Optimal Startup of Chemical Plants and its Practical Implementation

Marius Støre Govatsmark

Department of Chemical Engineering, NTNU

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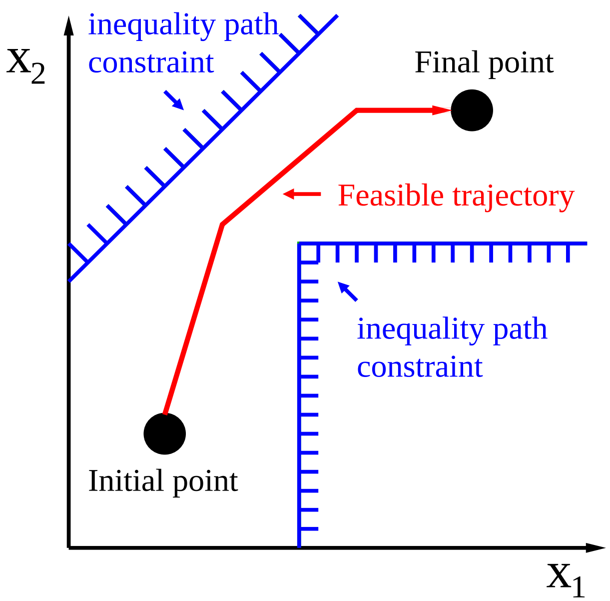
Outline

- Startup of chemical plants
- Industrial example: Startup
- Startup: Operating procedure
- Startup: PI-control with target setpoints
- Startup: Dynamic optimization (offline)
- Startup: Nonlinear model predictive control (online)
- Startup: Self-optimizing control
- Interesting research areas
- Concluding remarks

Startup of chemical plants

Def: Large change in operating point.



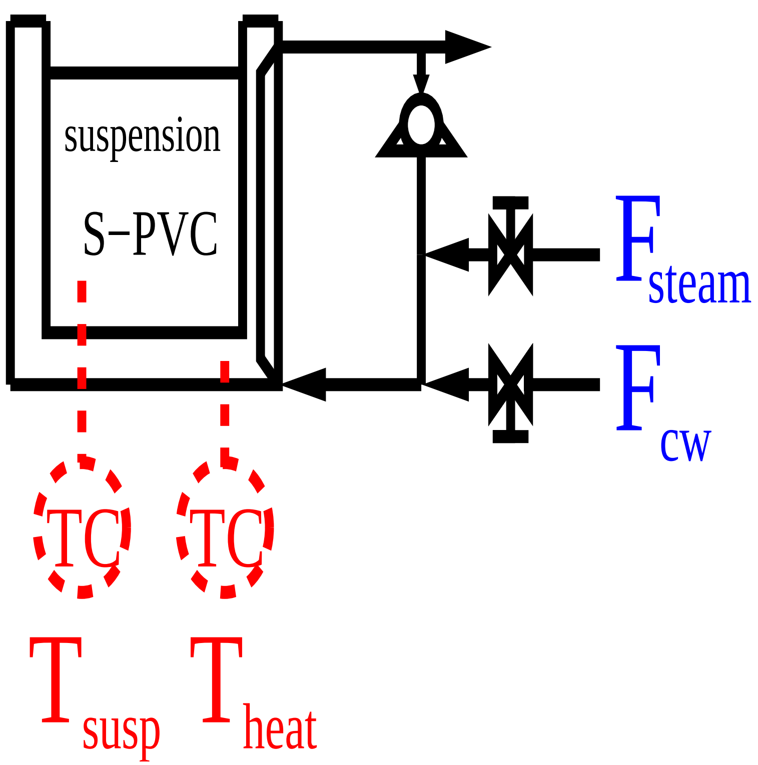
Characteristics:

- Strong nonlinearities
- Inequality path constraints
- Position constraints (e.g. final point)
- Varying physical/chemical phenomena

Objective:

- minimize transition time
- minimize energy / material usage
- minimize risk

Example: Startup
Norsk Hydro S-PVC plant, Porsgrunn, Norway



Manipulated variables:

$$u^T = [F_{cw} \quad F_{steam}]$$

Minimize startup time:

$$J = t_{final} - t_{init}$$

Constraints:

$$T_{heat}(t) \leq 90^o C$$
$$T_{susp}(t) \leq 52^o C$$
$$T_{susp}(t_{final}) = 50^o C$$

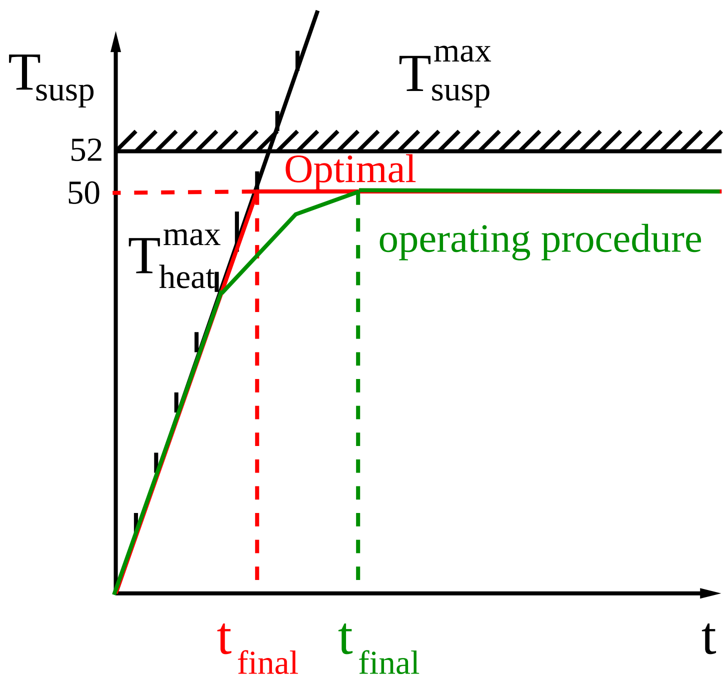
Disturbance:

$$d = T_{susp}(t_{init}) = [25 \pm 5^o C]$$

Measurements:

$$y^T = [T_{susp} \quad T_{heat}]$$

Startup: Operating procedure



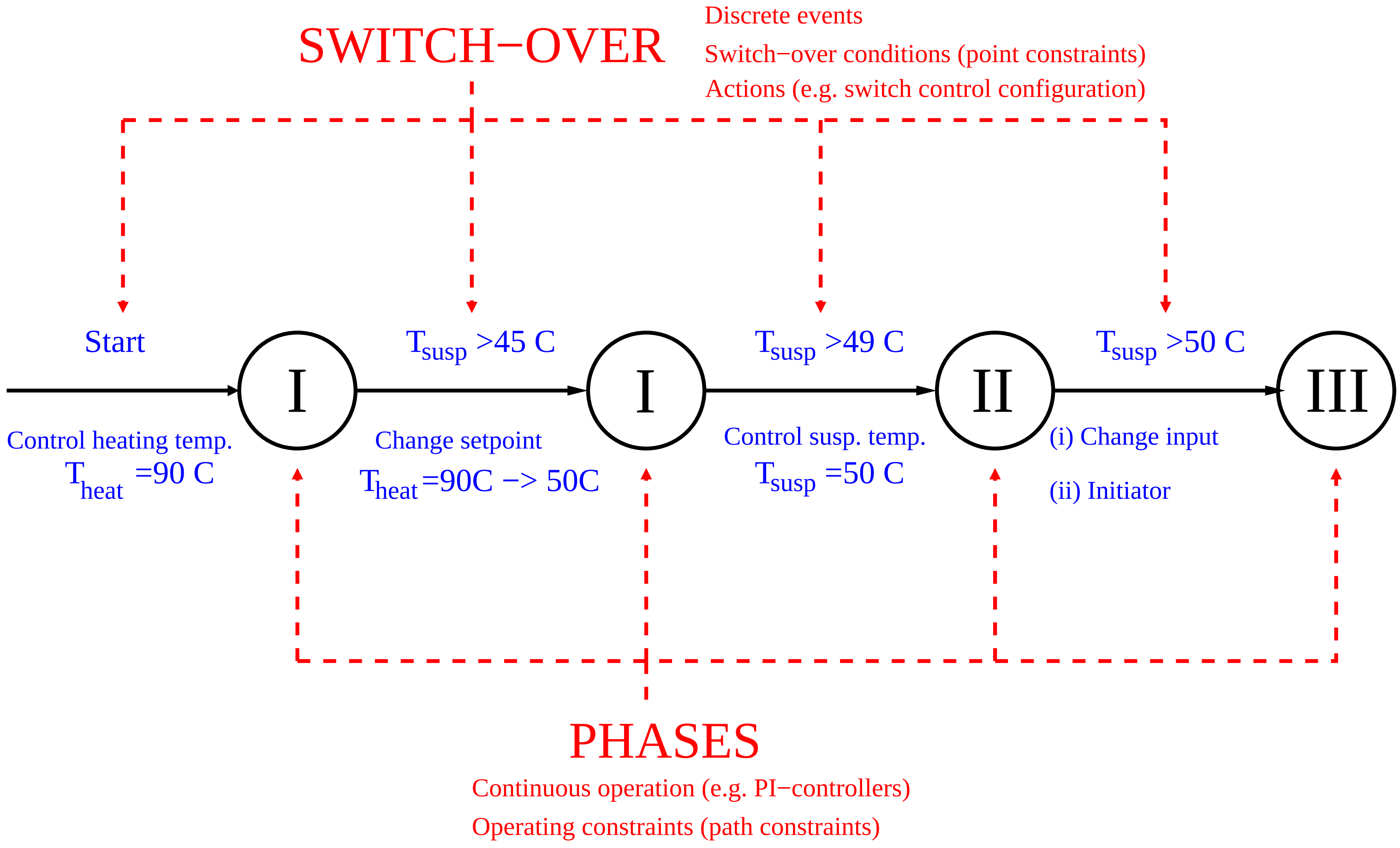
Design procedures:

- Trial and error
- Stephanopoulos and coworkers

Handling procedures:

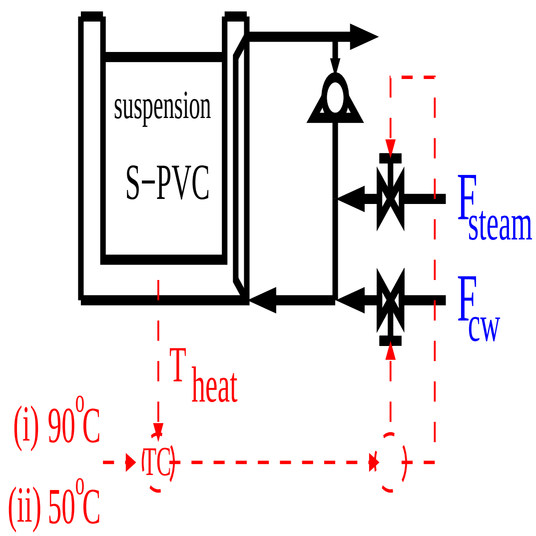
- Operators
- Automation
  - Fuzzy logic control (Bahar *et al.* 1994)
  - Discrete event system (Klein *et al.* 2000)

