Part 1.
Plantwide process control
«Control architectures»

Sigurd Skogestad

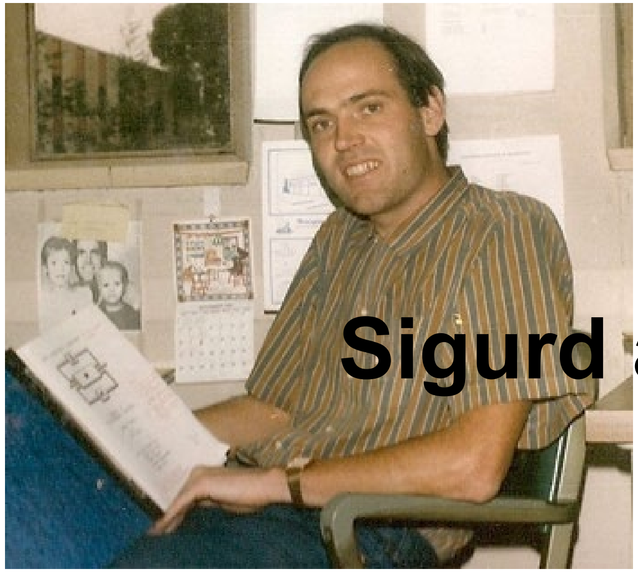
Plantwide control (Control archirecture)

- Objective: Put controllers on flow sheet (make P&ID)
- Two main objectives for control: Longer-term economics (CV1) and shorter-term stability (CV2)
- Regulatory (basic) control layer for CV2 and supervisory (advanced) control layer for CV1

How can we design a control system for a complete chemical plant?

Where do we start?

What should we control? And why?



Sigurd at Caltech (1984)

How we design a control system for a complete chemical plant?

- Where do we start?
- What should we control? and why?
- etc.
- etc.

Control system structure*

Alan Foss ("Critique of chemical process control theory", AIChE Journal,1973):

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?

*Current terminology: Control system architecture

Plantwide control =
Control structure (architecture) design

- *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 ("outputs")
  - Selection of manipulated variables ("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 (LQG, H-infinity, PID, decoupler, MPC etc.)

QUIZ
What are the three most important inventions of process control?

- Hint 1: According to Sigurd Skogestad
- Hint 2: All became commonly used in the 1940s

SOLUTION

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

Note: None of these are easily implemented using Model predictive control (MPC)

Main objectives of a control system

1. Economics: Implementation of acceptable (near-optimal) operation
2. Regulation: Stable operation

ARE THESE OBJECTIVES CONFLICTING?

- 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 (frequency range)

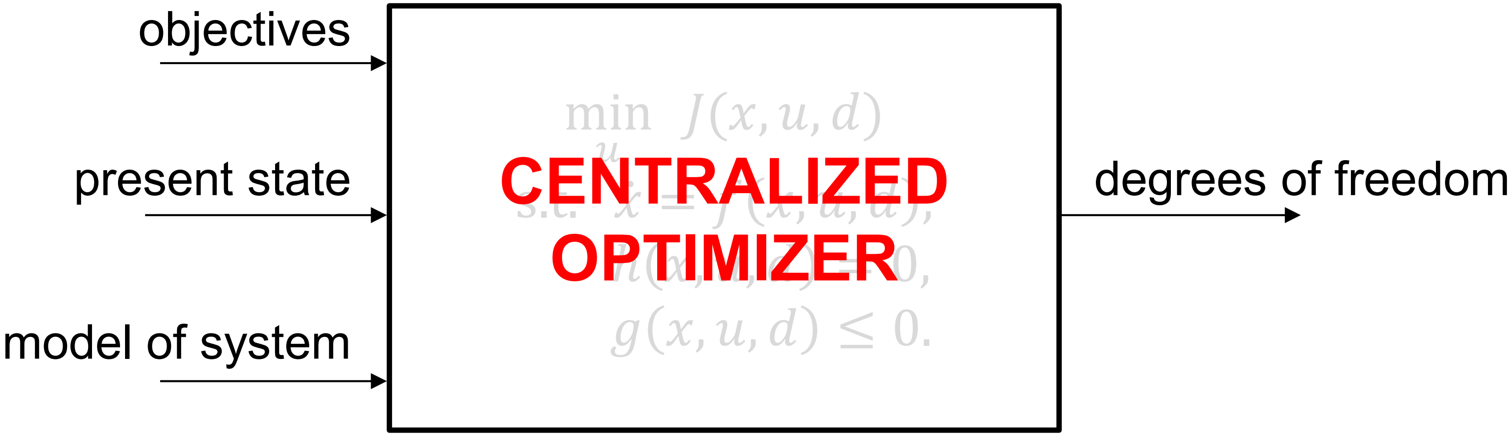
Optimal operation

General approach: minimize cost / maximize profit, subject to satisfying constraints (product quality, environment, resources)

Mathematically,

$$
\begin{aligned}
\min_{u} \quad & J(x,u,d) \\
\text{s.t.} \quad & \dot{x} = f(x,u,d), \\
& h(x,u,d) = 0, \\
& g(x,u,d) \leq 0.
\end{aligned}
$$

Optimal operation (in theory)



(example: Economic MPC)

Procedure:
- Obtain model of overall system
- Estimate present state
- Optimize all degrees of freedom

Problems:
- Model not available
- Optimization is complex
- Not robust (difficult to handle uncertainty)
- Slow response time

Engineering systems

- Most (all?) large-scale engineering systems are controlled using hierarchies of quite simple controllers
  - Large-scale chemical plant (refinery)
  - Commercial aircraft

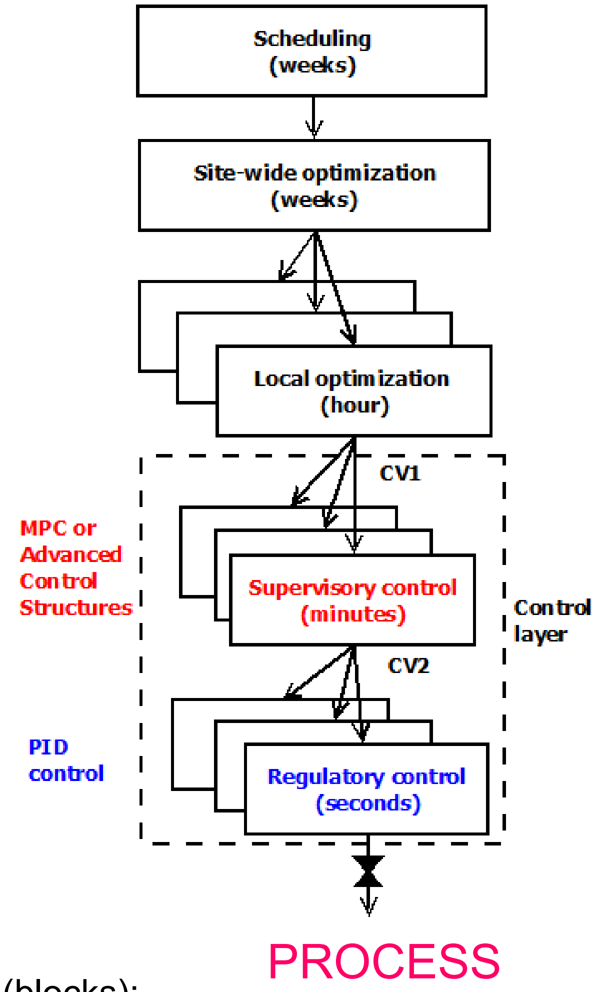
- 100’s of loops

- Simple components:

  on-off + PI-control + nonlinear fixes + some feedforward

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

In addition: Decomposition of controller into smaller elements (blocks): Feedforward element, nonlinear element, estimators (soft sensors), switching elements

Time scale separation: Control* layers

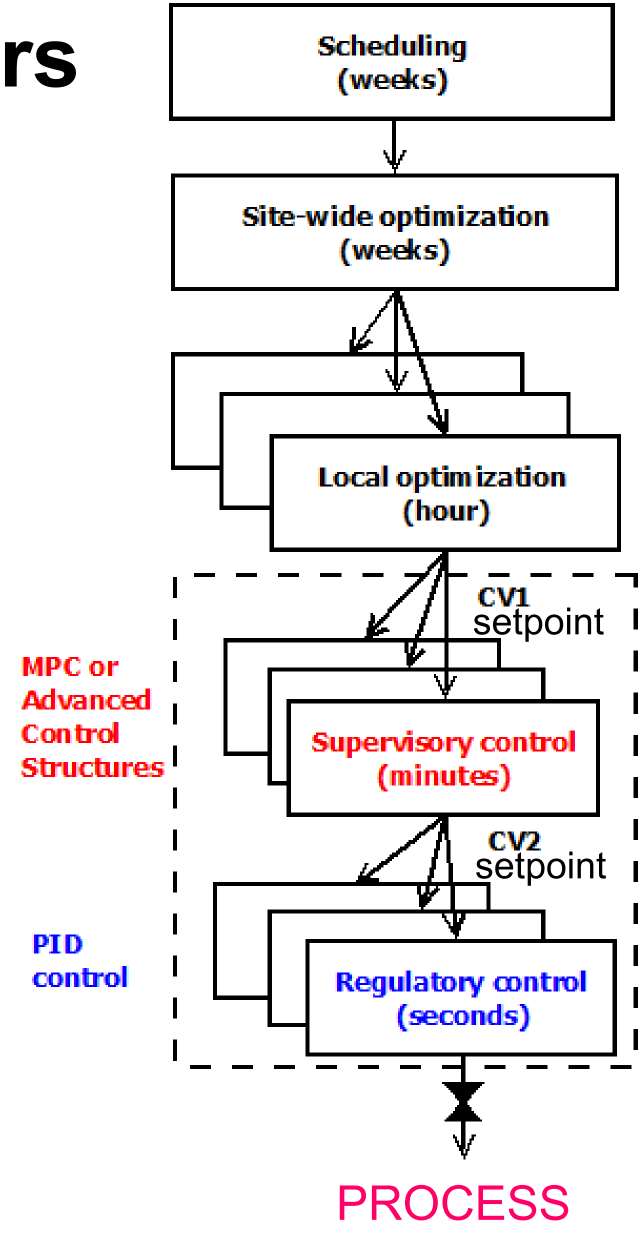
Two objectives for control: Stabilization and economics

- **Supervisory ("advanced") control layer**

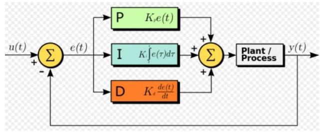
  Tasks:
  - Follow set points for CV1 from economic optimization layer
  - Switch between active constraints (change CV1)
  - Look after regulatory layer (avoid that MVs saturate, etc.)

