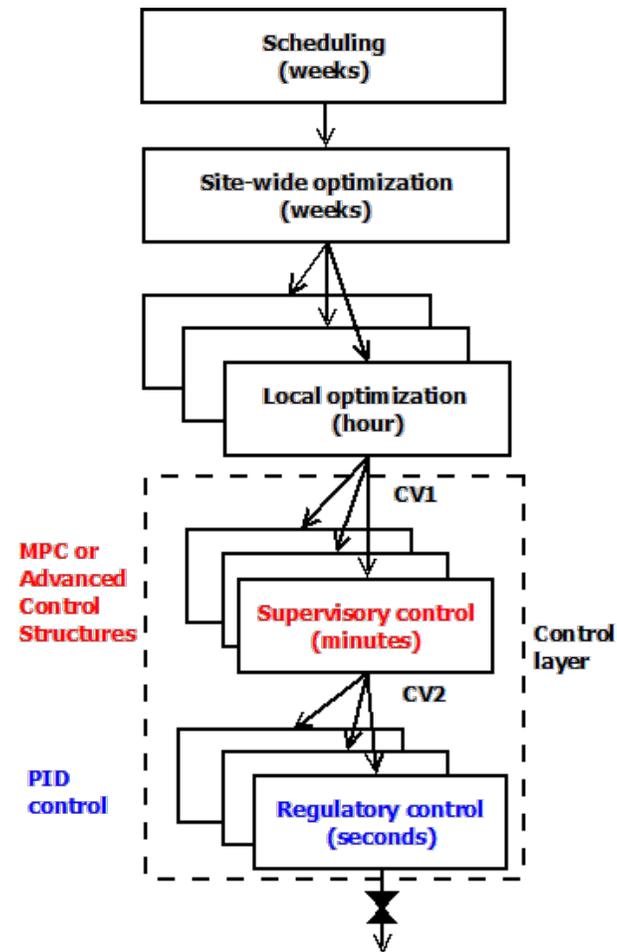
The response (from input to output) should be fast, large and in one direction.

Avoid time delay and inverse responses!

# Objectives of regulatory control layer

1. Allow for manual operation
2. Simple decentralized (local) PID controllers that can be tuned on-line
3. Take care of “fast” control
4. Track setpoint changes from the layer above
5. Local disturbance rejection
6. **Stabilization** (mathematical sense)
7. **Avoid “drift”** (due to disturbances) so system stays in “linear region”
  - **“stabilization”** (practical sense)
8. Allow for “slow” control in layer above (supervisory control)
9. Make control problem easy as seen from layer above
10. Use “easy” and “robust” measurements (pressure, temperature)
11. Simple structure
12. Contribute to overall economic objective (“indirect” control)
13. Should not need to be changed during operation

# Step 6: Design of Supervisory layer



# Objectives supervisory layer:

## 1. 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)
- Feedforward from disturbances (if helpful)
- Make use of extra inputs
- Make use of extra measurements

## 2. Keep an eye on stabilizing layer

- Avoid saturation in stabilizing layer

## 3. Switch control structures (CV1) depending on operating region

- Active constraints
- self-optimizing variables

## **Implementation supervisory control layer:**

- Alternative 1: Advanced regulatory control based on “simple elements” (decentralized control)
- Alternative 2: MPC

# Advanced regulatory control (ARC)

- Using simple standard elements

# QUIZ

What are the three most important inventions of process control?

- Hint 1: According to Sigurd Skogestad
- Hint 2: All were in use around 1940

## SOLUTION

1. PID controller, in particular, I-action
2. Cascade control
3. Ratio control

# The three main inventions of process control can only indirectly and with effort be implemented with MPC

1. Integral action with MPC: Need to add artificial integrating disturbance in estimator
  - ARC: Just add an integrator in the controller (use PID)
2. Cascade control with MPC: Need model for how  $u$  and  $d$  affect  $y_1$  and  $y_2$ .
  - ARC: Just need to know that control of  $y_2$  indirectly improves control of  $y_1$
3. Ratio control with MPC: Need model for how  $u$  and  $d$  affect property  $y$ 
  - ARC: Just need the insight that it is good for control of  $y$  to keep the ratio  $R=u/d$  constant

Because of this, MPC should be on top of a regulatory control layer with the setpoints for  $y_2$  and  $R$  as MVs.

# ARC: Standard Advanced control elements

Each element links a subset of inputs with a subset of outputs. Results in simple local design and tuning

First, there are some elements that are used to improve control for cases where simple feedback control is not sufficient:

- E1\*. Cascade control<sup>2</sup>
- E2\*. Ratio control
- E3\*. Valve (input)<sup>3</sup> position control (VPC) on extra MV to improve dynamic response.

