Economic process control Introduction

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Nitin Kaitsha, IIT Kanpur

Shortcourse PSE Asia 2019: Economic process control

13 January 2019 from 9:00-17:00 at Faculty of Engineering, Chulalongkorn University

Presenters:
Professor Sigurd Skogestad, NTNU
Professor Nitin Kaistha, IIT Kanpur (case studies)

This workshop focuses on the big picture when designing a control system for process plants. The overall objectives is to maximize the operational profit. This should also be the basis for designing the control system, all the way from PID controllers at the bottom of the control hierarchy to the possible RTO layer at the top.

09.00 Session 1. Overview of economic plantwide control
- Control degrees of freedom
- Our paradigm: Hierarchical decision (control) layer
- What is optimal operation?
- Distillation examples

10.15 Coffee break

10.30 Session 2. What to control? (Selection of primary controlled variables based on economics)
- Degrees of freedom
- Optimization
- Active constraints
- Self-optimizing control

11.30 Session 3. Where to set the production rate and bottleneck
- Inventory control
- Consistency
- Radiation rule
- Location of throughput manipulator (TPM)

12.00 Lunch

13.00 Session 4. Design of the regulatory control layer ("what more should we control")
- PID control, Stabilization
- Secondary controlled variables (measurements)
- Pairing with inputs

14.00 Case studies (1415: Coffee break)

15.00 Session 5 Design of supervisory control layer
- Cascade control
- Switching between active constraints
    - MV saturation (SIMO controllers) : Split range control / Controllers with different setpoints
    - CV changes: Selectors
- Decentralized versus centralized (MPC)

Session 6. Real-time optimization (see plenary tomorrow)
Alternative implementations:
- Conventional RTO
- Dynamic RTO / Economic MPC
- Combination with Self-optimizing control
- Extremum seeking control / NCO tracking / Hill climbing

16.00 Session 7 Case studies (throughout the day)

Part 1. Plantwide control

Introduction to plantwide control (what should we really control?)

Introduction.

- 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) and supervisory (advanced) control layer

Optimal operation (economics)

- Define cost J and constraints
- Active constraints (as a function of disturbances)
- Selection of economic controlled variables (CV1). Self-optimizing variables.

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Sigurd Skogestad

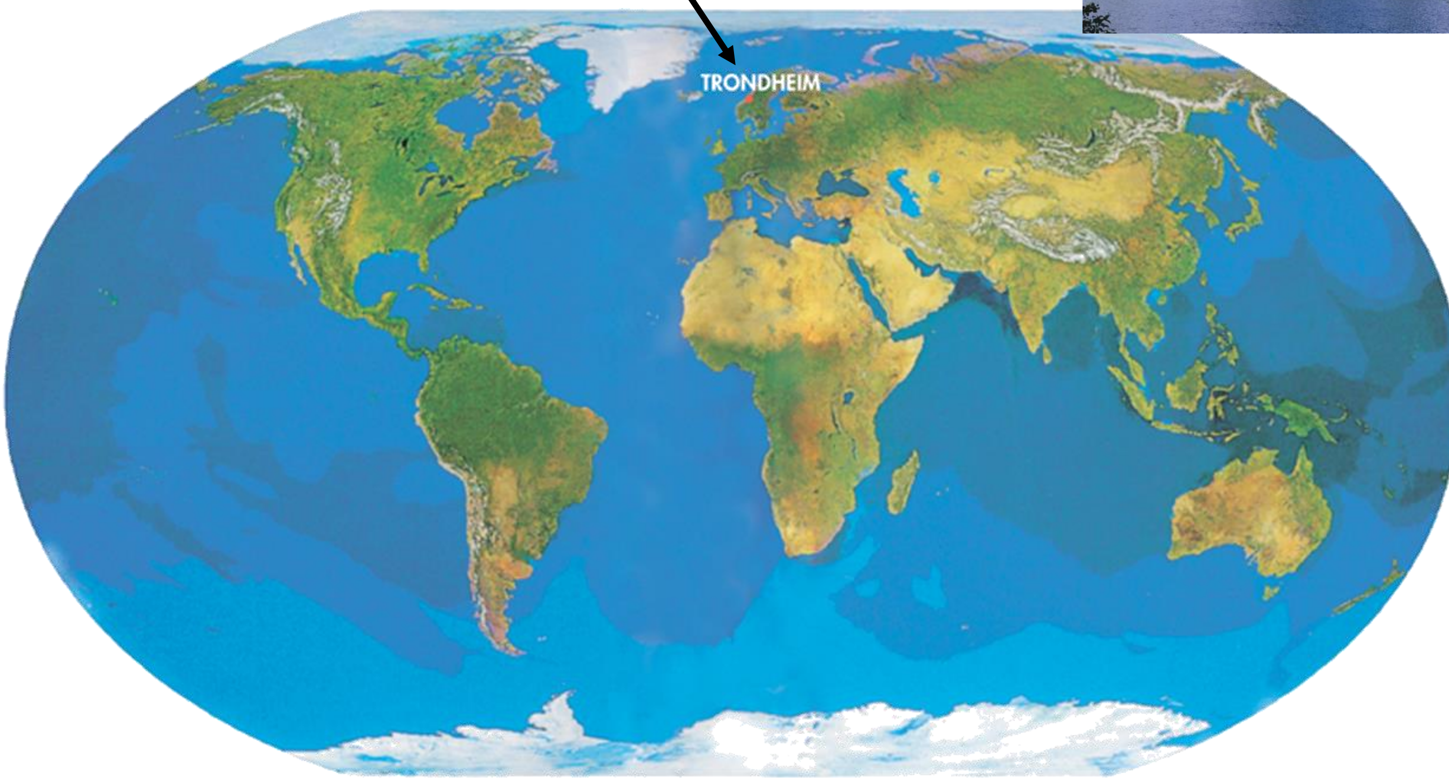
- 1955: Born in Flekkefjord, Norway
- 1978: MS (Siv.ing.) in chemical engineering at NTNU
- 1979-1983: Worked at Norsk Hydro co. (process simulation)
- 1987: PhD from Caltech (supervisor: Manfred Morari)
- 1987-present: Professor of chemical engineering at NTNU
- 1999-2009: Head of Department
- 2015-..: Director SUBPRO (Subsea research center at NTNU)

- Book: Multivariable Feedback Control (Wiley 1996; 2005)
  - 1989*: Ted Peterson Best Paper Award* by the CAST division of AIChE
  - 1990*: George S. Axelby Outstanding Paper Award* by the Control System Society of IEEE
  - 1992*: O. Hugo Schuck Best Paper Award* by the American Automatic Control Council
  - 2006: *Best paper award* for paper published in 2004 in *Computers and chemical engineering*.
  - 2011: *Process Automation Hall of Fame* (US)
  - 2012: *Fellow of American Institute of Chemical Engineers (AIChE)*
  - 2014: *Fellow of International Federation of Automatic Control (IFAC)*

Nitin Kaistha

- Professor at IIT Kanpur

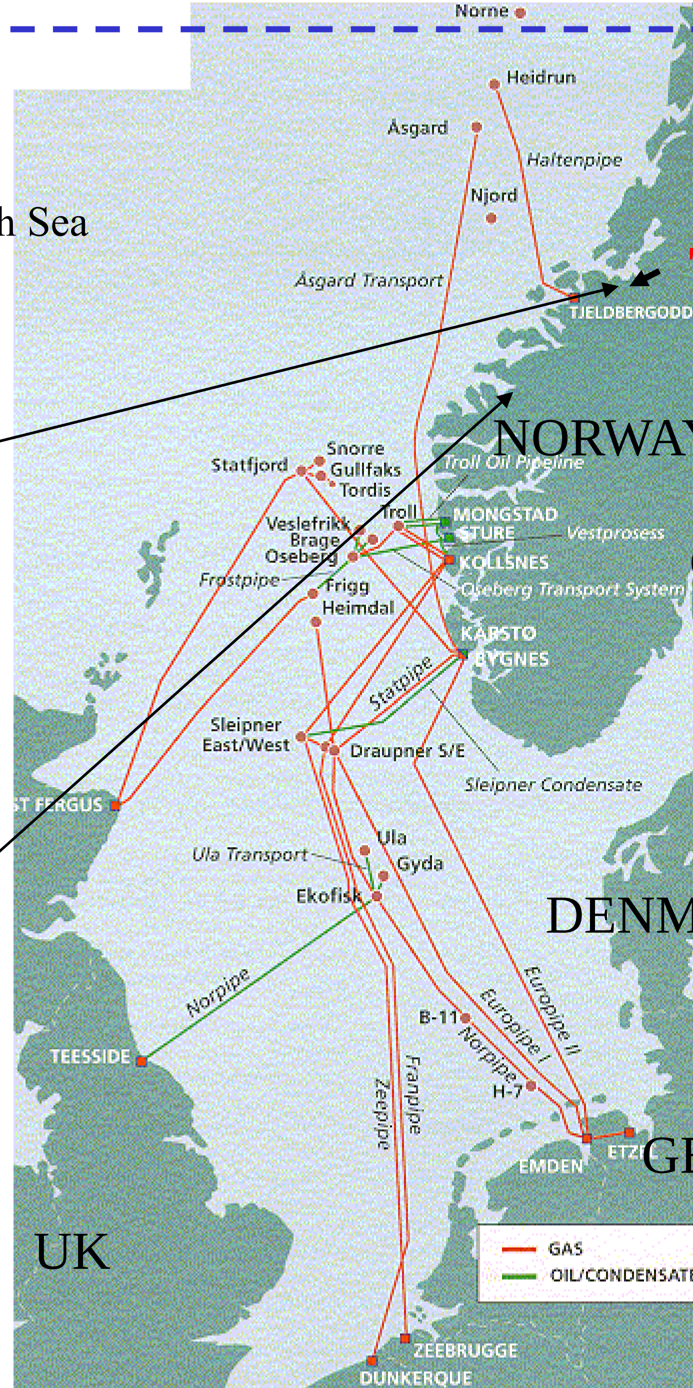
Trondheim, Norway



North Sea

Arctic circle

Trondheim



NORWAY

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UK

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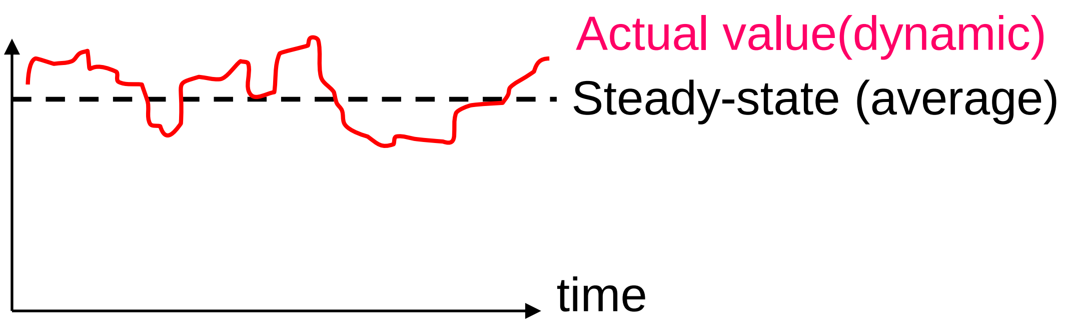
Why control?