Startup: Operating procedure

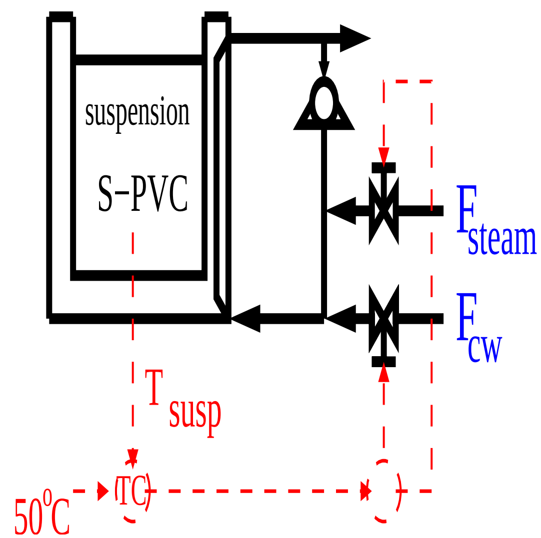


Varying control structures during startup

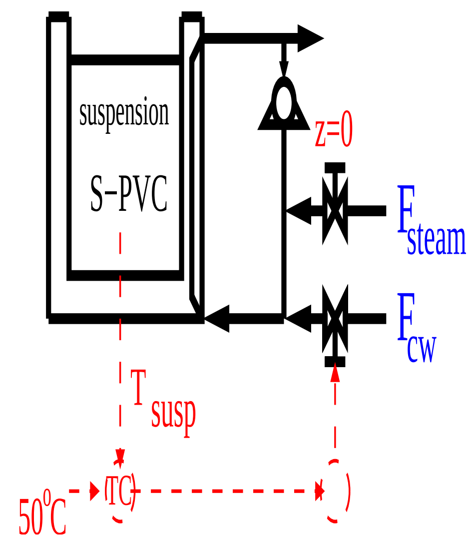
Structure I



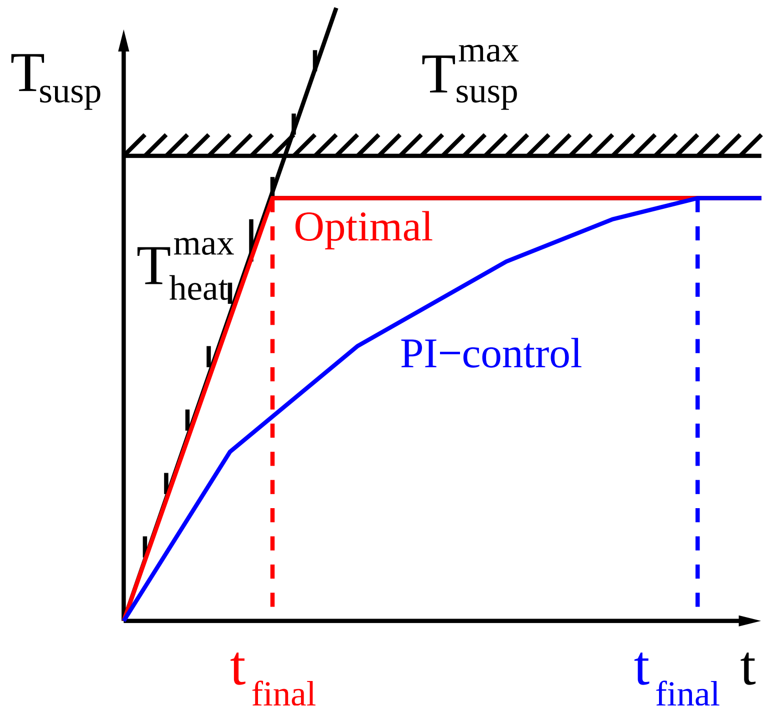
Structure II



Structure III



Startup: PI-control with target setpoints



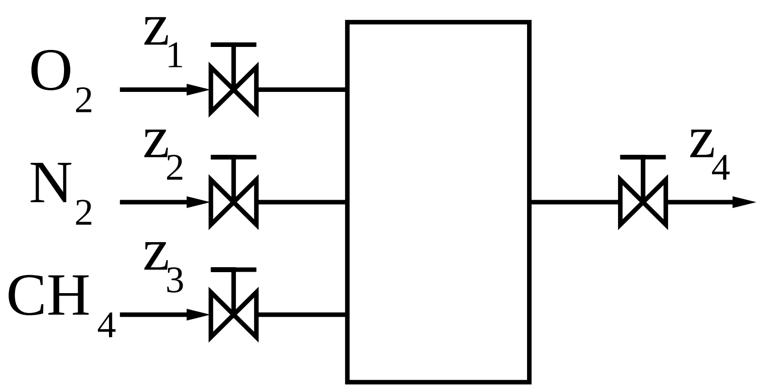
Slow control:

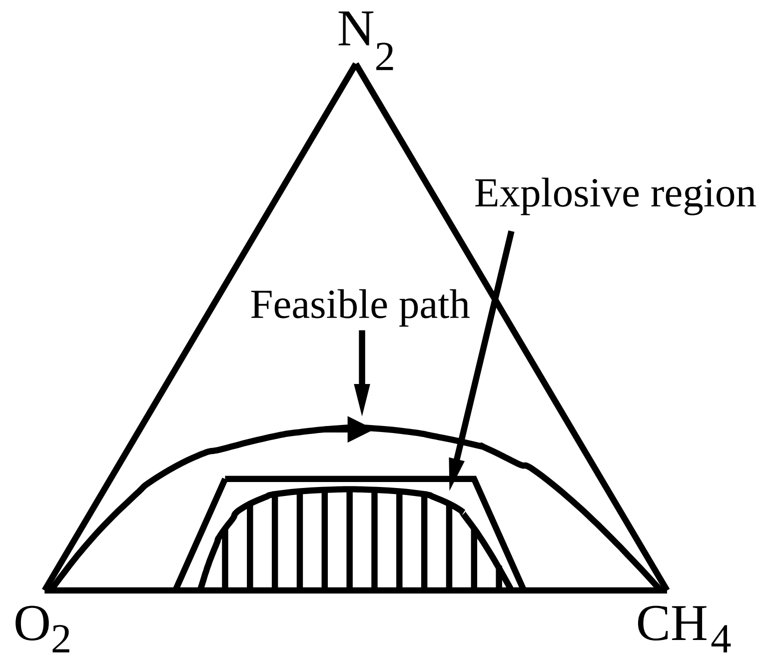
- Path constraints
- Nonlinearities

Phillips *et al.* (1988*a*), Wozny and Li (2003): "Conventional approach"

Startup: PI-control with target setpoints

Tank changeover problem (Rivas and Rudd 1974): Infeasible





Dynamic optimization

$$\min_{u(t),\dot{x}(t),x(t),t_f} \phi(x(t_f), u(t_f), d(t_f), t_f) + \int_0^{t_f} L(x(t), u(t), d(t), t)dt$$

$$\begin{array}{lll} plant: & \mathsf{f}(\dot{x}(t), x(t), u(t), d(t), t) = 0 & \forall t \in [0, t_f] \\ path\ constraints: & \mathsf{g}(\dot{x}(t), x(t), u(t), d(t), t) \leq 0 & \forall t \in [0, t_f] \\ position\ constraints: & \mathsf{k}_p(\dot{x}(t_p), x(t_p), u(t_p), d(t_p), t_p) \leq 0 & \forall p \in \{1, .., n_p\} \\ disturbances: & \mathsf{d(t)=d_0}(t) & \forall t \in [0, t_f] \end{array}$$

Gives:

Optimal profile u(t).