- **Regulatory control (PID layer):**
  - Stable operation (CV2)

*My definition of «control» is that the objective is to track setpoints



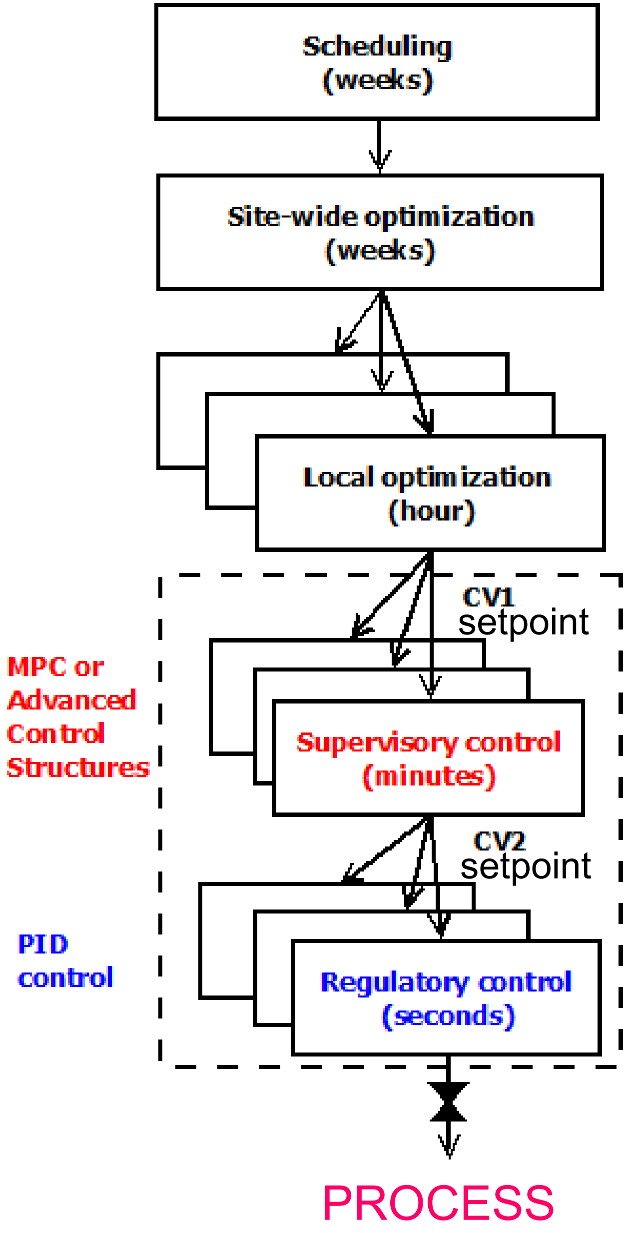
Control
layer



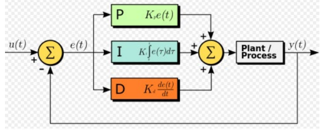
«Advanced» control

- Advanced: This is a relative term
- Usually used for anything than comes in addition to (or in top of) basic PID loops
- Mainly used in the «supervisory» control layer
- Two main options
  - **Standard «Advanced regulatory control» (ARC) elements**
    - Based on decomposing the control system
      - Cascade, feedforward, selectors, etc.
    - This option is preferred if it gives acceptable performance

  - **Model predictive control (MPC)**
    - Requires a lot more effort to implement and maintain
    - Use for interactive processes
    - Use with known information about future (use predictive capanulities)



Control layer



Combine control and optimization into one layer?

EMPC: Economic model predictive "control"

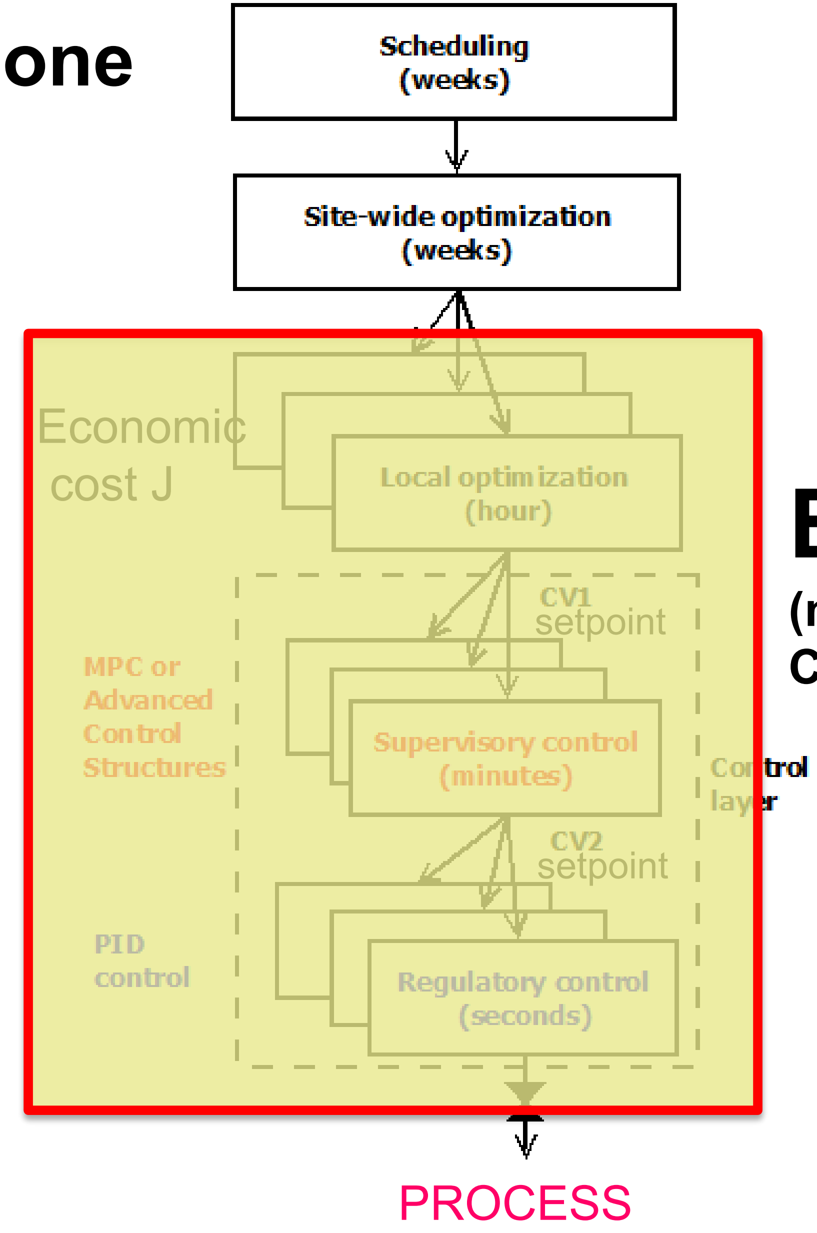
$J_{EMPC} = J + J_{control}$

- J [\$/s] $= J_{economic}$ = cost feed + cost energy – value products
- $J_{control} = \Sigma \Delta u_i^2$ (typical) - Penalize input usage

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

One layer (EMPC) is optimal theoreretically, 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 maintainance costly and time consuming



EMPC

(no setpoints CV1, CV2)

What about «conventional» RTO and MPC?

- Yes, it’s OK
- Both has been around for more than 50 years (since 1970s)
  - but the expected growth never came
- MPC is still used mostly in large-scale plants (petrochemical and refineries).
- MPC is far from replacing PID as some expected in the 1990s.

- But plants need to be run optimally:

  ⇒ Need something else than conventional RTO/MPC!

Alternative solutions for advanced control

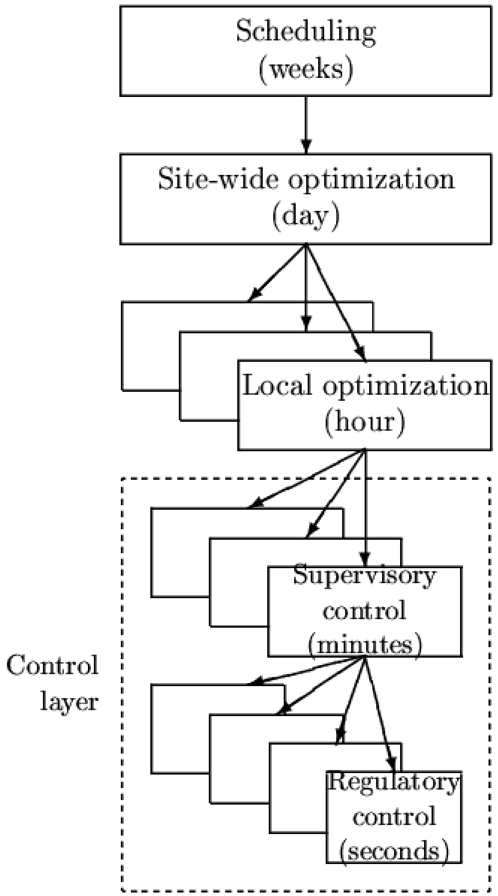
- Would like: Feedback solutions that can be implemented with minimum need for models

- **Machine learning?**
  - Requires a lot of data, not realistic for process control
  - And: Can only be implemented after the process has been in operation

- **"Classical advanced regulatory control" (ARC) based on single-loop PIDs?**
  - YES!
  - Extensively used by industry
  - Problem for engineers: Lack of design methods
    - Has been around since 1930's
    - But almost completely neglected by academic researchers
  - Main fundamental limitation: Based on single-loop (need to choose pairing)

ARC = Advanced regulatory control

Optimal operation and control objectives: What should we control?