- *Operation*



In practice never steady-state:

- Feed changes
- Startup
- Operator changes
- Failures
- …..

"Disturbances" (d's)

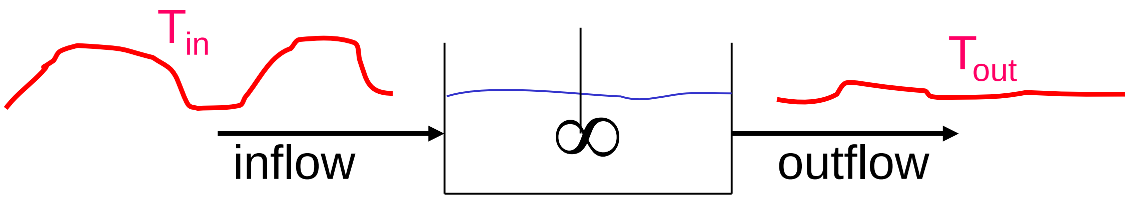
- Control is needed to reduce the effect of disturbances
- 30% of investment costs are typically for instrumentation and control

Countermeasures to disturbances (I)

I. Eliminate/Reduce the disturbance

(a) Design process so it is insensitive to disturbances

- Example: Use buffertank to dampen disturbances



(b) Detect and remove source of disturbances

- "Statistical process control"
- Example: Detect and eliminate variations in feed composition

Countermeasures to disturbances (II)

II. Process control

Do something (usually manipulate valve)
to **counteract** the **effect** of the disturbances

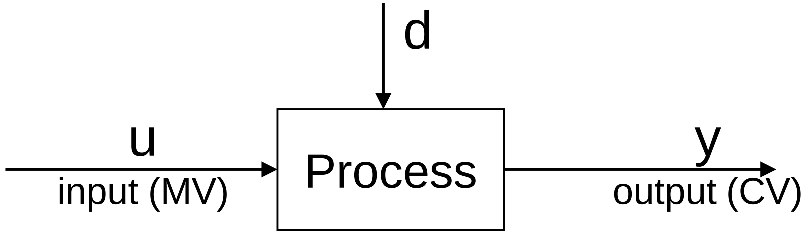
(a) Manual control: Need operator
(b) Automatic control: Need measurement + automatic valve + computer

Goals automatic control:

- Smaller variations
  - more consistent quality
  - More optimal
- Smaller losses (environment)
- Lower costs
- More production

Industry: Still large potential for improvements!

Classification of variables



Independent variables ("the cause"):

(a) Inputs (MV, u): Variables we can adjust (valves)

(b) Disturbances (DV, d): Variables outside our control

Dependent (output) variables ("the effect or result"):

(c) Primary outputs (CVs, y_1): Variables we want to keep at a given setpoint

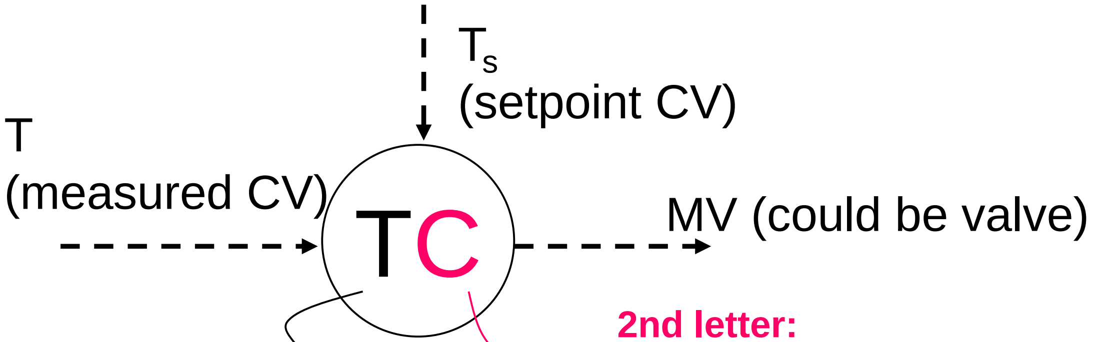
(d) Secondary outputs (y_2): Extra measurements that we may use to improve control

Inputs for control (MVs)

- Usually: Inputs (MVs) are valves.
  - Physical input is valve position (z), but we often simplify and say that flow (q) is input

- Valve Equation $q(\mathrm{m}^3/\mathrm{s}) = C_v f(z)\sqrt{\Delta p/\rho}$

Notation feedback controllers (P&ID)

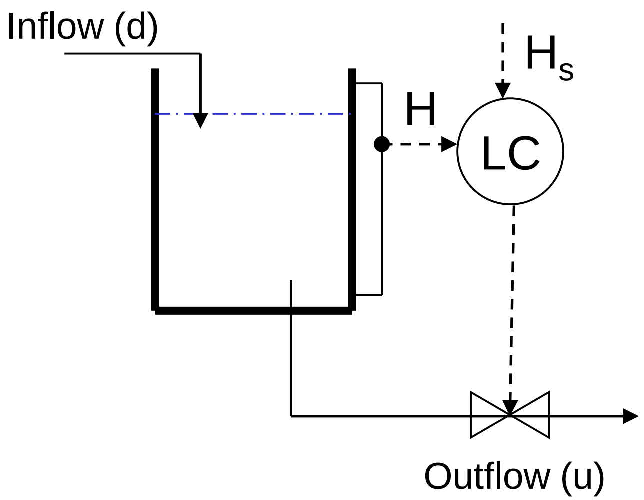


2nd letter:
C: controller
I: indicator (measurement)
T: transmitter (measurement)

1st letter: Controlled variable (CV) = What we are trying to control (keep constant)

- **T: temperature**
- **F: flow**
- **L: level**
- **P: pressure**
- DP: differential pressure (Δp)
- A: Analyzer (composition)
- C: composition
- X: quality (composition)
- H: enthalpy/energy

Example: Level control



CLASSIFICATION OF VARIABLES:

INPUT (u): OUTFLOW (Input for control!)
OUTPUT (y): LEVEL
DISTURBANCE (d): INFLOW

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

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

Plantwide control = Control structure 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.).
- That is: **Control structure design** includes all the decisions we need make to get from ``PID control'' to "PhD" control

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Main objectives 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)

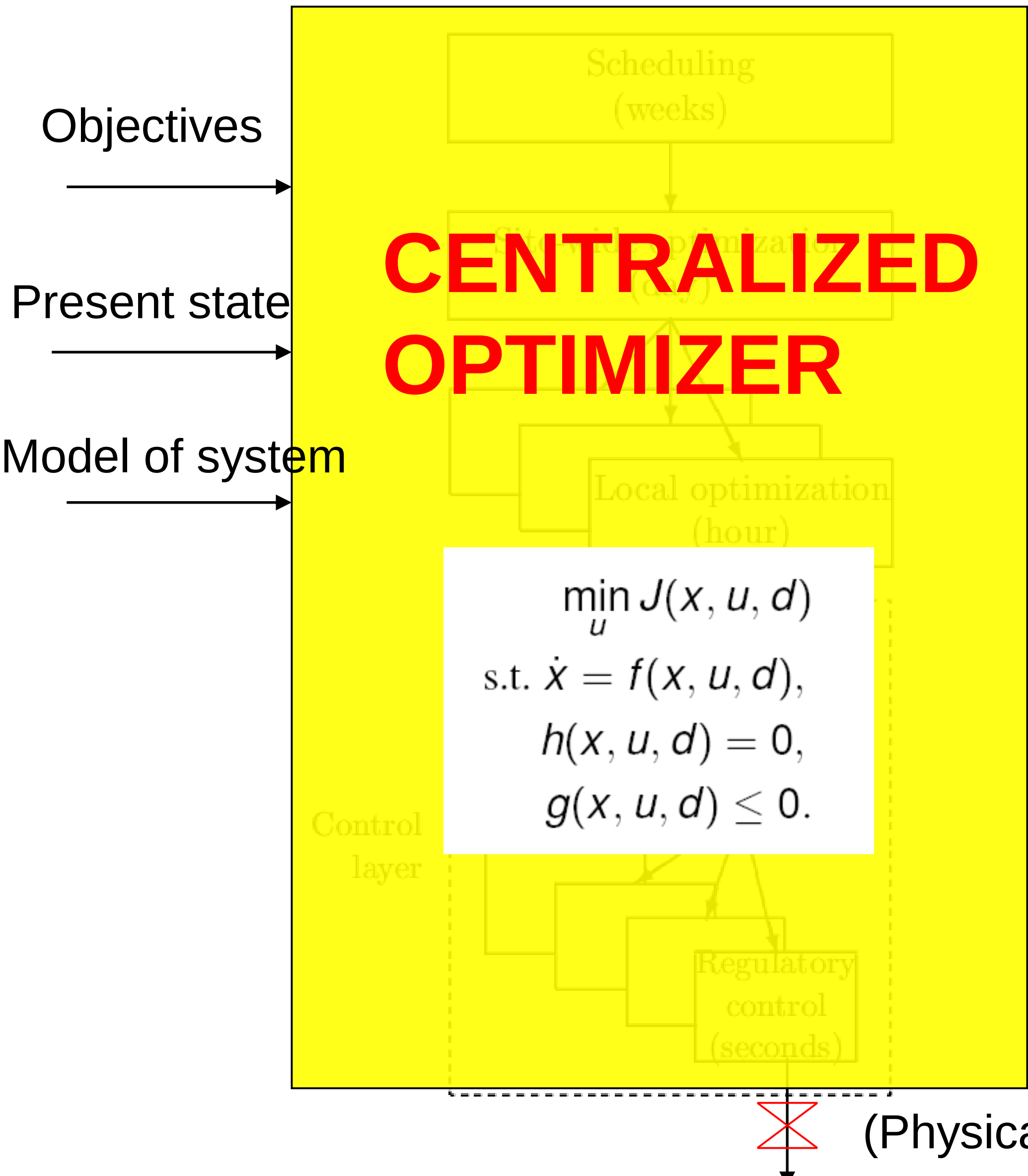
Optimal operation (economics)

Example of systems we want to operate optimally

- Process plant
  - minimize J=economic cost
- Runner
  - minimize J=time
- «Green» process plant
  - Minimize J=environmental impact (with given economic cost)

- General multiobjective:
  - Min J (scalar cost, often $)
  - Subject to satisfying constraints (environment, resources)

Theory: Optimal operation



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

Theory:
- Model of overall system
- Estimate present state
- Optimize all degrees of freedom