Cervantes *et al.* (2002): polyethylene plant

Modeling

Discrete events

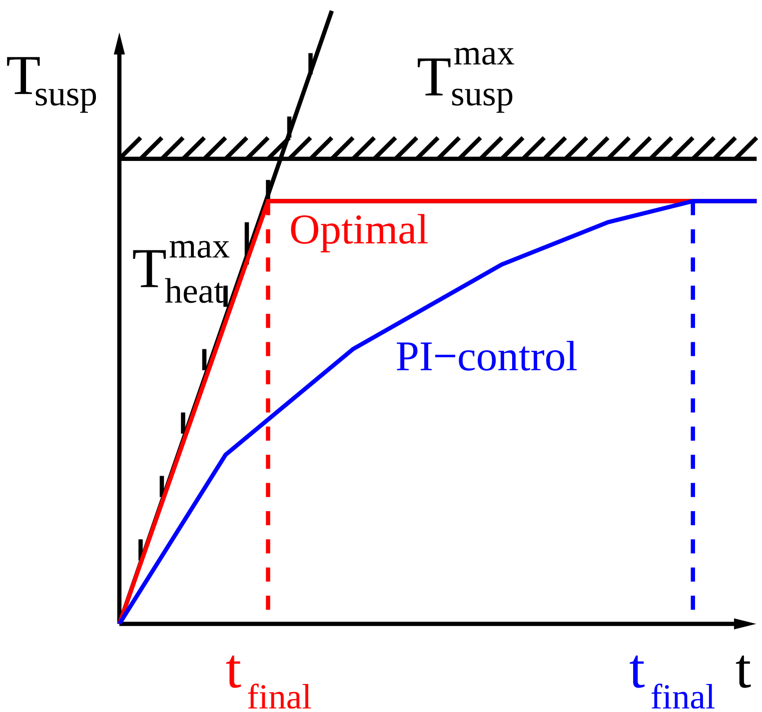
Jumps

Ref: Barton *et al.* (1998)

Dynamic optimization methods

Ref: Biegler *et al.* (2002), Barton *et al.* (1998).

Startup: Dynamic optimization (offline)



Sensitive to uncertainty

- Disturbances:
  - Initial states
  - Final states
  - Parameters
- Implementation error
  - Measurement errors
- Model error

Reduce sensitivity by feedback:

1. regularly update model and re-optimize
2. change implemented variables (controlled variables)

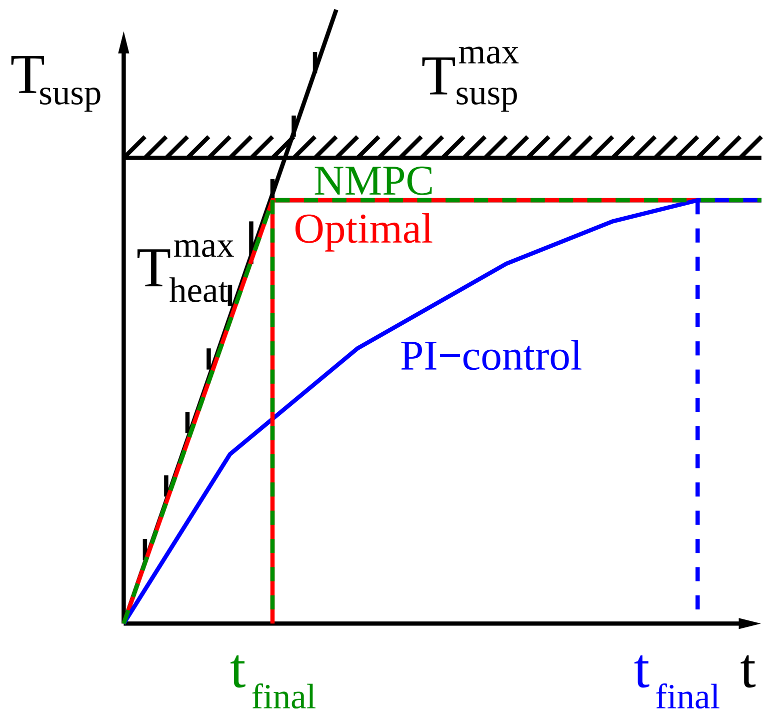
Nonlinear model predictive control (NMPC)