CV_1 (economics)

CV_2 (stabilization)

Skogestad procedure for control structure design:

I. Top Down (analysis)
  - Step S1: Define operational objective (cost) and constraints
  - Step S2: Identify degrees of freedom and optimize operation for disturbances
  - Step S3: Implementation of optimal operation
    - What to control? (CV1) (self-optimizing control)
  - Step S4: Where set the production rate (TPM)? (Inventory control)

II. Bottom Up (design)
  - Step S5: Regulatory control: What more to control (CV2)?
  - Step S6: Supervisory control
  - Step S7: Real-time optimization

TPM = Throughput manipulator

Step S1. Define optimal operation (economics)

- Usually easy!
- What are the economic goals of the operation?
- Typical cost function*:

J = cost feed + cost energy – value products [$/s]

*No need to include fixed costs (capital costs, operators, maintainance) at ”our” time scale (hours)

Note: J=-P where P= Operational profit

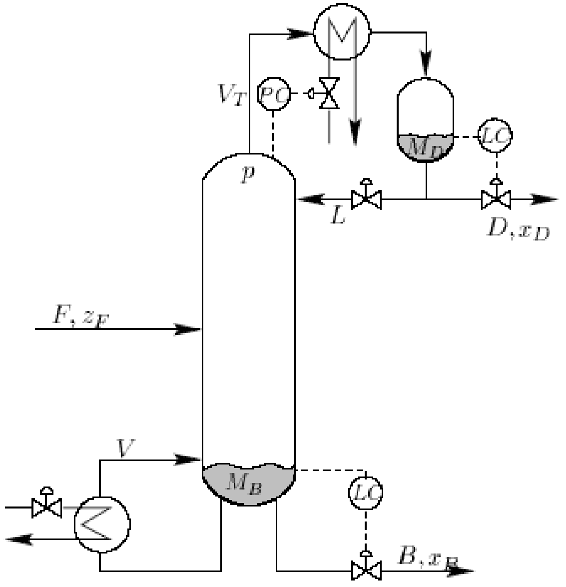
Example: distillation column

- Distillation at steady state with given p and F: N=2 DOFs, e.g. L and V (u)
- **Cost to be minimized (economics)**

cost energy (heating + cooling)

$$J = -P \text{ where } P = \underbrace{p_D D + p_B B}_{\text{value products}} - p_F F - p_V V$$

cost feed



- **Constraints**

  Purity D: For example, $x_{D,\text{ impurity}} \leq \max$

  Purity B: For example, $x_{B,\text{ impurity}} \leq \max$

  Flow constraints: $\min \leq D, B, L$ etc. $\leq \max$

  Column capacity (flooding): $V \leq V_{max}$, etc.

  Pressure: 1) p given (d) 2) p free (u): $p_{min} \leq p \leq p_{max}$

  Feed: 1) F given (d) 2) F free (u): $F \leq F_{max}$

- Optimal operation: Minimize J with respect to steady-state DOFs (u)

Skogestad procedure for control structure design:

I. Top Down
   - Step S1: Define operational objective (cost J) and constraints (easy!)
   - Step S2: (a) Identify degrees of freedom and (b) optimize operation for disturbances
     - Usually not easy! So often based on process insight
   - Step S3: Implementation of optimal operation
     - What to control? (primary CV's) (self-optimizing control)
   - Step S4: Where set the production rate? (Inventory control)

II. Bottom Up
   - Step S5: Regulatory control: What more to control (secondary CV's)?
   - Step S6: Supervisory control
   - Step S7: Real-time optimization

Step S2a: Degrees of freedom (DOFs) for operation

IMPORTANT!

DETERMINES THE NUMBER OF VARIABLES TO CONTROL!

- **No. of CV1 = No. of steady-state DOFs**

How many? NOT as simple as one may think!

To find all operational (dynamic) degrees of freedom:

- Count valves! (N_{valves})
- "Valves" also includes adjustable compressor power, etc.
  Anything we can manipulate!

BUT: not all these have a (steady-state) effect on the economics

How many Steady-state degrees of freedom (DOFs)?

Methods to obtain no. of steady-state degrees of freedom (N_{ss}):

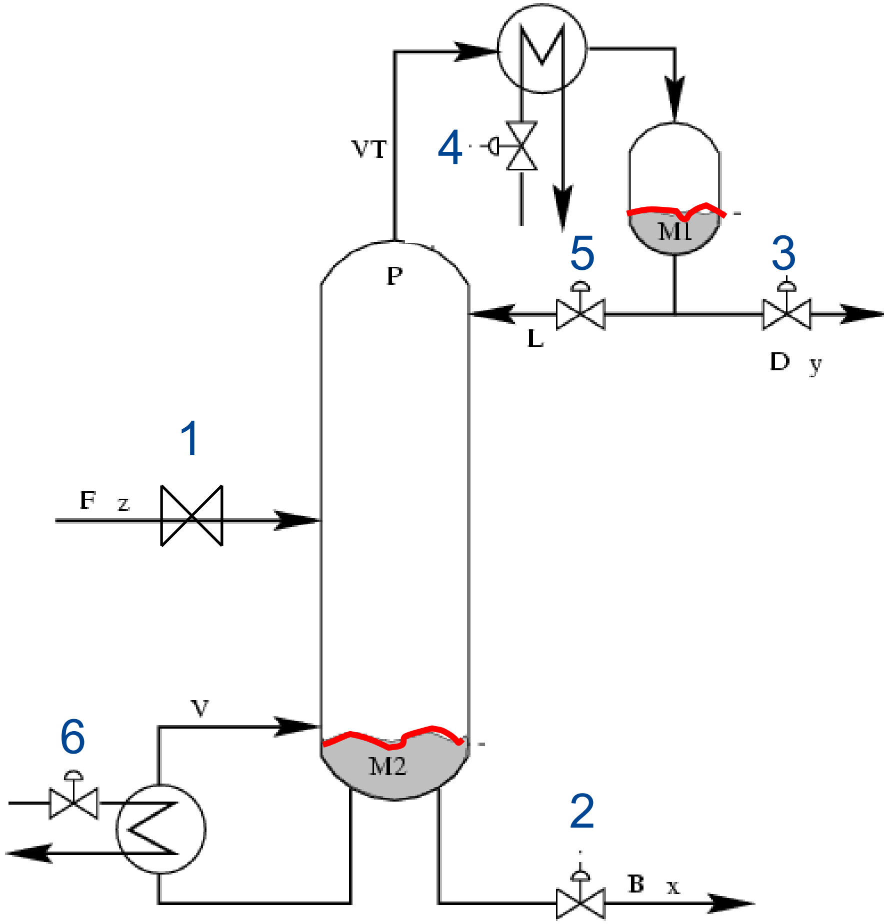
1. Equation-counting
   - N_{ss} = no. of variables – no. of equations/specifications
   - Very difficult in practice
2. **Valve-counting (easier!)**
   - $N_{ss} = N_{valves} - N_{0ss} - N_{specs}$
   - **N_{valves}: include also variable speed for compressor/pump/turbine**
   - **N_{specs}: Fixed variables (which are not later included in constraints)**
   - **N_{0ss} = variables with no steady-state effect**
     - **Inputs/MVs with no steady-state effect (e.g. extra bypass)**
     - **Outputs/CVs with no steady-state effect that need to be controlled (e.g., liquid levels)**
3. Potential number for some units (useful for checking!)
4. Correct answer: Will eventually find it when we perform optimization

CV = controlled variable

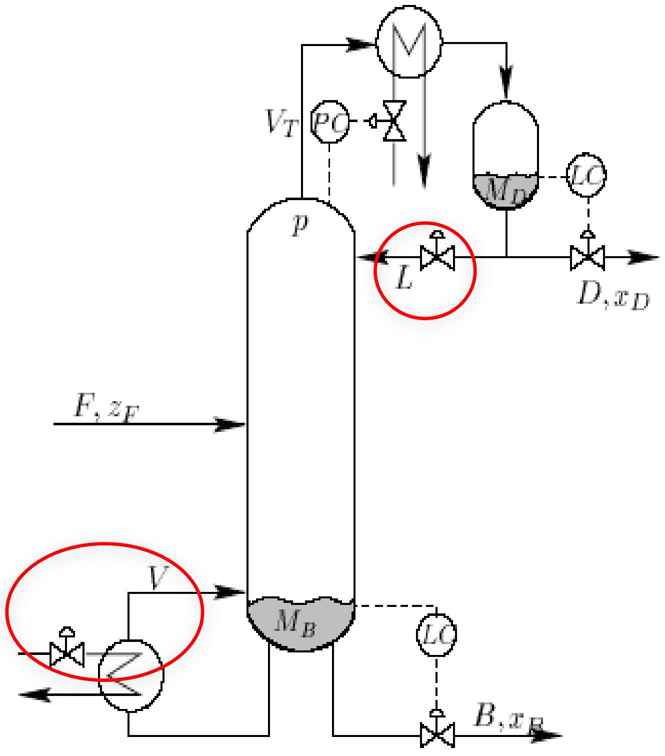
$N_{ss} = N_{valves} - N_{0ss} - N_{specs}$

Steady-state DOFs

Example: Distillation column



With levels and pressure controlled and given feed (LV-configuration):



$N_{valves} = 6$, $N_{0ss} = 2^*$,

$N_{DOF,SS} = 6 - 2 = 4$ (including feed and pressure as DOFs.

If feed and pressure are fixed: N_{specs}=2 and $N_{DOF,ss}$=4-2 = 2)

*N_{0ss} : no. controlled variables with no steady-state effect (here: levels M1 and M2)

NEED TO IDENTIFY 2 more CV's

- Typical: Top and btm composition

Step S2b: Optimize for expected disturbances

- What are the optimal values for our degrees of freedom u (MVs)?

J = cost feed + cost energy – value products

- Minimize J with respect to u for given disturbance d (usually steady-state):

$$\min_{u} J(x, u, d)$$

subject to:

- Model equations : $\dot{x} = f(x, u, d) = 0$
- Operational constraints: $g(x, u, d) \leq 0$

OFTEN VERY TIME CONSUMING

- Commercial simulators (Aspen, Unisim/Hysys) are set up in “design mode” and often work poorly in “operation (rating) mode”.
- Optimization methods in commercial simulators often poor
  - We can use Matlab or even Excel “on top”

Step S2b: Optimize for expected disturbances

- Need good model, usually steady-state
- Optimization is time consuming! But it is offline
- **Main goal: Identify ACTIVE CONSTRAINTS (optimal to maintain)**
- A good engineer can often guess the active constraints:

Example:

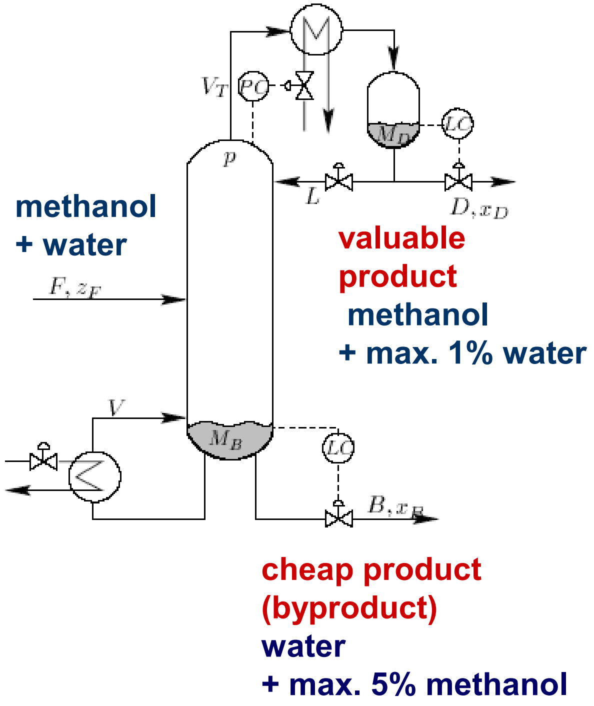
Cost J = T [h]
Constraint: v ≤ 50 km/h
Control implementation: Cruise control with setpoint 50 km/h (active constraint)