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

Process control:
- Excellent candidate for centralized control

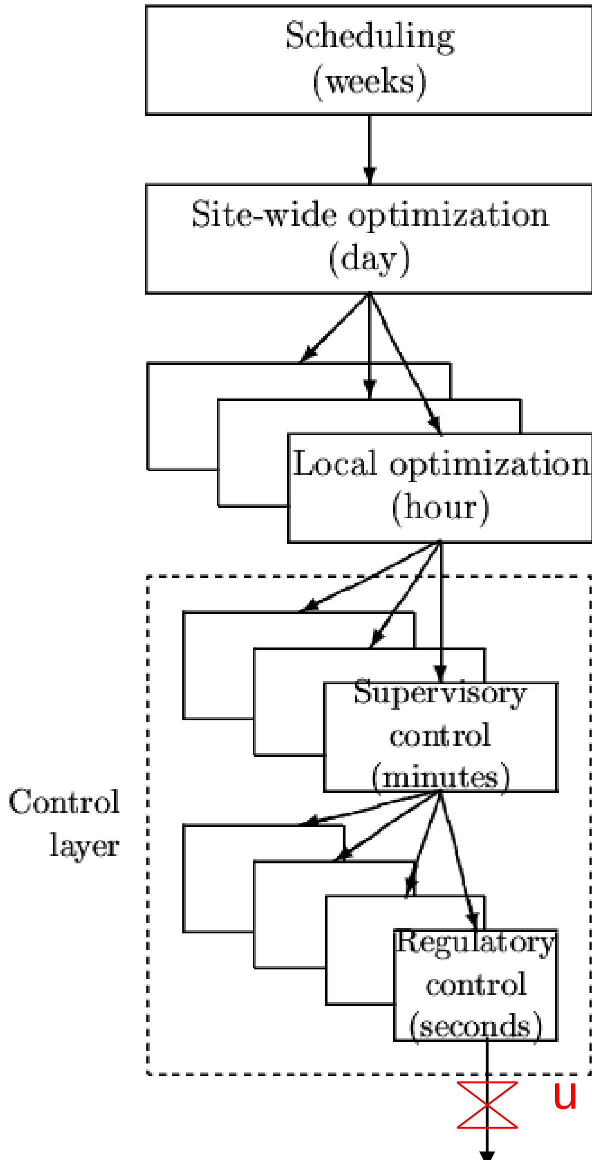
Practice: Engineering systems

- 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

Same in biological systems

Practical operation: Hierarchical structure



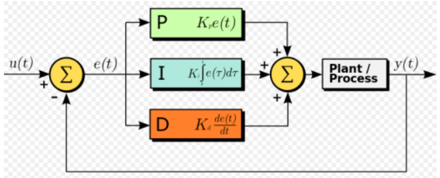
Manager

Process engineer

Operator/RTO

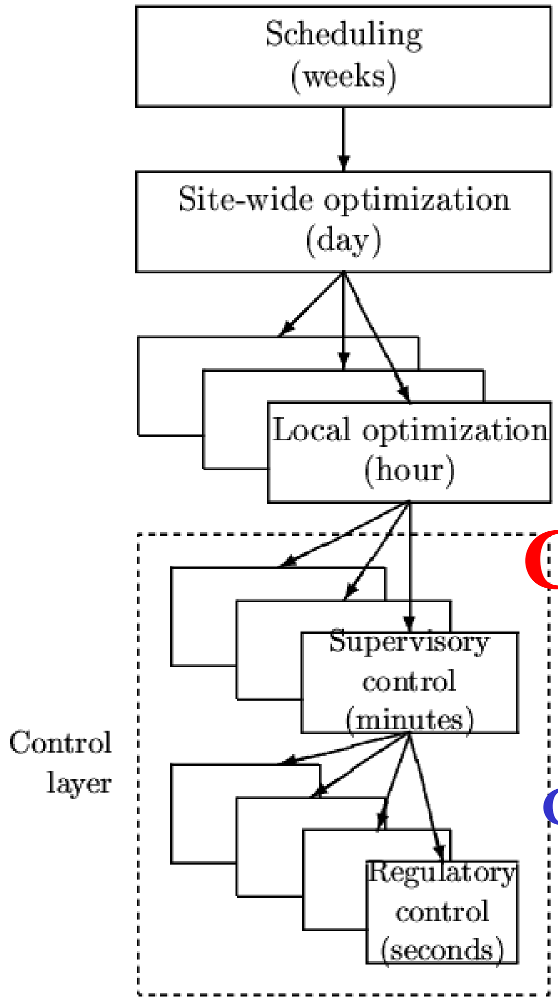
Operator/"Advanced control"/MPC

PID-control



u = valves

Translate optimal operation into simple control objectives: What should we control?



$CV_1 = c$? (economics)

$CV_2 =$? (stabilization)

Outline

- Skogestad procedure for control structure design

  I Top Down

  - <u>Step S1</u>: Define operational objective (cost) and constraints
  - <u>Step S2:</u> Identify degrees of freedom and optimize operation for disturbances
  - <u>Step S3</u>: Implementation of optimal operation
    - What to control ? (CV1) (active constraints, self-optimizing variables)
  - <u>Step S4:</u> Where set the production rate? (Inventory control)

  II Bottom Up

  - <u>Step S5</u>: Regulatory control: What more to control to stabilize (CV2) ?
  - <u>Step S6</u>: Supervisory control: Control CV1 and keep feaiable oparationm
  - <u>Step S7:</u> Real-time optimization

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Step S1. Define optimal operation (economics)

- What are we going to use our degrees of freedom (u=MVs) for?
- Typical cost function*:

J = cost feed + cost energy – value products

- *No need to include fixed costs (capital costs, operators, maintainance) at ”our” time scale (hours)
- Note: J=-P where P= Operational profit

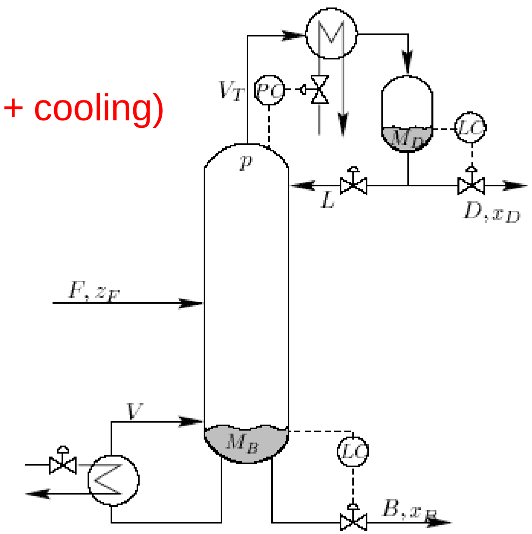
Optimal operation distillation column

- 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 $x_{D,\text{ impurity}} < \max$

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

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

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

  Pressure: p has given setpoint (can be given up, but need $p_{min} < p < p_{max}$)

  Feed: F has given setpoint (can be given up)

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

Step S2. Optimize

(a) Identify degrees of freedom

(b) Optimize for expected disturbances

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

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Step S2a: Degrees of freedom (DOFs) for operation

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

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Steady-state degrees of freedom (DOFs)

IMPORTANT!

DETERMINES THE NUMBER OF VARIABLES TO CONTROL!

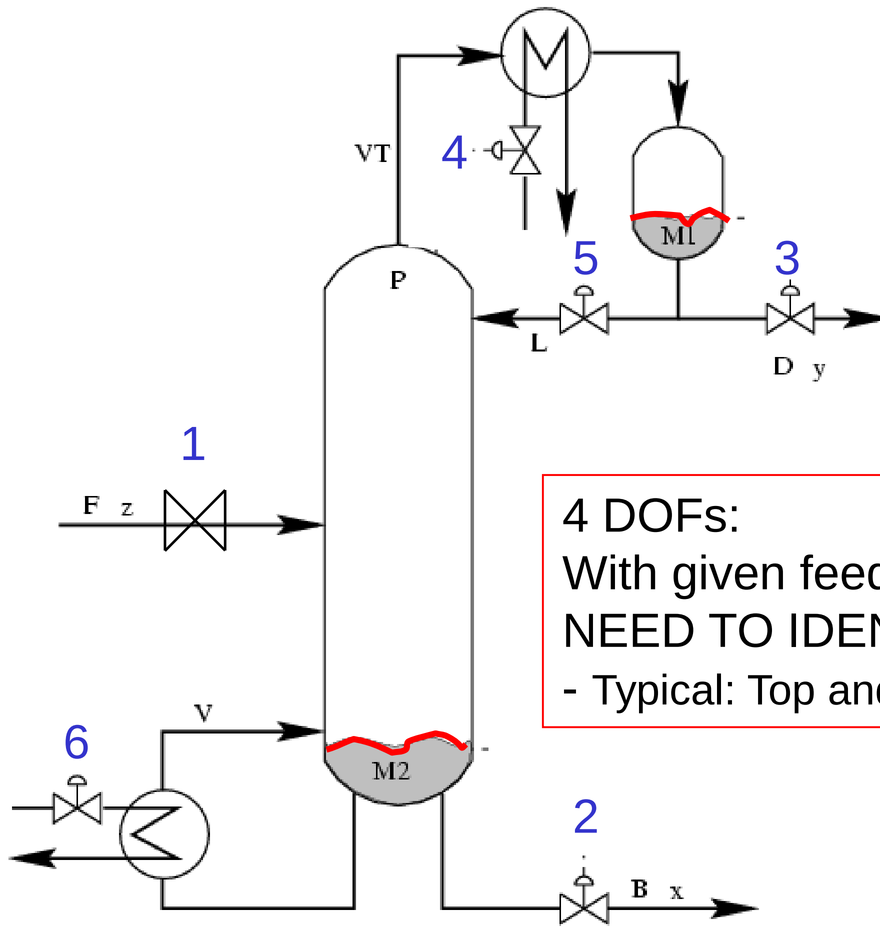
- **No. of primary CVs = No. of steady-state DOFs**

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}: Equality constraints (normally 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

Typical Distillation column



4 DOFs:
With given feed and pressure:
NEED TO IDENTIFY 2 more CV's
- Typical: Top and btm composition

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

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

$^*N_{0y}$: no. controlled variables (liquid levels) with no steady-state effect

Step S2b: Optimize for expected disturbances

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

$$J = \text{cost feed} + \text{cost energy} - \text{value products}$$

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

$$\min_u \; J(u,x,d)$$

subject to:

Model equations (e,g, Hysys): $f(u,x,d) = 0$

Operational constraints: $g(u,x,d) < 0$

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 use Matlab or even Excel "on top"

…. BUT A GOOD ENGINEER CAN OFTEN GUESS THE SOLUTION (active constraints)

Step S3: Implementation of optimal operation

- Have found 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)

Objective: Move optimization into control layer

Optimal operation of runner

- Cost to be minimized, J=T
- One degree of freedom (u=power)
- What should we control?