Next, there are some control elements used for cases when we reach constraints:

- E4\*. Selective (limit, override) control (for output switching)
- E5\*. Split range control (for input switching)
- E6\*. Separate controllers (with different setpoints) as an alternative to split range control (E5)
- E7\*. VPC as an alternative to split range control (E5)

All the above seven elements have feedback control as a main feature and are usually based on PID controllers. Ratio control seems to be an exception, but the desired ratio setpoint is usually set by an outer feedback controller. There are also several features that may be added to the standard PID controller, including

- E8\*. Anti-windup scheme for the integral mode
- E9\*. Two-degrees of freedom features (e.g., no derivative action on setpoint, setpoint filter)
- E10. Gain scheduling (Controller tunings change as a given function of the scheduling variable, e.g., a disturbance, process input, process output, setpoint or control error)

In addition, the following more general model-based elements are in common use:

- E11\*. Feedforward control
- E12\*. Decoupling elements (usually designed using feedforward thinking)
- E13. Linearization elements
- E14\*. Calculation blocks (including nonlinear feedforward and decoupling)
- E15. Simple static estimators (also known as inferential elements or soft sensors)

Finally, there are a number of simpler standard elements that may be used independently or as part of other elements, such as

- E16. Simple nonlinear static elements (like multiplication, division, square root, dead zone, dead band, limiter (saturation element), on/off)
- E17\*. Simple linear dynamic elements (like lead-lag filter, time delay, etc.)
- E18. Standard logic elements

<sup>2</sup> The control elements with an asterisk \* are discussed in more detail in this paper.

# Standard Advanced control elements

First, there are some elements that are used to improve control for cases where simple feedback control is not sufficient:

- E1\***. Cascade control<sup>2</sup>
- E2\***. Ratio control
- E3\***. Valve (input)<sup>3</sup> position control (VPC) on extra MV to improve dynamic response.

Next, there are some control elements used for cases when we reach constraints:

- E4\***. Selective (limit, override) control (for output switching)
- E5\***. Split range control (for input switching)
- E6\***. Separate controllers (with different setpoints) as an alternative to split range control (E5)
- E7\***. VPC as an alternative to split range control (E5)

All the above seven elements have feedback control as a main feature and are usually based on PID controllers. Ratio control seems to be an exception, but the desired ratio setpoint is usually set by an outer feedback controller. There are also several features that may be added to the standard PID controller, including

- E8\***. Anti-windup scheme for the integral mode
- E9\***. Two-degrees of freedom features (e.g., no derivative action on setpoint, setpoint filter)
- E10**. Gain scheduling (Controller tunings change as a given function of the scheduling variable, e.g., a disturbance, process input, process output, setpoint or control error)

In addition, the following more general model-based elements are in common use:

- E11\***. Feedforward control
- E12\***. Decoupling elements (usually designed using feedforward thinking)
- E13**. Linearization elements
- E14\***. Calculation blocks (including nonlinear feedforward and decoupling)
- E15**. Simple static estimators (also known as inferential elements or soft sensors)

Finally, there are a number of simpler standard elements that may be used independently or as part of other elements, such as

- E16**. Simple nonlinear static elements (like multiplication, division, square root, dead zone, dead band, limiter (saturation element), on/off)
- E17\***. Simple linear dynamic elements (like lead-lag filter, time delay, etc.)
- E18**. Standard logic elements

**Gives a decomposed control system:**

- Each element links a subset of inputs with a subset of outputs
- Results in simple local tuning



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Review article

Advanced control using decomposition and simple elements

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- Layered decomposition
- Vertical decomposition
- Network architectures

ABSTRACT

The paper explores the standard advanced control elements commonly used in industry for designing advanced control systems. These elements include cascade, ratio, feedforward, decoupling, selectors, split range, and more, collectively referred to as “advanced regulatory control” (ARC). Numerous examples are provided, with a particular focus on process control. The paper emphasizes the shortcomings of model-based optimization methods, such as model predictive control (MPC), and challenges the view that MPC can solve all control problems, while ARC solutions are outdated, ad-hoc and difficult to understand. On the contrary, decomposing the control systems into simple ARC elements is very powerful and allows for designing control systems for complex processes with only limited information. With the knowledge of the control elements presented in the paper, readers should be able to understand most industrial ARC solutions and propose alternatives and improvements. Furthermore, the paper calls for the academic community to enhance the teaching of ARC methods and prioritize research efforts in developing theory and improving design methods.

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## 8.2. The harder problem: Control structure synthesis

As a third approach, machine learning may prove to be useful. Machine learning has one of its main strengths in pattern recognition, in a similar way to how the human brain works. I have observed over the years that some students, with only two weeks of example-based teaching, are able to suggest good process control solutions with feedback, cascade, and feedforward/ratio control for realistic problems, based on only a flowsheet and some fairly general statements about the control objectives. This is the basis for believing that machine learning (e.g., a tool similar to **ChatGPT**) may provide a good initial control structure, which may later be improved, either manually or by optimization. It is important that such a tool has a graphical interface, both for presenting the problem and for proposing and improving solutions.