Example Step S2b: Active constraints for distillation

- **Both products (D, B) generally have purity specs**

- **Rule 1: Purity spec. always active for valuable product**
  - Reason: 1. Maximize amount of valuable product (D or B)
    - Avoid product "give-away" (So "sell water as methanol")
  - Reason 2: Save energy (because overpurification costs energy)

- **Rule 2: May overpurify (not control) cheap product**
  - Reason: Increase amount of valuable product ("reduce loss of methanol in bottom product")
  - This typically results in an unconstrained optimum because overpurification costs energy ("optimal purity of cheap product")



Step S2b: Optimize for expected disturbances

min J = cost feed + cost energy – value products

Generally: Two main cases (modes) depending on market conditions:

Mode 1 (low product price). Given throughput (feed rate)

Mode 2 (high product price). Maximum production (more constrained)

Comment: Depending on prices, Mode 1 may include many subcases (active constraints regions)

Mode 1. Given feedrate

Amount of products is then usually indirectly given and

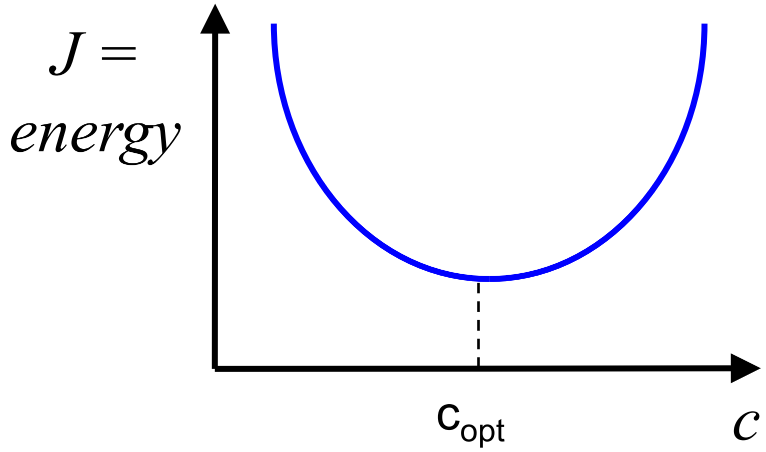
$$J = \underbrace{\text{cost feed} - \text{value products}}_{\text{Often constant}} + \textit{cost energy}$$

Optimal operation is then usually *unconstrained*

"maximize efficiency (energy)"

Control:

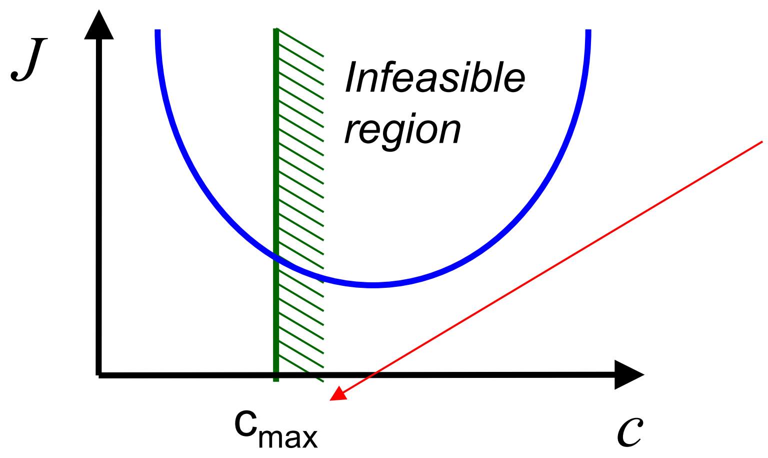
- Operate at optimal trade-off
- NOT obvious what to control
- CV = Self-optimizing variable



Mode 2. Maximum production

J = cost feed + cost energy – value products

- Assume feed rate is degree of freedom
- Assume products much more valuable than feed
- Optimal operation is then to maximize product rate
- **“max. constrained”, prices do not matter**



Control:
- Focus on tight control of bottleneck
- “Obvious what to control”
- CV = ACTIVE CONSTRAINT
- $CV_s = c_{max}$

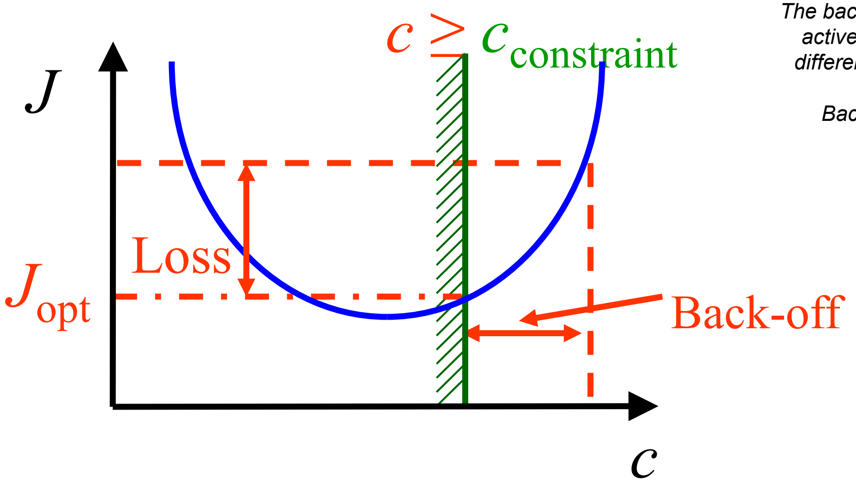
Step S3. Implementation of optimal operation

- Assume we have analyzed the optimal way of operation. How should it be implemented?

- **What to control?** (primary CV's)
  1. Active constraints
  2. Self-optimizing variables (for unconstrained degrees of freedom)

1. Control of Active output constraints

Need back-off



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 |

a) If constraint can be violated dynamically (only average matters)
- Required Back-off = "measurement bias" (steady-state measurement error for *c*)

b) If constraint <u>cannot</u> be violated dynamically ("hard constraint")
- Required Back-off = "measurement bias" + maximum dynamic control error

Want tight control of hard output constraints to reduce the back-off. "Squeeze and shift"-rule

Motivation for better control of active constraints: Squeeze and shift rule

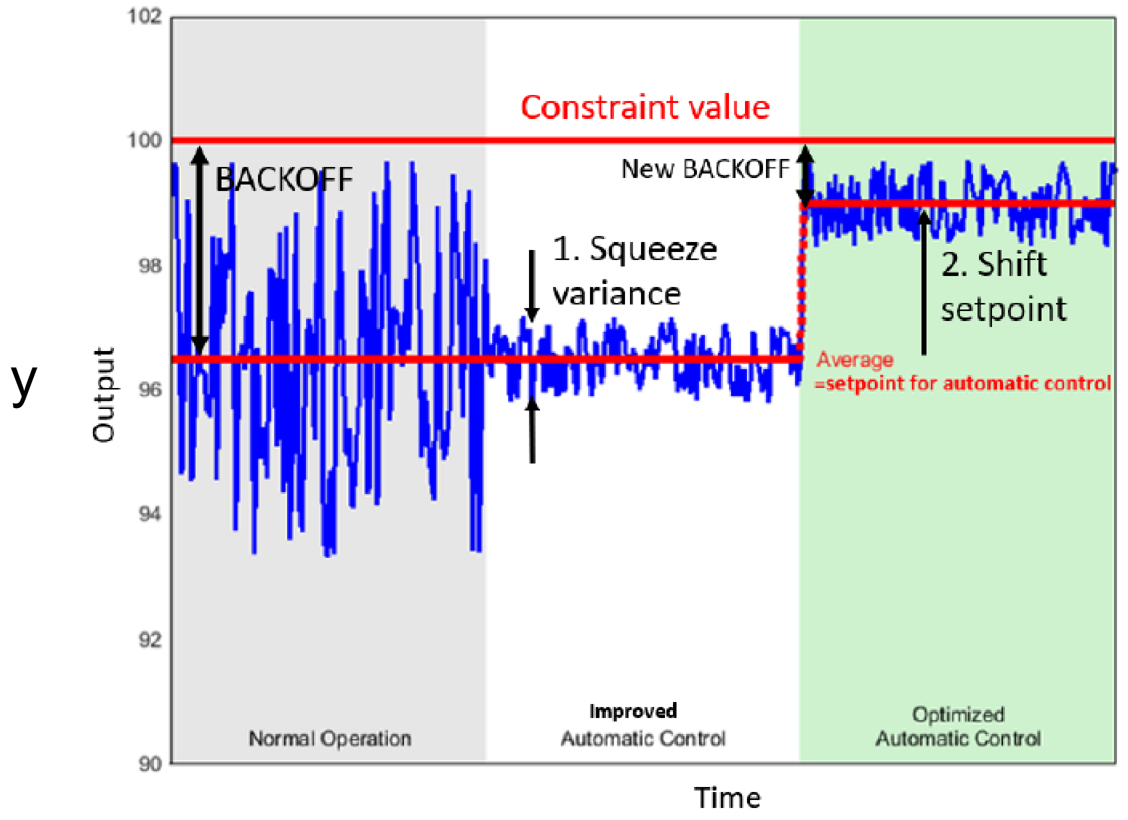
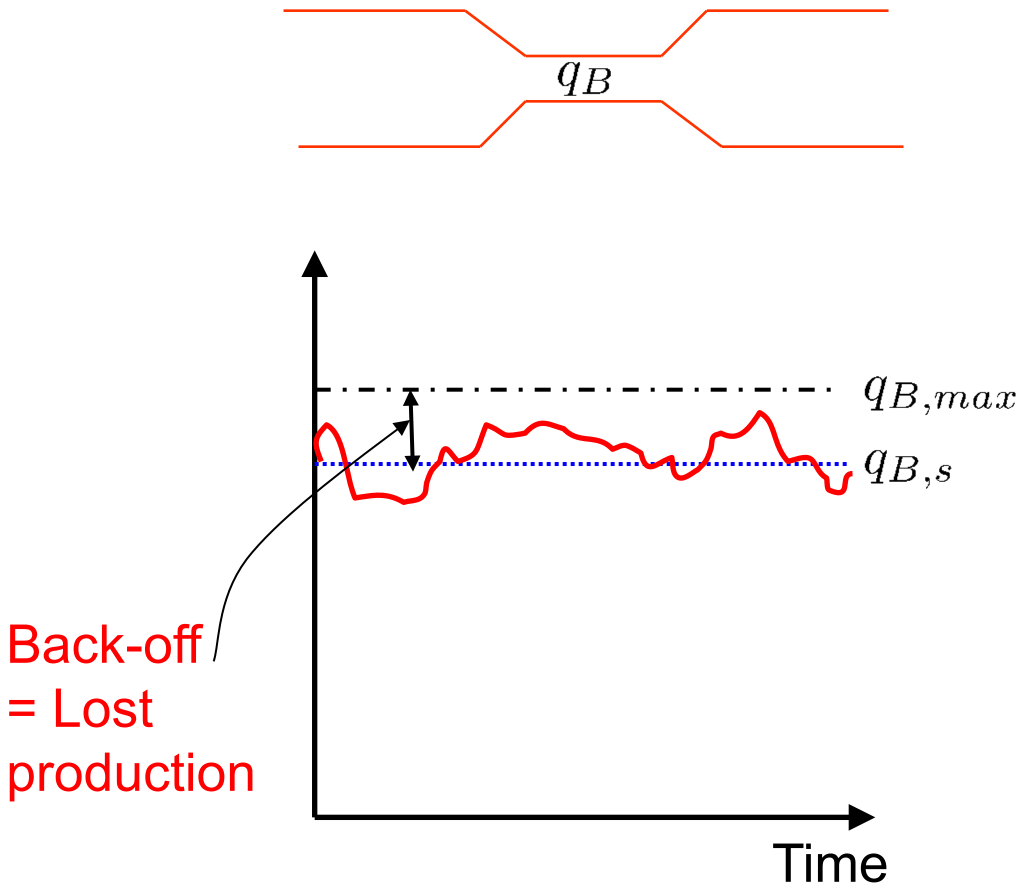