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1. Optimal operation of Sprinter

- 100m. J=T
- Active constraint control:
  - Maximum speed ("no thinking required")
  - CV = power (at max)

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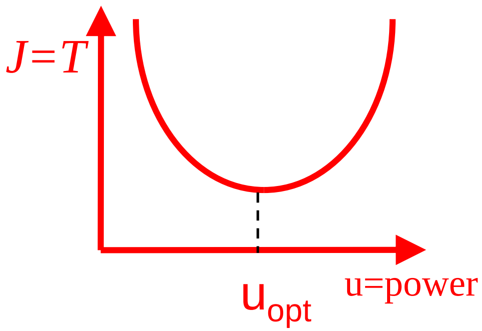
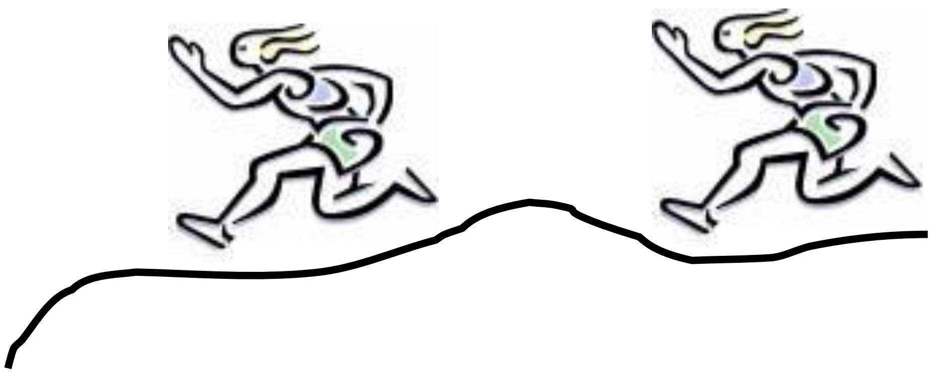
2. Optimal operation of Marathon runner

- 40 km. J=T
- What should we control? CV=?
- Unconstrained optimum





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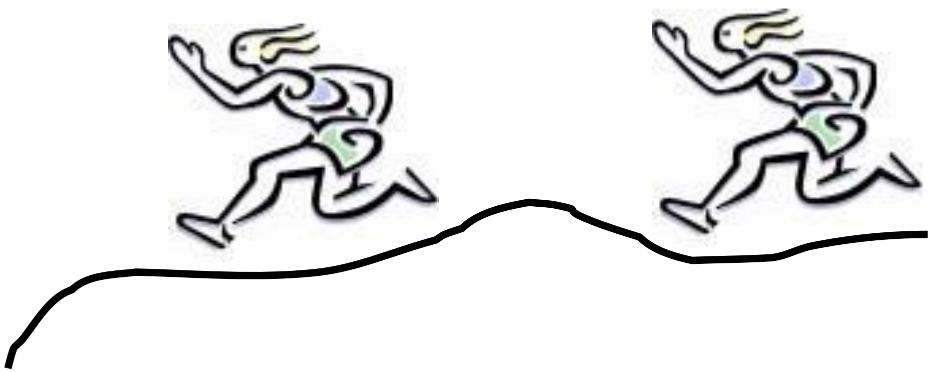
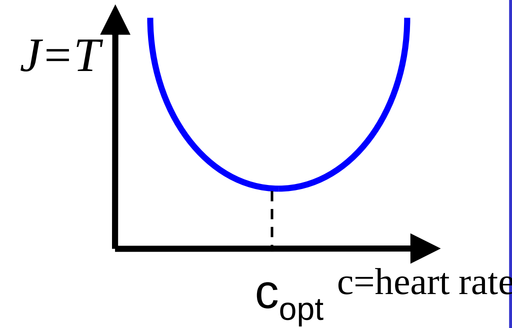
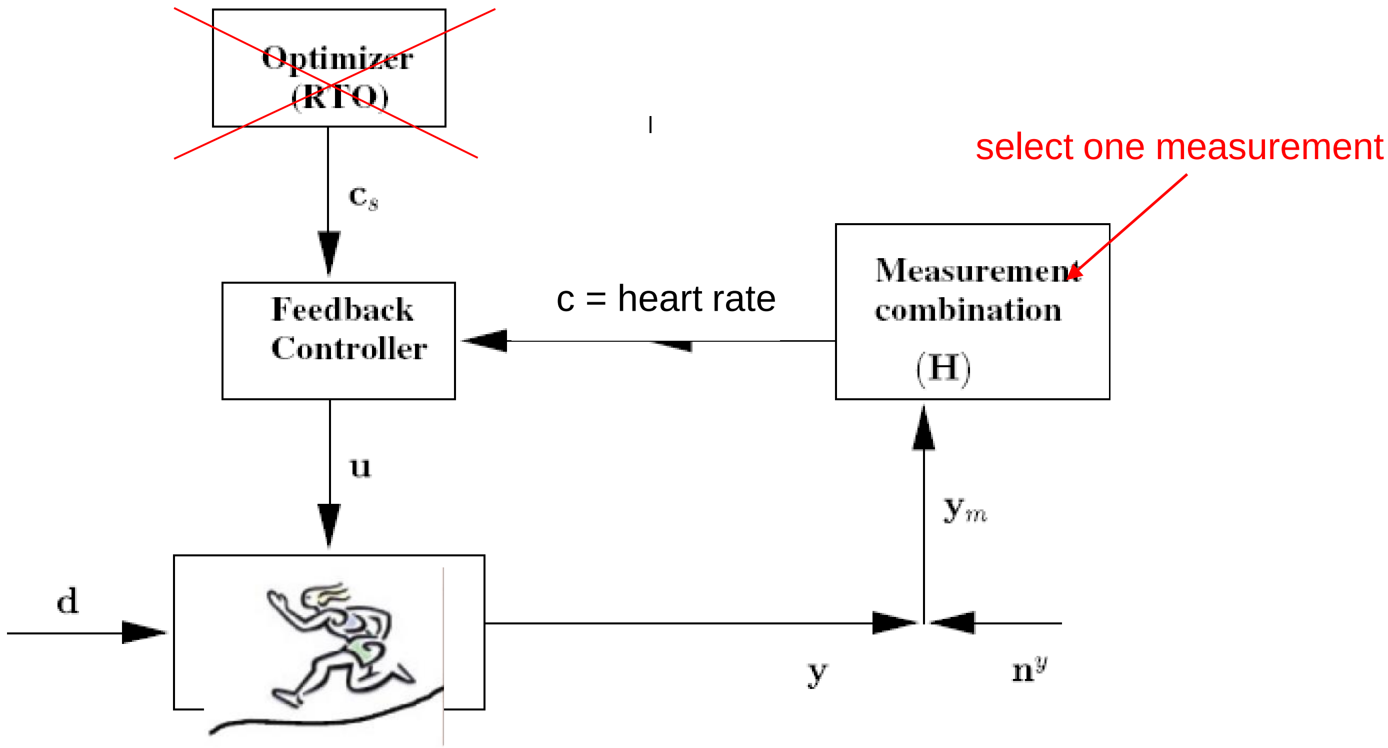
Self-optimizing control: Marathon (40 km)

- Any self-optimizing variable (to control at constant setpoint)?
    - c_1 = distance to leader of race
    - c_2 = speed
    - c_3 = heart rate
    - c_4 = level of lactate in muscles



Conclusion Marathon runner





- CV = heart rate is good "self-optimizing" variable
- Simple and robust implementation
- Disturbances are indirectly handled by keeping a constant heart rate
- May have infrequent adjustment of setpoint (c_s)

Step 3. What should we control (c)?

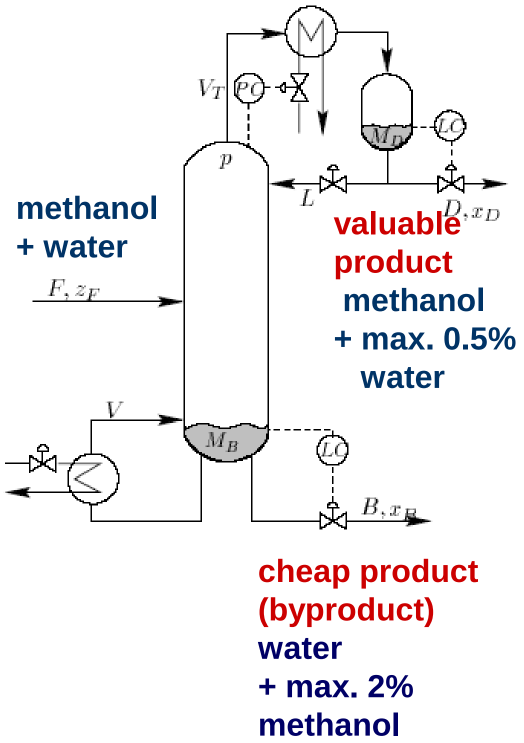
(primary controlled variables $y_1=c$)

Selection of controlled variables c

1. **Control active constraints!**
2. **Unconstrained variables: Control self-optimizing variables!**

Expected active constraints distillation

- **Both products (D,B) generally have purity specs**
- **Valuable product: Purity spec. always active**
  - Reason: Amount of valuable product (D or B) should always be maximized
    - Avoid product “give-away” (“Sell water as methanol”)
    - Also saves energy



Control implications:

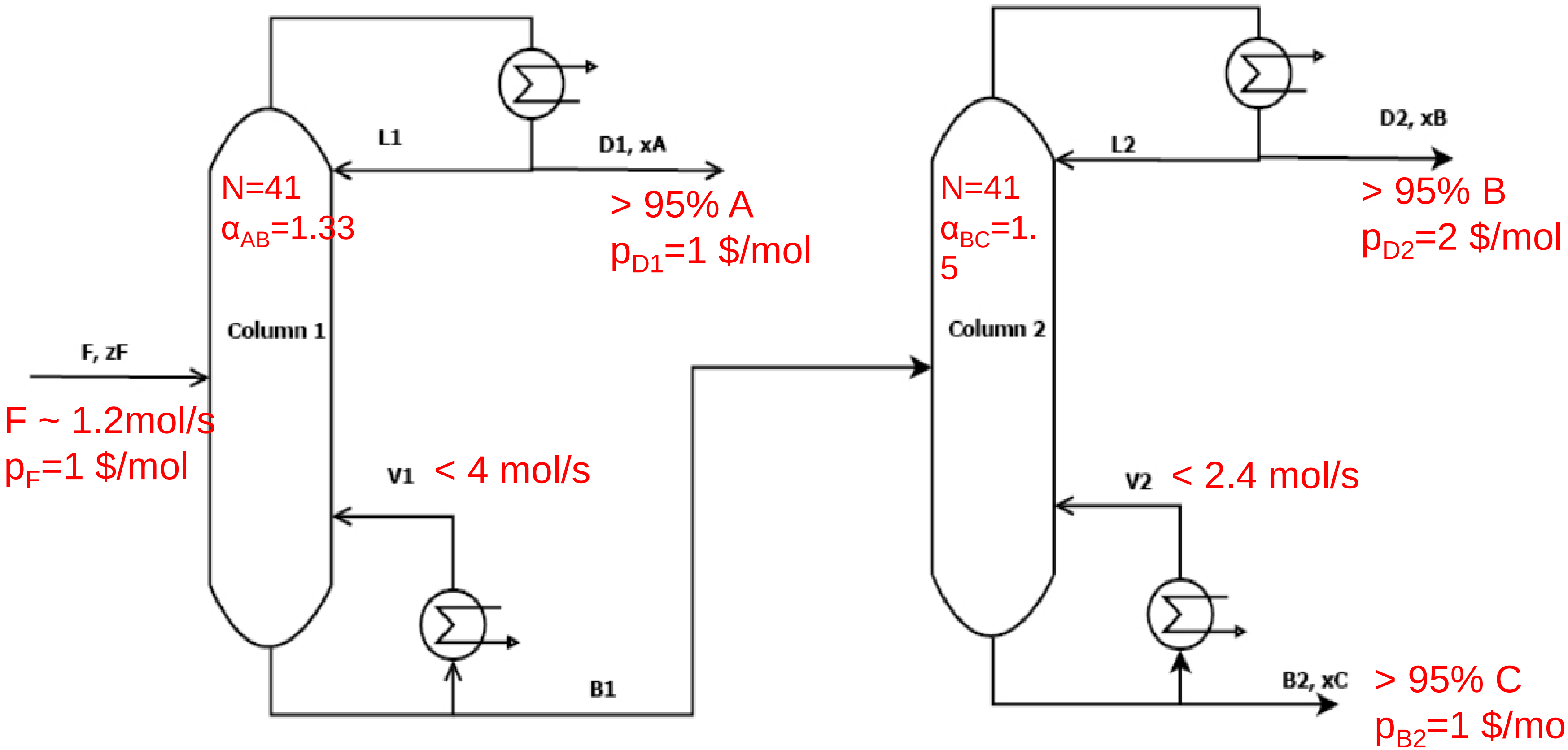
1. ALWAYS Control valuable product at spec. (active constraint).
2. May overpurify (not control) cheap product