$$\min_{u_k, x_k, y_k} \sum_{k=1}^{p} (y_k - y_{s,k})^T Q_{y,k} (y_k - y_{s,k}) + \sum_{k=1}^{m} (u_k - u_{r,k})^T Q_{u,k} (u_k - u_{r,k})$$

$$\begin{array}{lll}
plant: & \mathsf{f}_k(x_{k+1}, x_k, u_k, d_k) = 0 & \forall k \in \{1, .., p\} \\
outputs: & \mathsf{y}_k = g_k(x_k, u_k, d_k) & \forall k \in 1, .., m\} \\
input\ constraints: & \mathsf{u}_{max} \le u_k \le u_{min} & \forall k \in \{1, .., m\} \\
state\ constraints: & \mathsf{x}_{max} \le x_k \le x_{min} & \forall k \in \{1, .., s\} \\
output\ constraints: & \mathsf{y}_{max} \le y_k \le y_{min} & \forall k \in \{1, .., p\}
\end{array}$$

Ref: Biegler (1998), Santos *et al.* (2001)

Startup: Nonlinear model predictive control (NMPC)



Advantages:

- Feasibility
- Nonlinearity

Disadvantages:

- Modeling
- Computation online
- Objective function?

Self-optimizing control for dynamic system (ideal)

$$\min_{\color{red}{c(..),c_s(t)},u(t),\dot{x}(t),x(t),t_f} \phi(x(t_f),u(t_f),d(t_f),t_f) + \int_0^{t_f} L(x(t),u(t),d(t),t)dt$$

$$\begin{array}{lll}
plant: & \mathrm{f}(\dot{x}(t),x(t),u(t),d(t),t) = 0 & \forall t \in [0,t_f] \\
path\ constraints & \mathrm{g}(\dot{x}(t),x(t),u(t),d(t),t) \leq 0 & \forall t \in [0,t_f] \\
\color{red}{controlled\ variables} & \mathrm{c}(\dot{x}(t),x(t),u(t),d(t),t) = c_s(t) + d_c(t) & \forall t \in [0,t_f] \\
disturbances & \mathrm{d(t)}=\mathrm{d}_0(t) + \Delta d(t) & \forall t \in [0,t_f] \\
impl.\ errors & \mathrm{d}_c(t) = d_{c,0}(t) + \Delta d_c(t) & \forall t \in [0,t_f]
\end{array}$$

Find: Controlled variables $c(\dot{x}(t),x(t),u(t),d(t),t)$ and corresponding setpoint $c_s(t)$

Implement:

Operating policy: Less sensitive to uncertainty (i.e. more robust operation)

"Self-optimizing control" for dynamic system

Main issue: Identify:

- Controlled variables with setpoints
- Switching conditions.

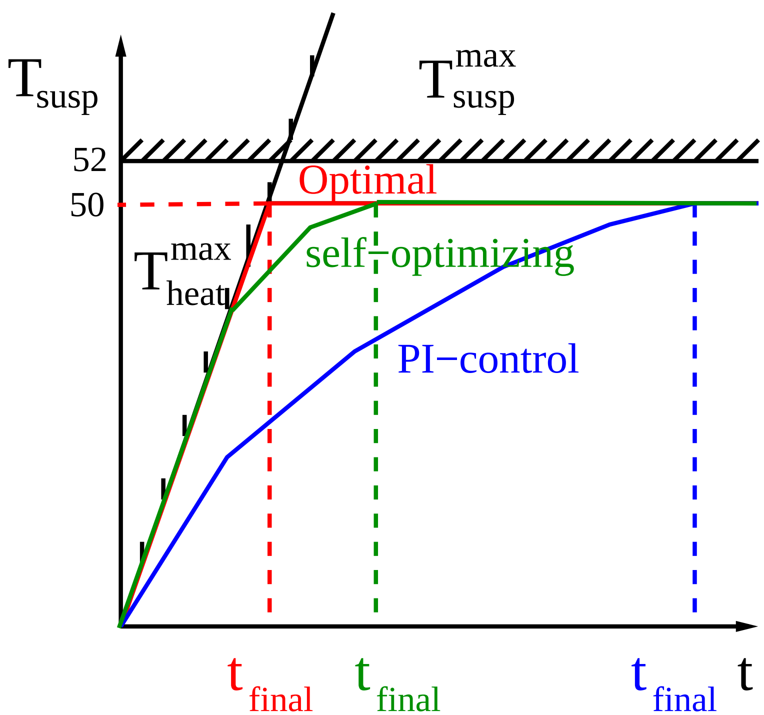
"Method":