Figure 8: Squeeze and shift rule: Squeeze the variance by improving control and shift the setpoint closer to the constraint (i.e., reduce the backoff) to optimize the economics (Richalet et al., 1978).

Example: max. throughput.

Want tight bottleneck control to reduce backoff!

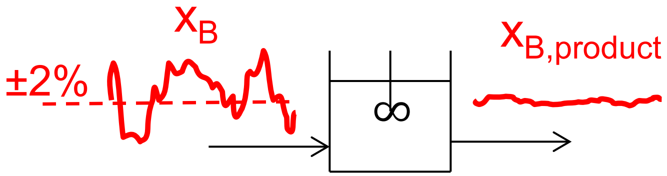


Example active constraint: purity on distillate

x_B = purity of product > 95% (min.)

- D_2 directly to customer (hard constraint)
  - Measurement error (bias): 1%
  - Control error (variation due to poor control): 2%
  - Backoff = 1% + 2% = 3%
  - Setpoint x_{Bs}= 95 + 3% = 98% (to be safe)
  - Can reduce backoff with better control ("squeeze and shift")

- D_2 to large mixing tank (soft constraint)
  - Measurement error (bias): 1%
  - Backoff = 1%
  - Setpoint x_{Bs}= 95 + 1% = 96% (to be safe)
  - Do not need to include control error because it averages out in tank



2. Unconstrained optimum

Control “self-optimizing” variable! (More on this soon!)

- Which variable is best?
- Often not obvious

What are good self-optimizing variables?

1. Optimal value of CV is constant
2. CV is “sensitive” to MV (large gain)

Note: Tight control of the self-optimizing variable is usually not important because optimum should be flat.

Conclusion optimal operation

ALWAYS:

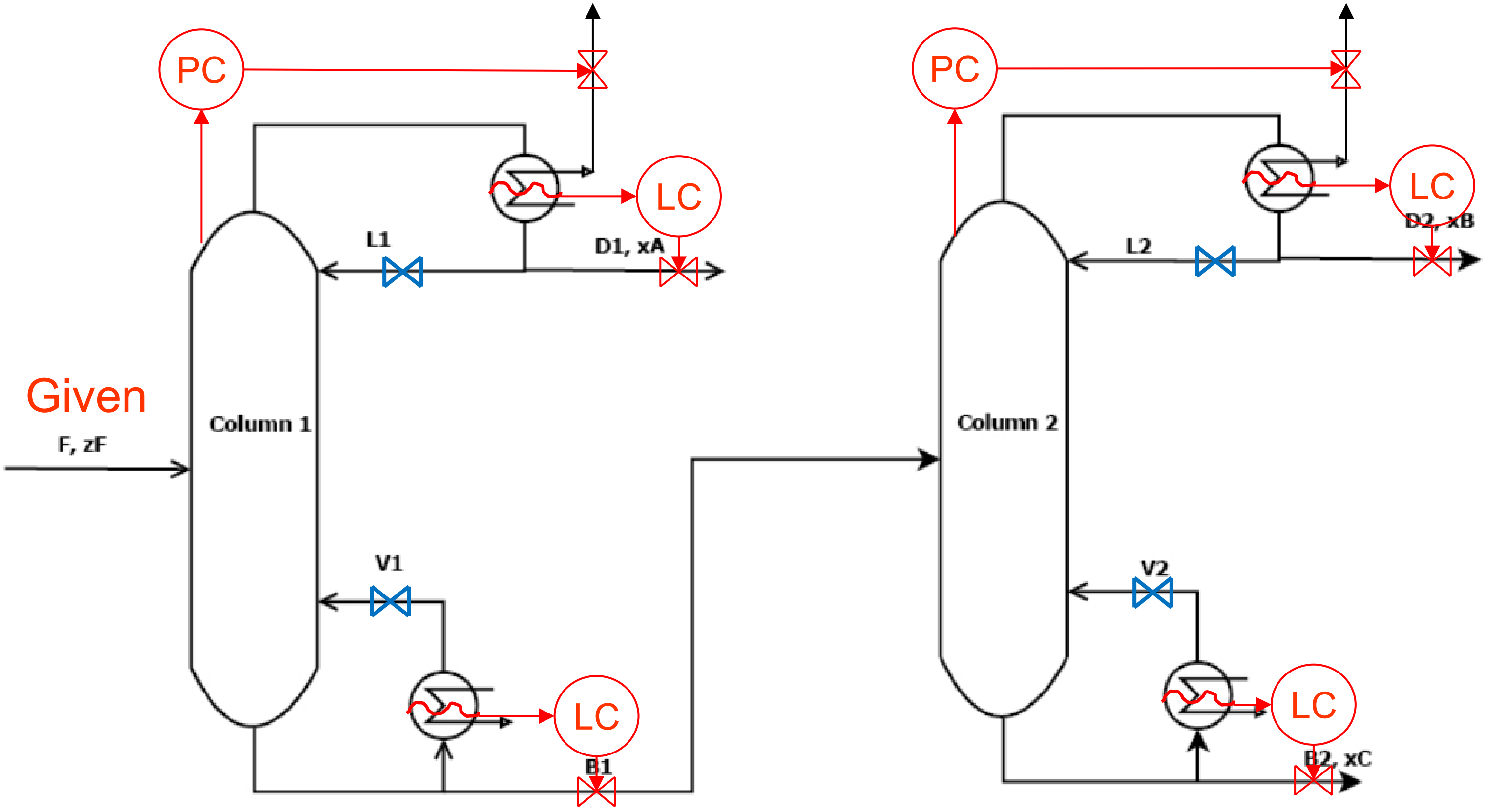
1. Control active constraints and control them tightly!!
   - Good times: Maximize throughput → tight control of bottleneck

2. Identify "self-optimizing" CVs for remaining unconstrained degrees of freedom

- Use offline analysis to find expected operating regions and prepare control system for this!
  - One control policy when prices are low (nominal, unconstrained optimum)
  - Another when prices are high (constrained optimum = bottleneck)

ONLY if necessary: consider RTO <u>on top of this</u>

Example Steps 1, 2 & 3: Distillation columns in series

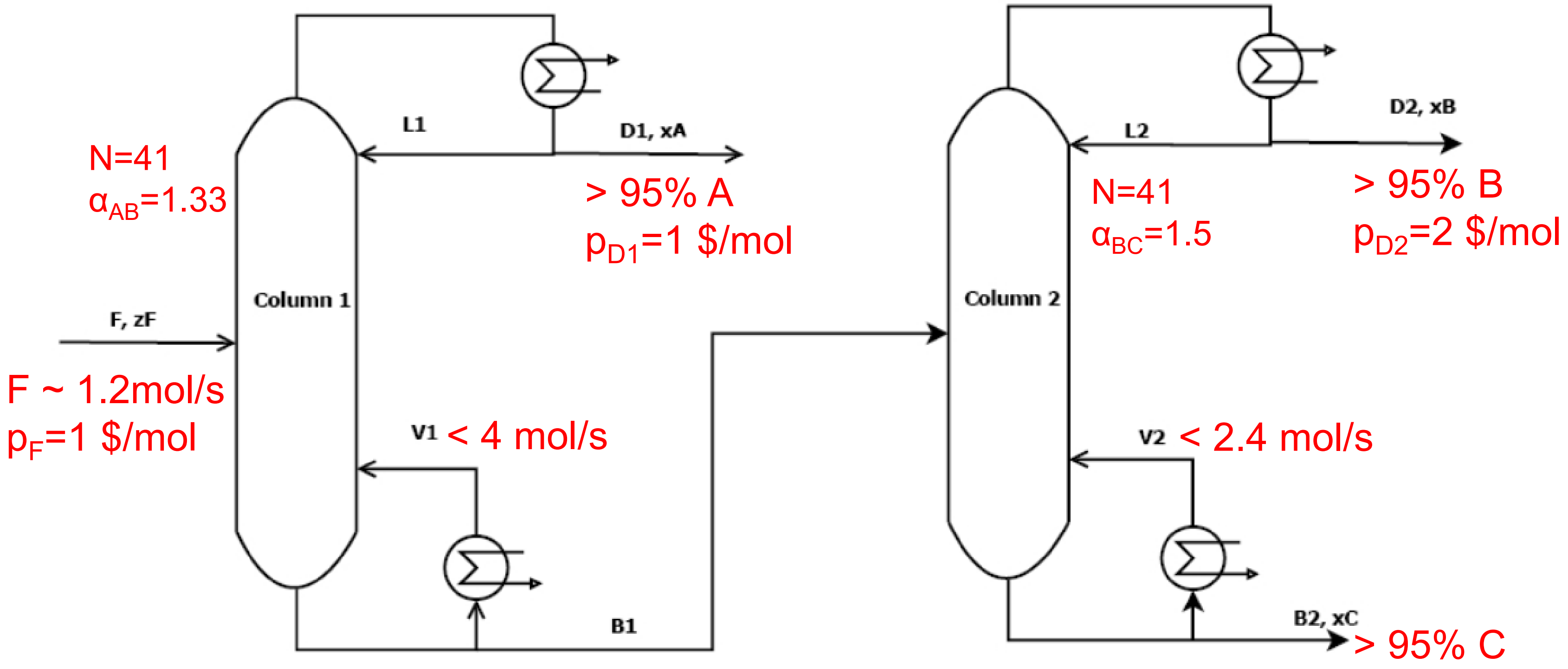


Given feed and pressures: We have 4 remaining steady-state MVs (L1, V1, L2, V2)
What more should we control?
HINT: CONTROL ACTIVE CONSTRAINTS