Example with Quiz: Optimal operation of two distillation columns in series

QUIZ 1

Operation of Distillation columns in series

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



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)

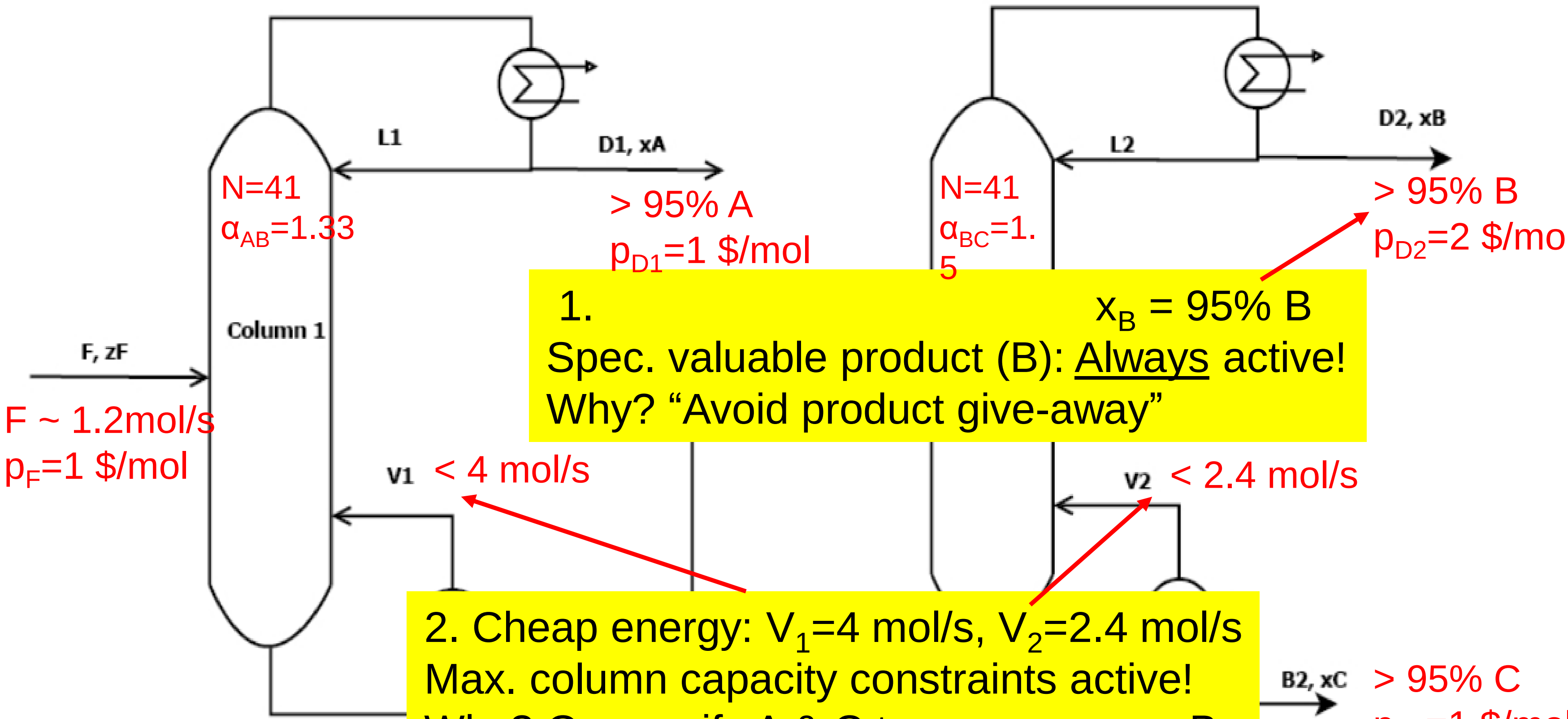
QUIZ: What are the expected active constraints?
1. Always. 2. For low energy prices.

DOF = Degree Of Freedom

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

Operation of Distillation columns in series

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



1. x_B = 95% B
Spec. valuable product (B): <u>Always</u> 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)

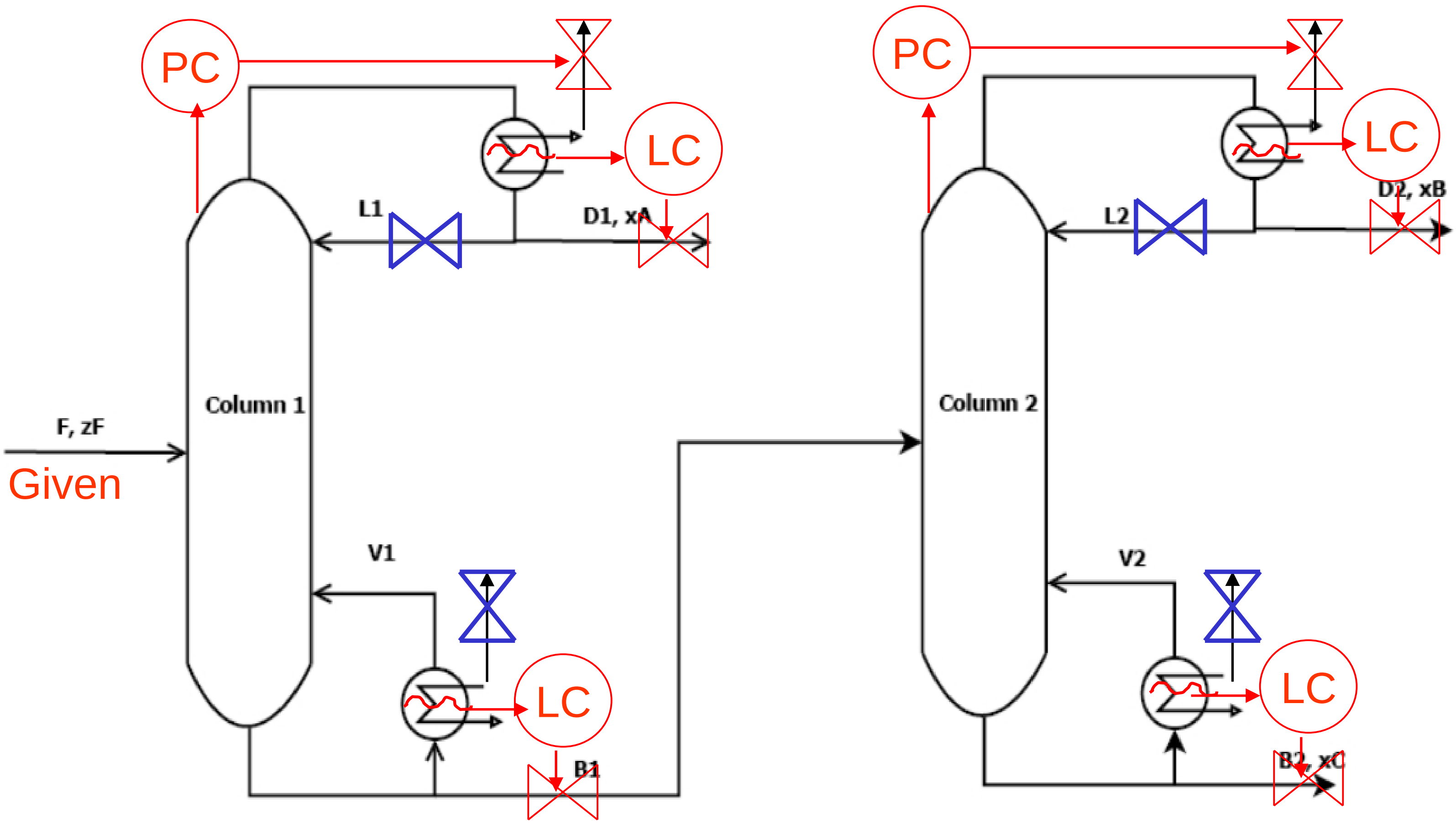
QUIZ: What are the expected active constraints?
1. Always. 2. For low energy prices.

DOF = Degree Of Freedom

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

QUIZ 2

Control of Distillation columns in series

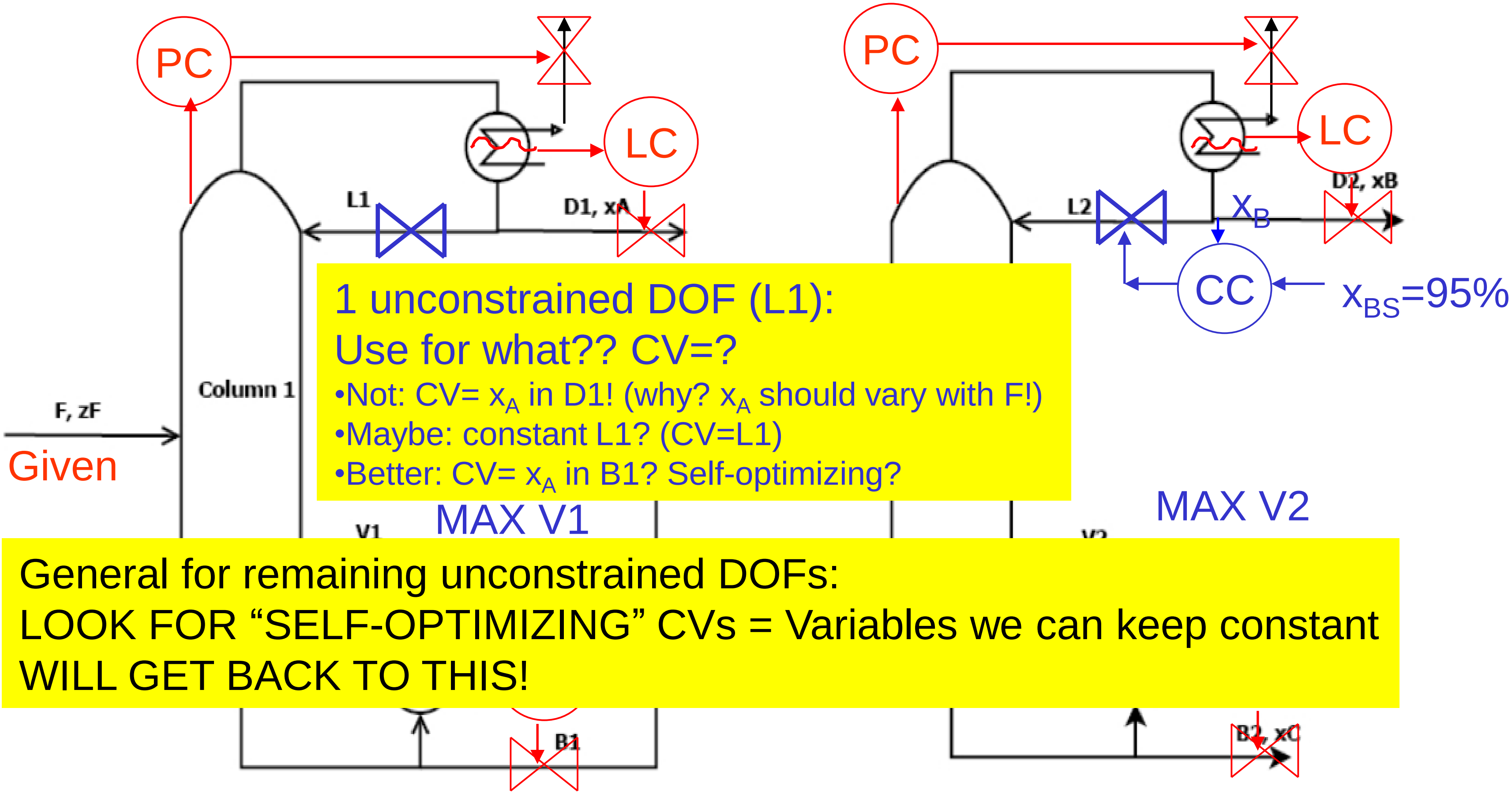


QUIZ. Assume low energy prices (pV=0.01 $/mol).
How should we control the columns?
HINT: CONTROL ACTIVE CONSTRAINTS

Red: Basic regulatory loops

SOLUTION QUIZ 2

Control of Distillation columns in series



1 unconstrained DOF (L1):
Use for what?? CV=?

- 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?

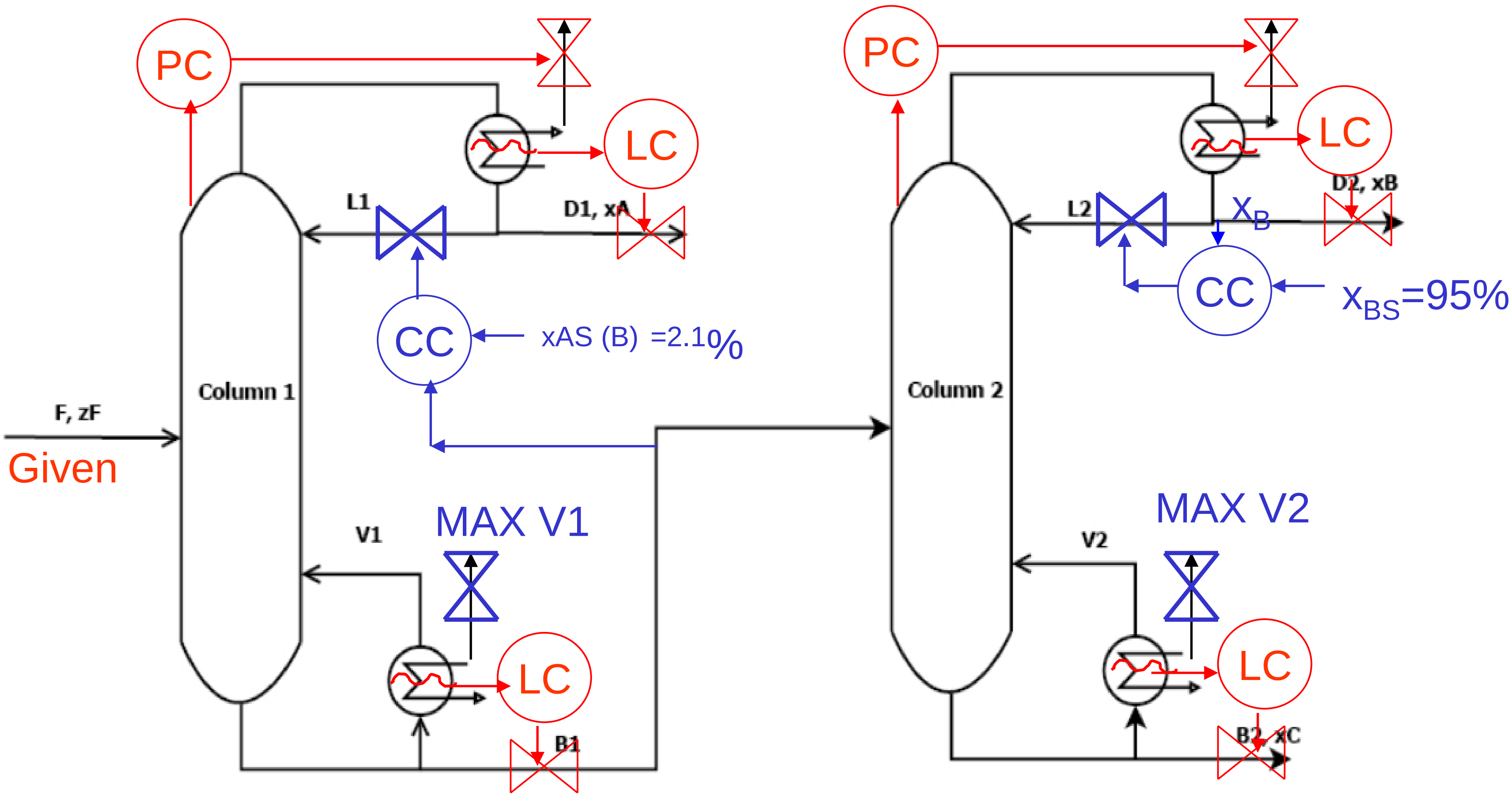
General for remaining unconstrained DOFs:
LOOK FOR “SELF-OPTIMIZING” CVs = Variables we can keep constant
WILL GET BACK TO THIS!

QUIZ. Assume low energy prices (pV=0.01 $/mol).
How should we control the columns?
HINT: CONTROL ACTIVE CONSTRAINTS

Red: Basic regulatory loops

Solution.

Control of Distillation columns. Cheap energy



Comment: Distillation column control in practice

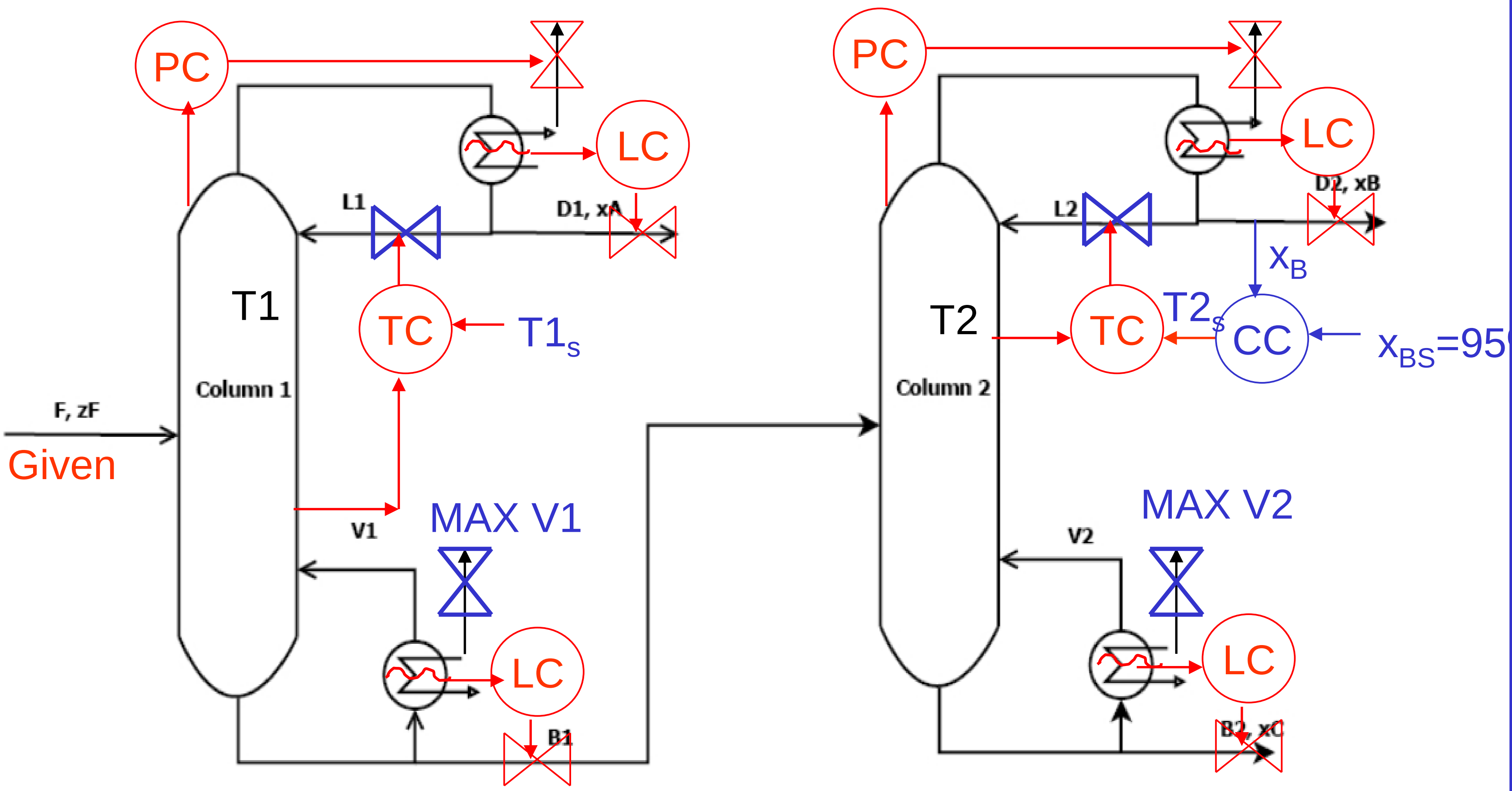
1. Add stabilizing temperature loops
   - In this case: use reflux (L) as MV because boilup (V) may saturate
   - T_{1s} and T_{2s} then replace L_1 and L_2 as DOFs.