Red: Basic regulatory loops

Step S1: Cost and constraints

- 4 steady-state DOFs (e.g., L and V in each column)
- 5 (important) constraints: 3 product composition + 2 max. heat input



Cost (J) = - Profit = $p_F F + p_V(V_1+V_2) - p_{D1}D_1 - p_{D2}D_2 - p_{B2}B_2$

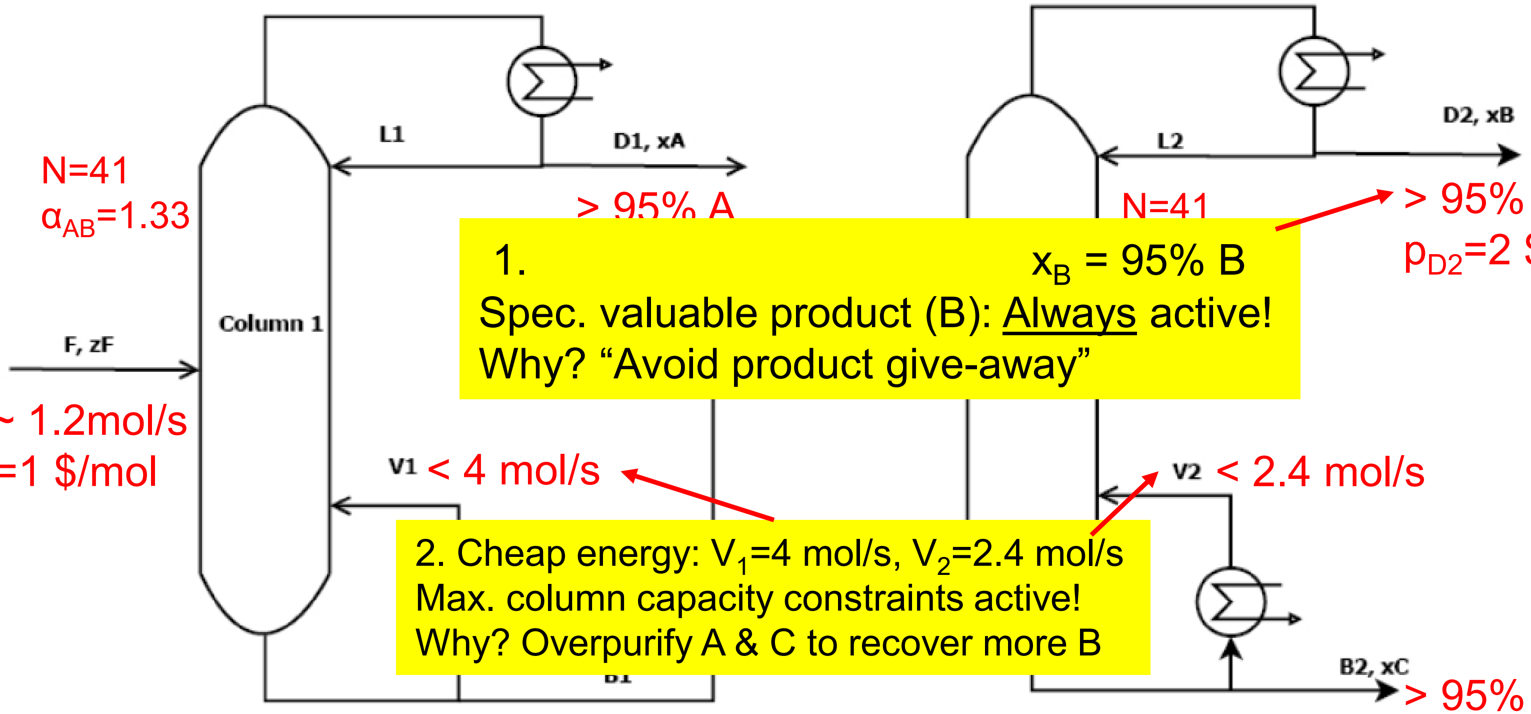
Energy price: p_V=0-0.2 $/mol (varies)

DOF = Degree Of Freedom

Ref.: M.G. Jacobsen and S. Skogestad (2011)

Step S2. Optimal operation

With given feed and pressures (disturbances): 4 steady-state DOFs
(e.g., L and V in each column)



N=41
α_{AB}=1.33

F ~ 1.2mol/s
p_F=1 \$/mol

> 95% A

N=41

> 95% B
p_{D2}=2 \$/mol

> 95% C
p_{B2}=1 \$/mol

V1 < 4 mol/s

V2 < 2.4 mol/s

1. x_B = 95% B
Spec. valuable product (B): Always active!
Why? “Avoid product give-away”

2. Cheap energy: V_1=4 mol/s, V_2=2.4 mol/s
Max. column capacity constraints active!
Why? Overpurify A & C to recover more B

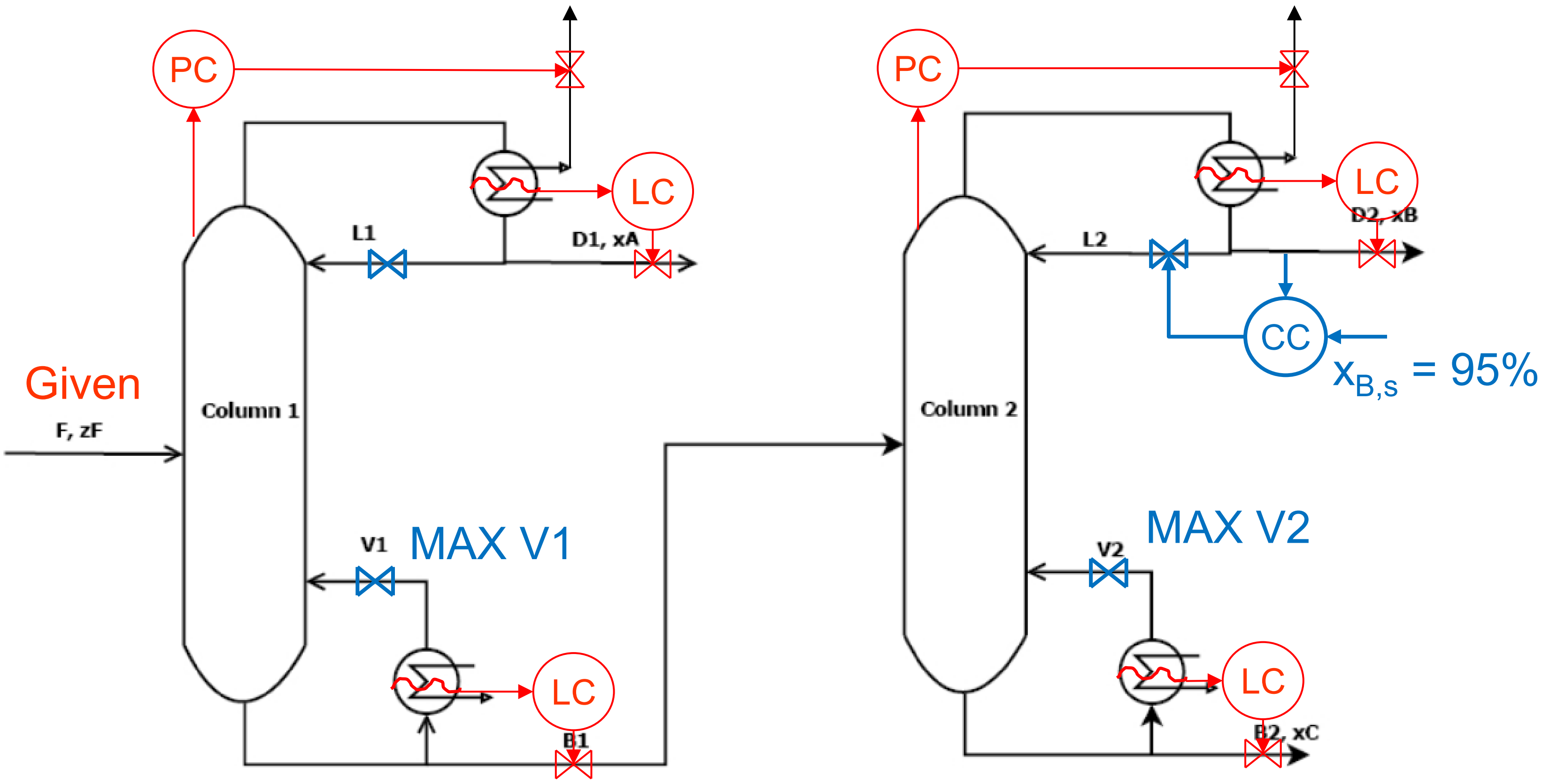
$$\text{Cost (J)} = -\text{Profit} = p_F F + p_V(V_1+V_2) - p_{D1}D_1 - p_{D2}D_2 - p_{B2}B_2$$

Energy price: p_V=0-0.2 \$/mol (varies)

DOF = Degree Of Freedom
Ref.: M.G. Jacobsen and S. Skogestad (2011)

QUIZ: What are the expected active constraints?
1. Always. 2. For low energy prices.