2. Replace V1=max and V2=max by DPmax-controllers (assuming max. load is limited by flooding)

- See next slide

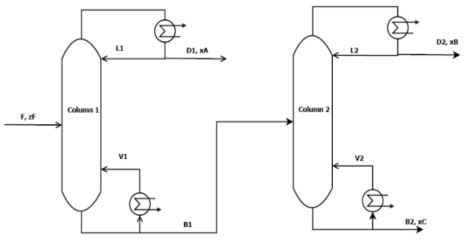
Comment: In practice

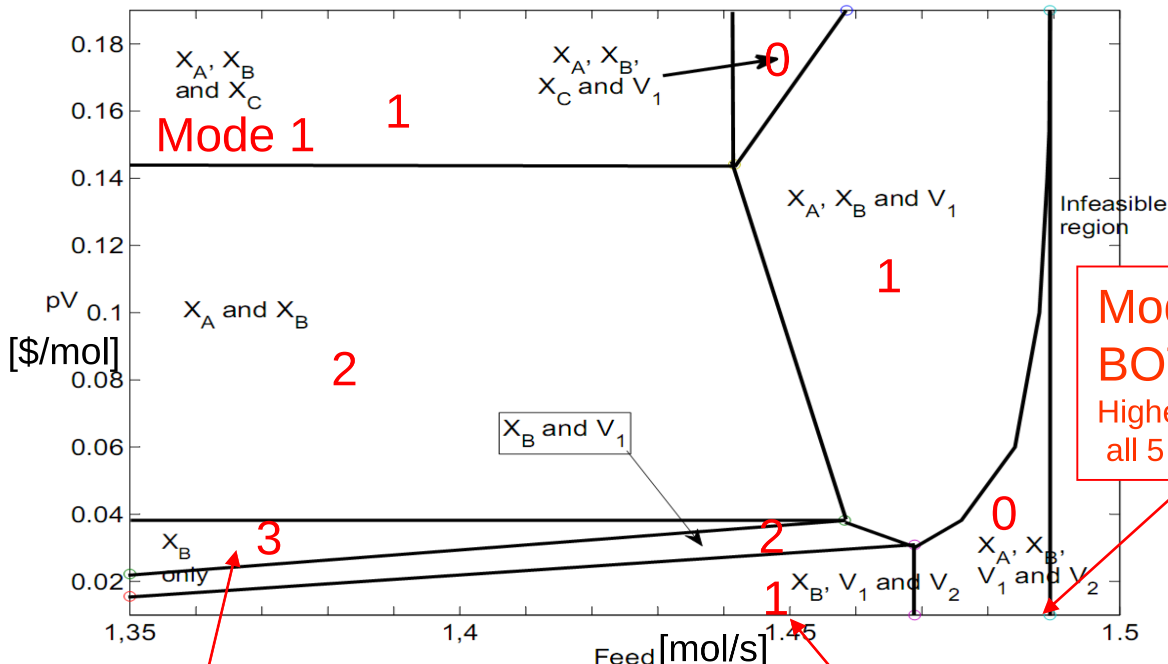
Control of Distillation columns in series



Distillation example: Not so simple

Active constraint regions for two distillation columns in series





Mode 2: operate at BOTTLENECK

Higher F infeasible because all 5 constraints reached

Cheap energy: 1 remaining unconstrained DOF (L1)

-> Need to find 1 additional CVs ("self-optimizing")

More expensive energy: 3 remaining unconstrained DOFs

-> Need to find 3 additional CVs ("self-optimizing")

CV = Controlled Variable

How many active constraints regions?

- Maximum: $$2^{n_c}$$

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

More on: Optimal operation

minimize J = cost feed + cost energy – value products

Two main cases (modes) depending on marked conditions:

Mode 1. Given feedrate

Mode 2. Maximum production

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

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Mode 1. Given feedrate

Amount of products is then usually indirectly given and

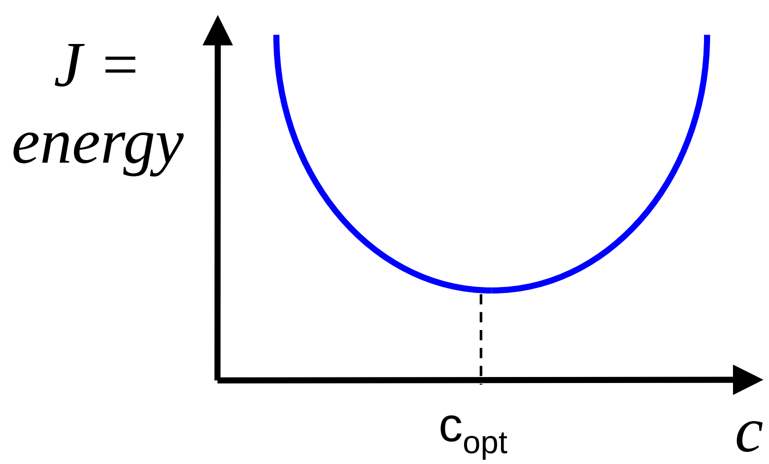
$$J = \underbrace{\text{cost feed} - \text{value products}}_{\text{Often constant}} + \textit{cost energy}$$

Optimal operation is then usually *unconstrained*

"maximize efficiency (energy)"

Control:

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

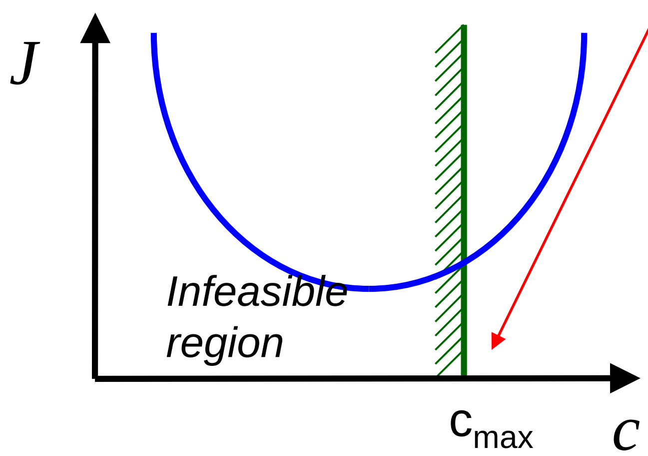


Mode 2. Maximum production

J = cost feed + cost energy – value products

- Assume feedrate 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**

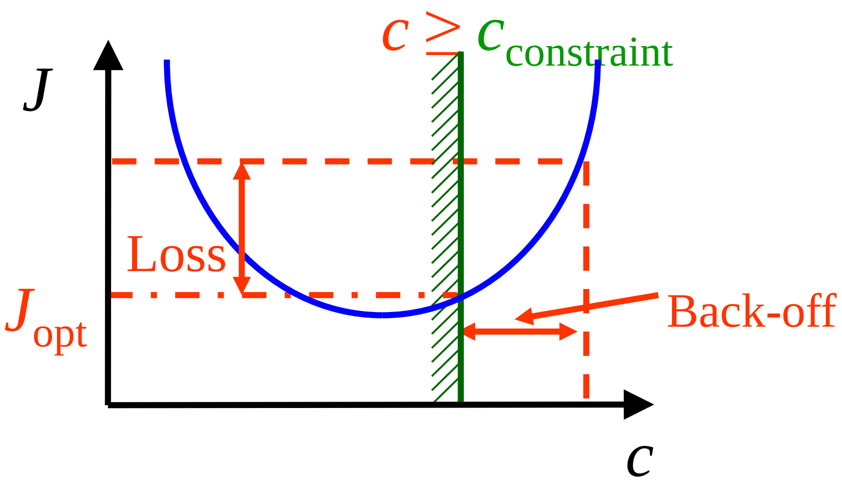
Control:
- Focus on tight control of bottleneck
- “Obvious what to control”
- CV = ACTIVE CONSTRAINT



CV = Active constraint

More on: Active output constraints

Need back-off



a) If constraint can be violated dynamically (only average matters)

- Required Back-off = "measurement bias" (steady-state measurement error for c)

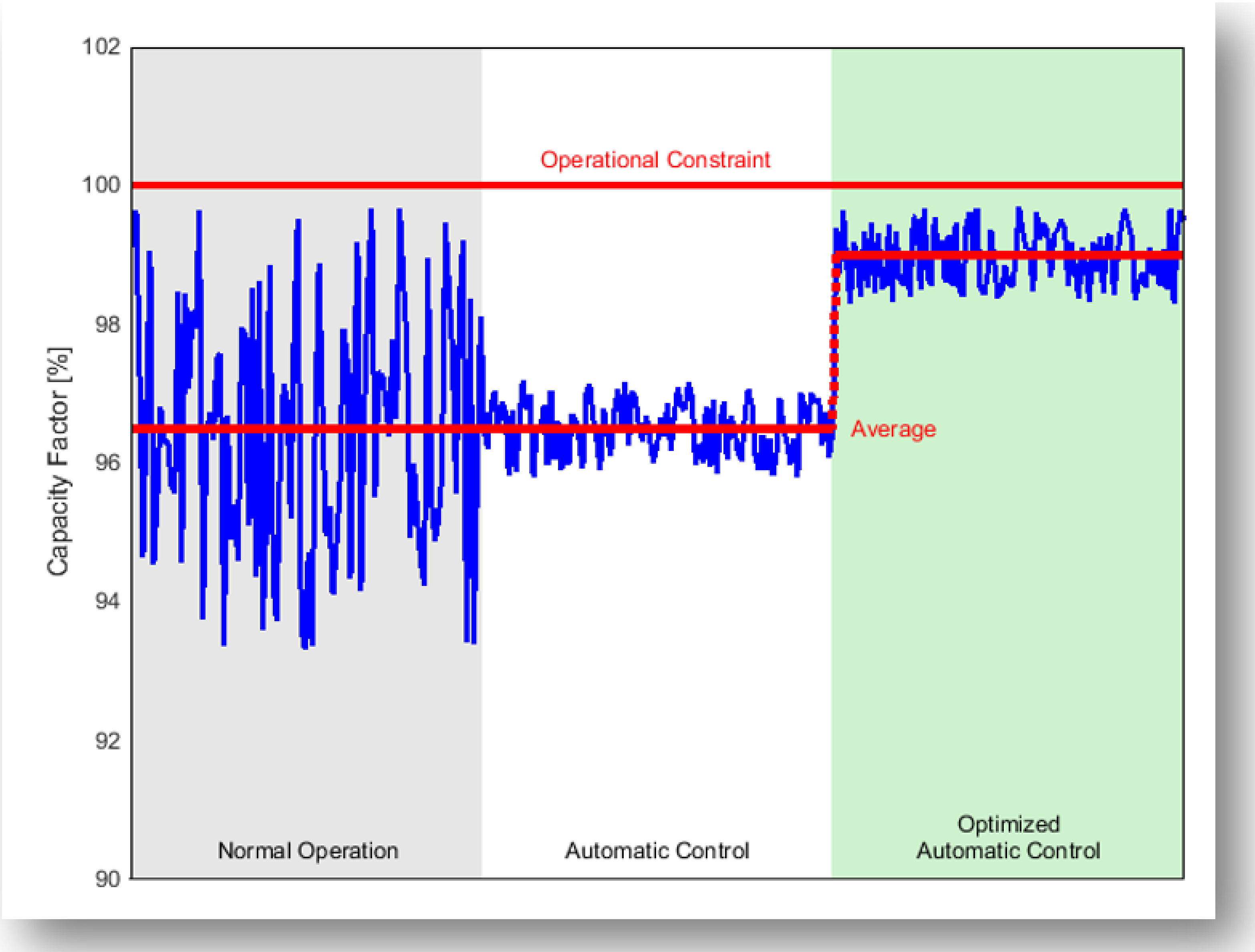
b) If constraint <u>cannot</u> be violated dynamically ("hard constraint")

- 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

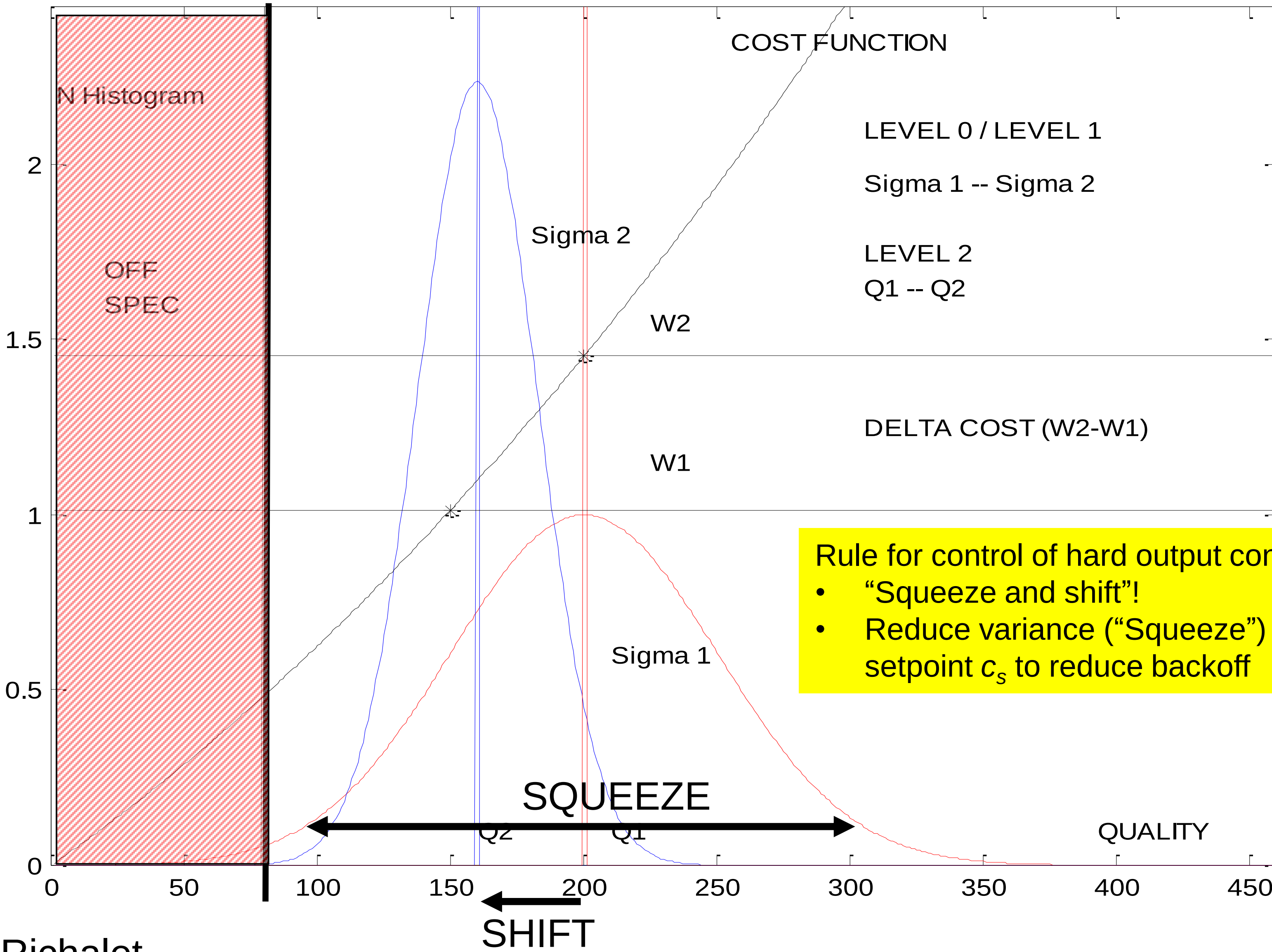
CV = Active constraint

Hard Constraints: «SQUEEZE AND SHIFT»



CV = Active constraint

Hard Constraints: «SQUEEZE AND SHIFT»

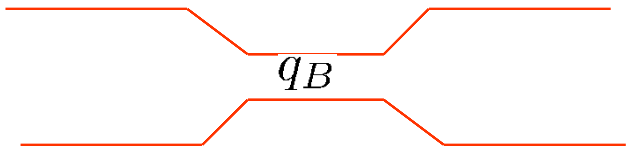


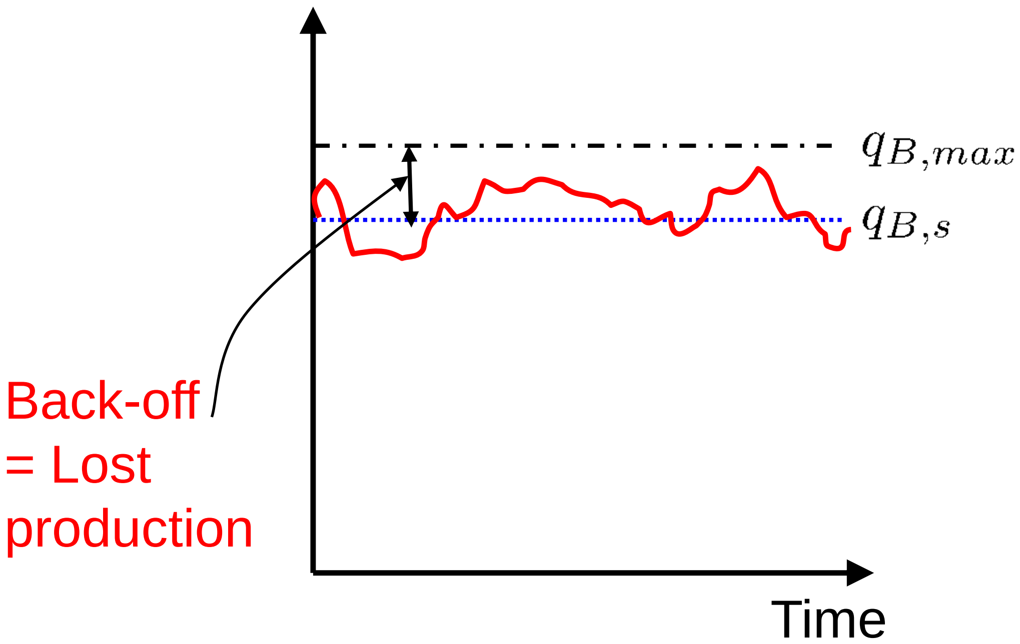
Rule for control of hard output constraints:

- “Squeeze and shift”!
- Reduce variance (“Squeeze”) and “shift” setpoint c_s to reduce backoff

CV = Active constraint

Example. Optimal operation = max. throughput. Want tight bottleneck control to reduce backoff!

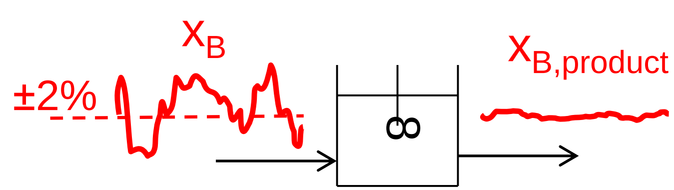




Example back-off.
x_B = purity product > 95% (min.)

- D_1 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_1 to <u>large</u> 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



Unconstrained optimum: Control “self-optimizing” variable.

- Which variable is best?
- Often not obvious (marathon runner)

Q1: I have to set back-off on output active constraints, right? Should I set it before evaluating losses? How much back-off should be set, 5%?
[Sigurd Skogestad] 5% is a good start, but it depends mostly on the measurement accuracy. Note that it must be a hard output constraint.
The value of the active constraint is usually considered a disturbance; this also takes care of the backoff variation

Q2: When I consider input active constraints as disturbance too. Ex. Temp feed to SOFC (T0f, T0a) are active at upper bound. Should I set +10% on this disturbance change or just -10%.
Is it possible to set +10% in this case?
[Sigurd Skogestad] Yes, the actual available input could be larger than what you think

Q3: In region I, I would like to try same structure as in region II which means Uf,MCFC set at 75% (upper bound) but it is an input active constraint
which considered as disturbance too. Hence Uf,MCFC doesn’t a disturbance in region I, am I right? Then region I have 4 disturbances but region II and III have 5 disturbances, am I right?

[Sigurd Skogestad] No, because also the value of the active constraint is usually considered as a disturbance

Q4: Do I need back-off on input active constraints too?

Answer: Input constraint: Normally not, except of it is a bottleneck constraint and the TPM is located somewhere else

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More on: WHAT ARE GOOD "SELF-OPTIMIZING" VARIABLES?

- *Intuition:* "Dominant variables" (Shinnar)

- More precisely
  1. Optimal value of CV is constant
  2. CV is "sensitive" to MV (large gain)

GOOD "SELF-OPTIMIZING" CV=c

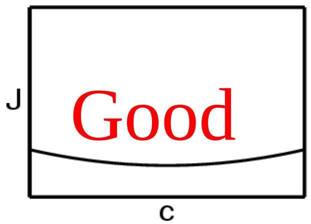
1. *Optimal value* c_{opt} is constant (*independent* of disturbance d):

   Want small optimal sensitivity, $F_c = \frac{\Delta c_{opt}}{\Delta d}$

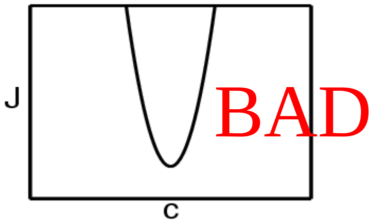
2. c is "sensitive" to MV=u (to reduce effect of measurement noise)

   Want large gain $G = \frac{\Delta c}{\Delta u}$

   Equivalently: Optimum should be flat



(b) Flat optimum: Implementation easy



(c) Sharp optimum: Sensitive to implementation erros

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>

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 = energy) - and we make the stupid choice os selecting CV = V = J
     - Then setting J < Jmin: Gives infeasible operation (cannot meet constraints)
     - and setting J > Jmin: Forces us to be nonoptimal (which may require strange operation)

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