Step S3: Control 3 Active constraints:

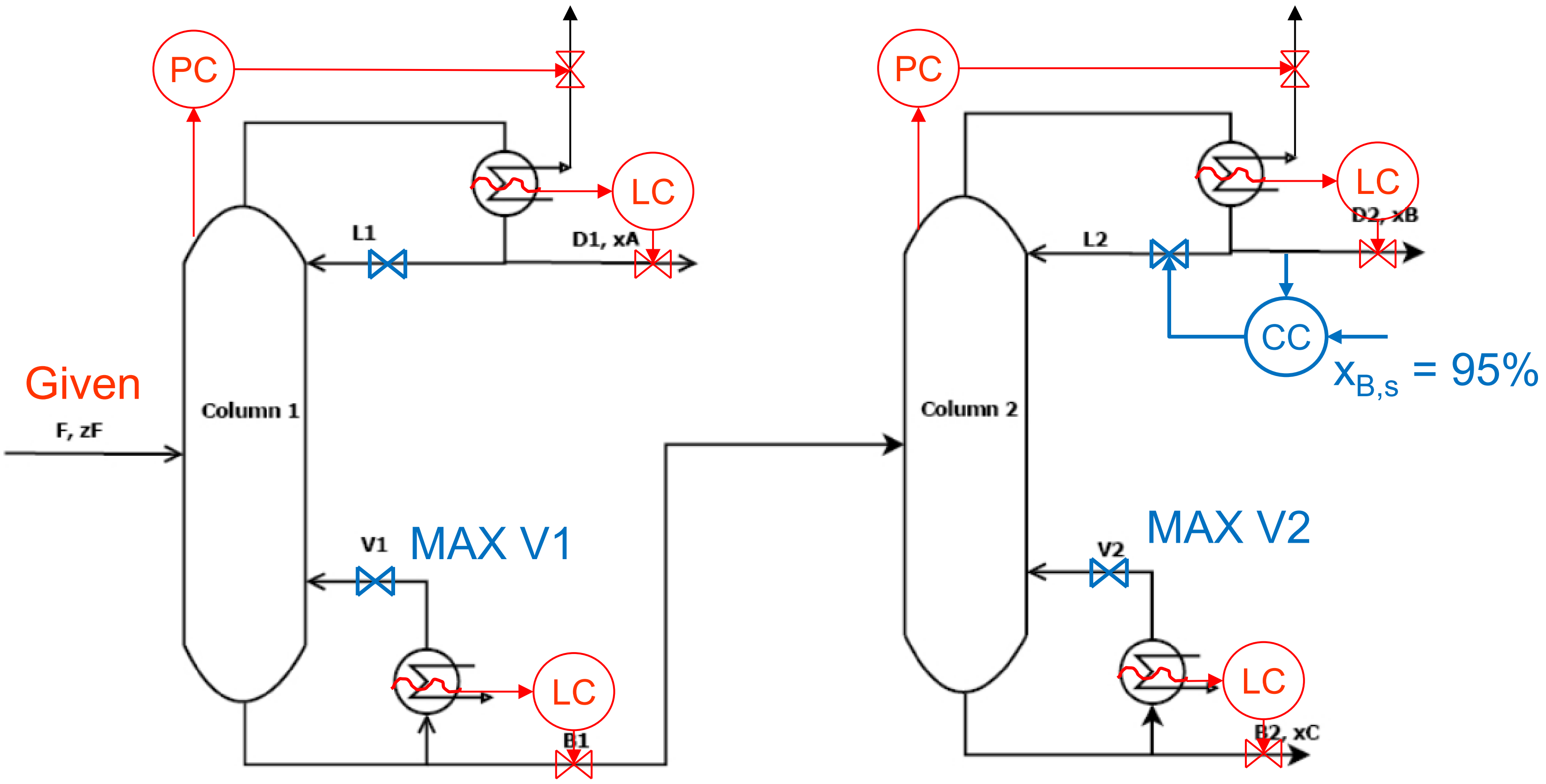


L1 not used. What more should we control?

Optimal to "overpurify" D1 - but optimal overpurification is **uncontrained** and varies with feedrate.
LOOK FOR "SELF-OPTIMIZING" CVs = Variables we can keep constant

Red: Basic regulatory loops

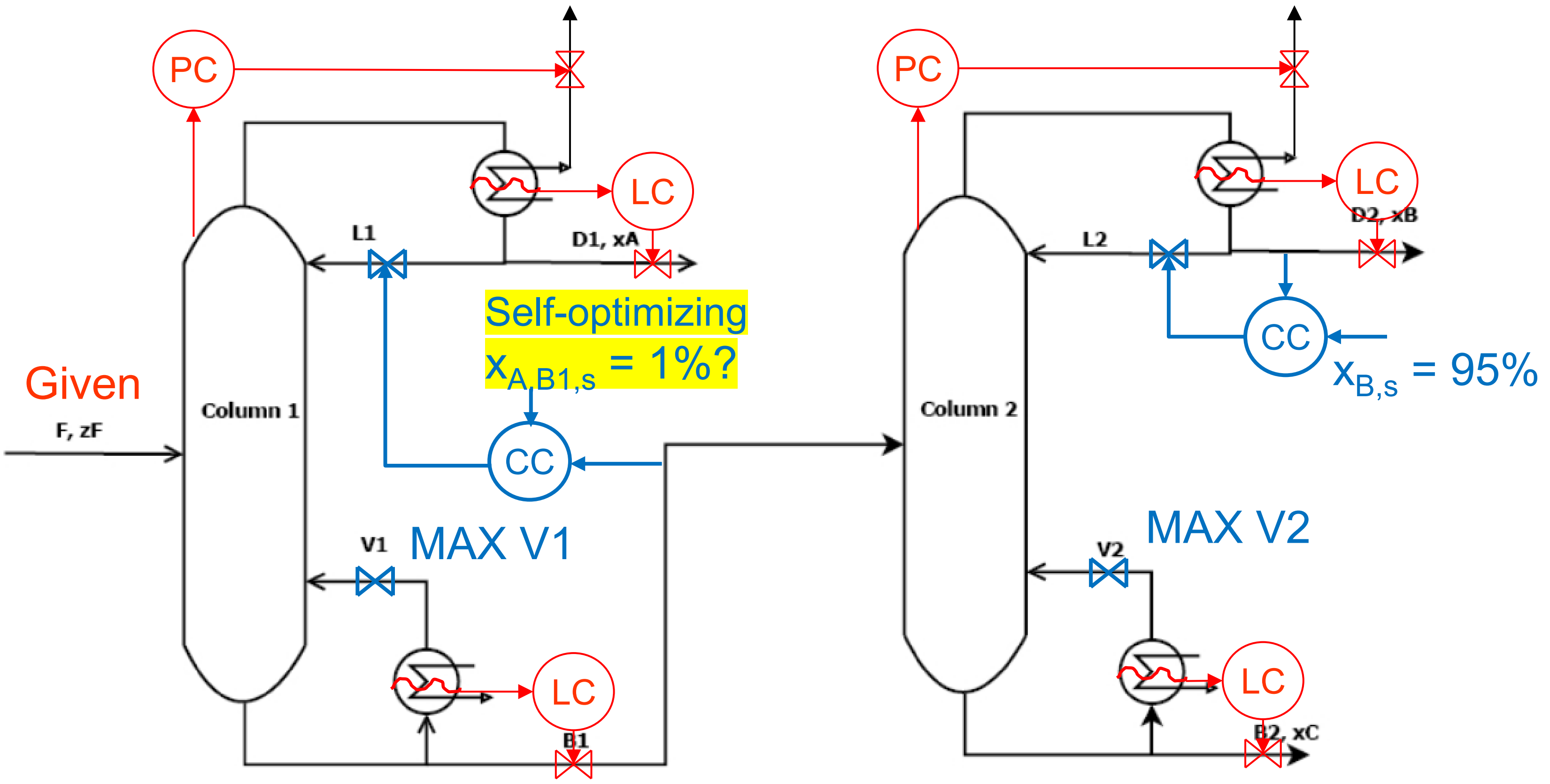
Step S3: Control 3 Active constraints:



What CV should L1 be paired with?
- Not: CV= x_A in D1! (why? x_A should vary with F!)
- Maybe: constant L1? (CV=L1)
- Better: CV= x_A in B1? Self-optimizing?

Red: Basic regulatory loops

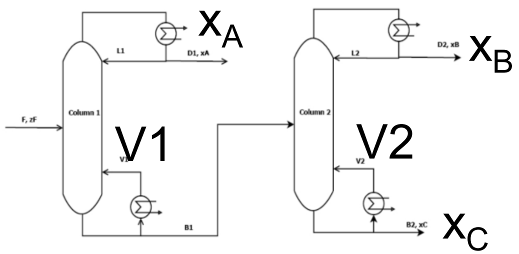
Step S3: Control 3 Active constraints + 1 self-optimizing

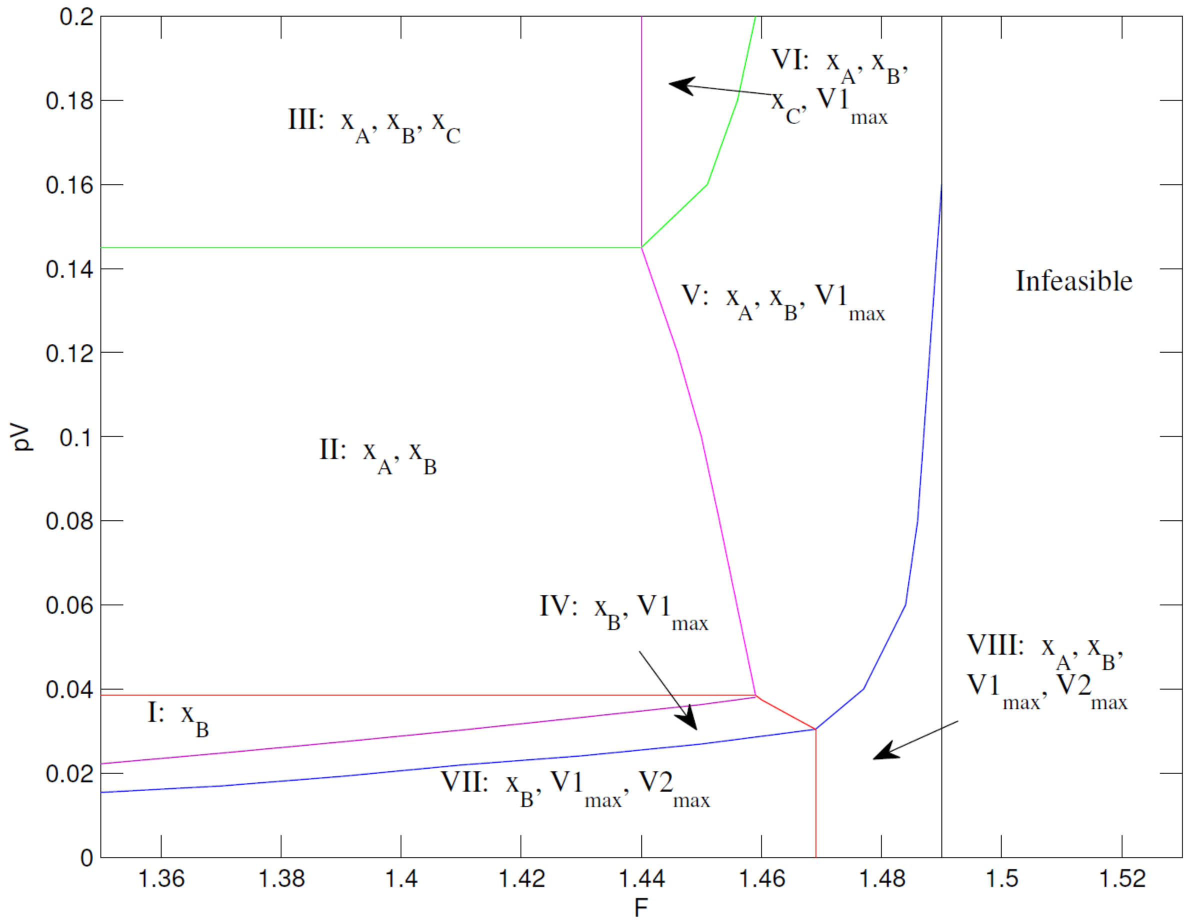


Red: Basic regulatory loops

Distillation example: Not so simple

Vary feedrate (F) and energy price (pV): 8 active constraint regions





- The figure shows the active constraints (between 1 and 4) in each region. x_B in D2 is always active.
- On the previous slide we only considered region VII («cheap energy» with pV small).
- In the «infeasible» region there are 5 constraints (xA, xB, xC, V1max, V2max) but only 4 DOFs. Must reduce F

How many active constraints regions?

- Maximum: 2^{n_c}

  where n_c = number of constraints

Distillation
$n_c = 5$
$2^5 = 32$

BUT there are usually fewer in practice

- Certain constraints are always active (reduces effective n_c)
- Only n_u can be active at a given time

  n_u = number of MVs (inputs)
- Certain constraints combinations are not possibe
  - For example, max and min on the same variable (e.g. flow)
- Certain regions are not reached by the assumed disturbance set

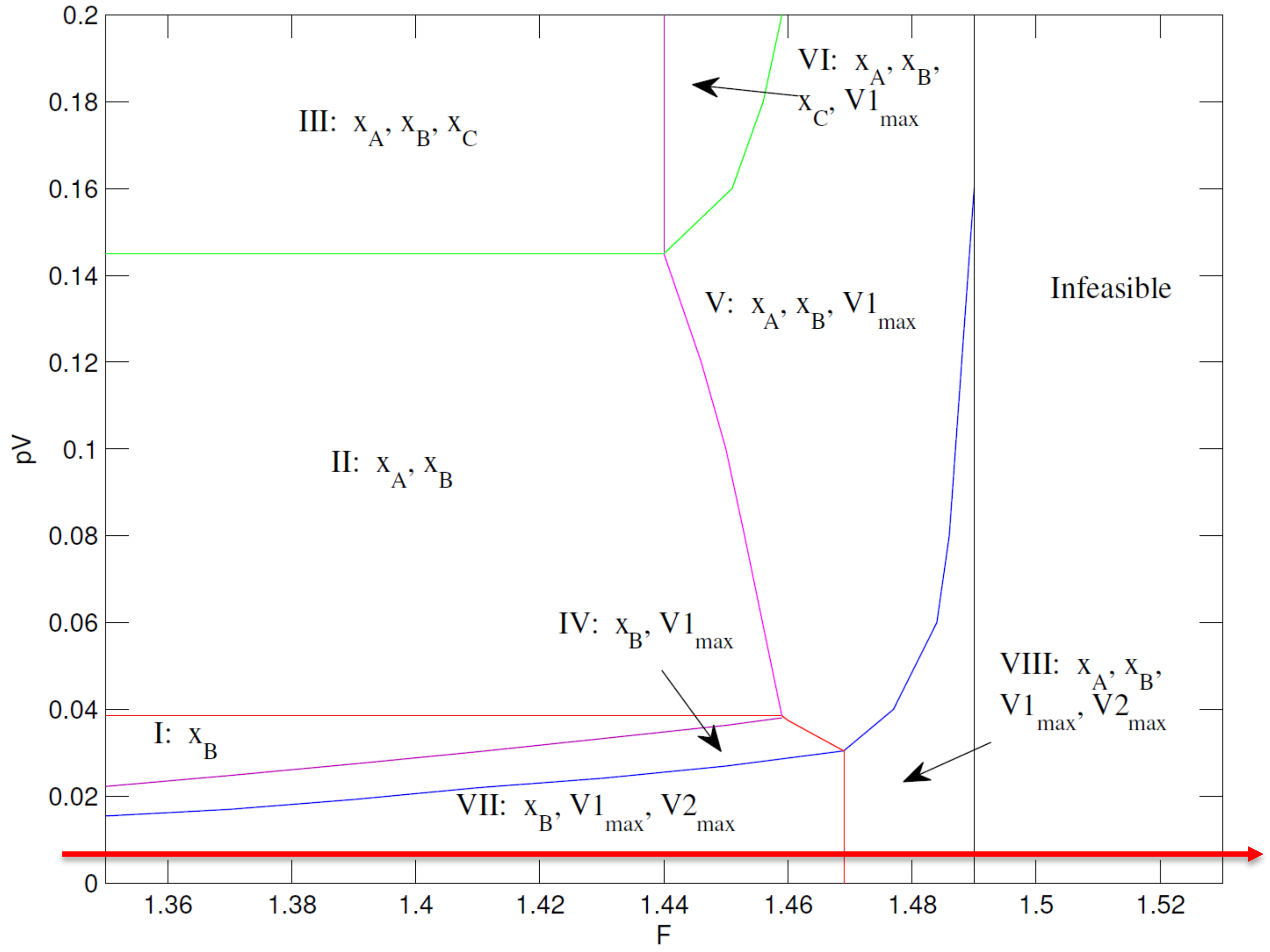
x_B always active
$2^4 = 16$

-1 = 15

In practice = 8

This seems complicated….. But knowledge about all regions is rarely (if ever) needed….
In practice: We use the control system to switch when constraints are encountered…..
It's much simpler and in many cases optimal…. Try

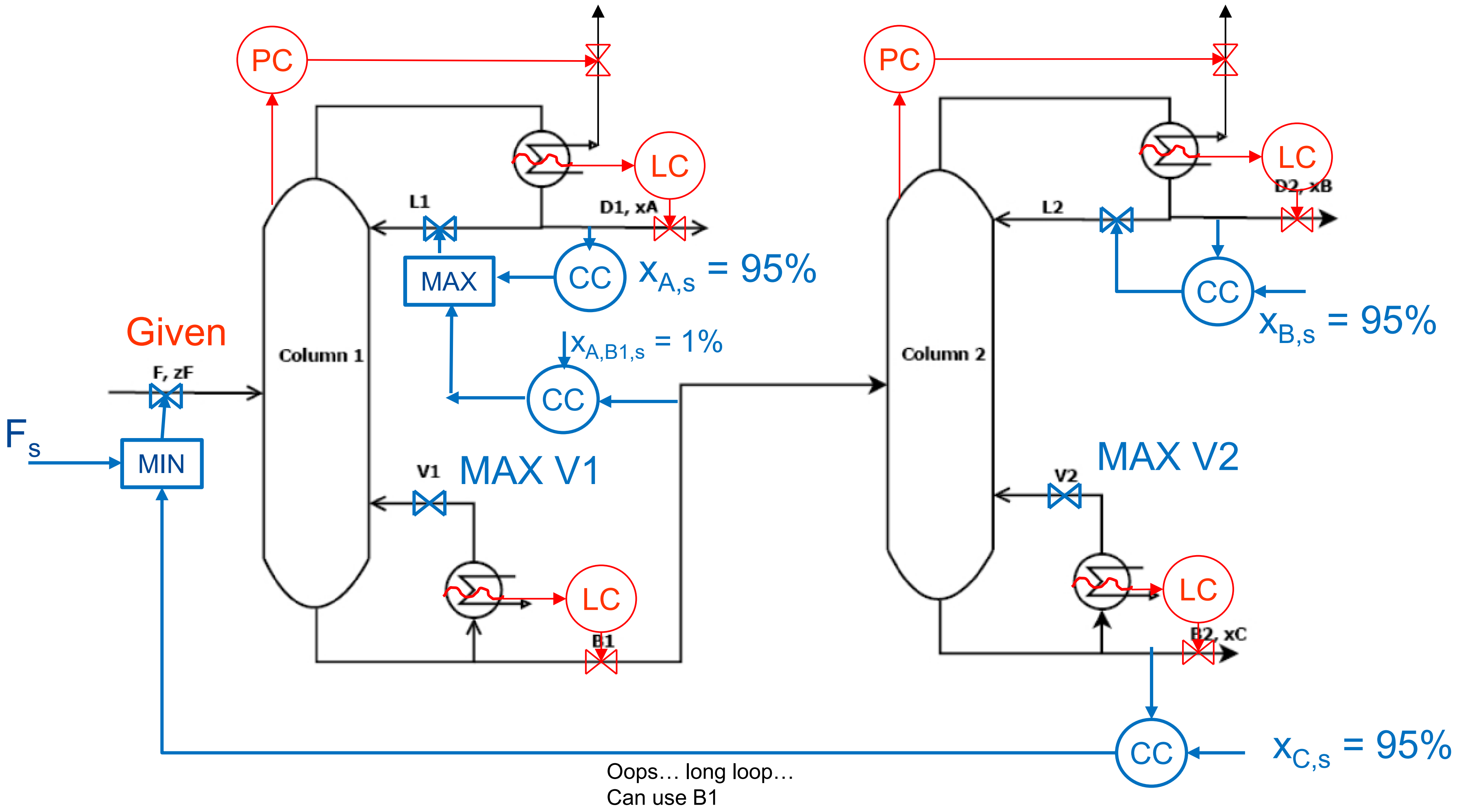
Preview: How handle increase in F (still with low pV)?



How to control in three regions (VII, VIII and Infeasible)?

Preview: Control of distillation columns in series in three regions

(but finding a simple control structure with constant setpoints that works in all regions is not possible;
One solution: 4 composition loops + RTO that optimizes composition setpoints)



Red: Basic regulatory loops