- Dynamic optimization - nominal and expected uncertainty
- Initial controlled variables and setpoints:
  - Active constraints
  - Nominal setpoints (afo time)
- Show loss as function of time
  - Regions with large loss
- Simplify operating policy
  - Profile $\rightarrow$ constant setpoints

Ref: Bonvin and coworkers

Startup: Self-optimizing control

In practice: A robust operating policy (procedure)



Advantages:

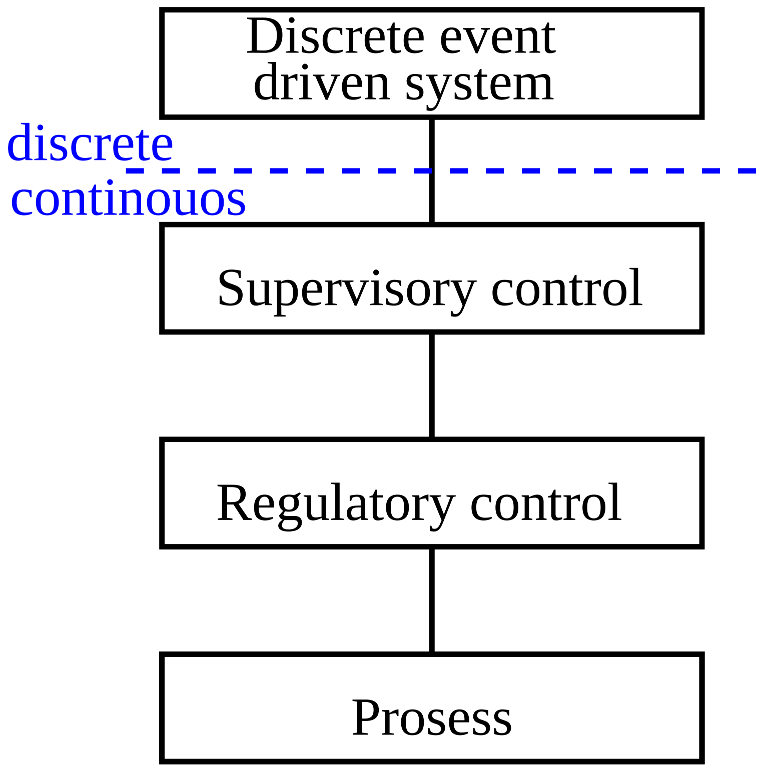
- Simple and close to optimal operation
- Less sensitive to uncertainties
- Reduced online computation

Disadvantages:

- Identify c
- Modeling and model uncertainty
- Offline computations

Handling extreme uncertainties $\rightarrow$ safety system.

Interesting research areas



- Operating procedure
- Dominating phases
  (e.g. semi-continuous phase in distillation column).
- Dominating subprocess
  (e.g. reactor or columns.)
- Discrete event driven system

Concluding remarks

- Open
- Complex
- Challenging
- Safety
- Piece-wise self-optimizing dynamic control

Coments

Consider "bottlenecks" with respect to startup time or cost:

- Parts in a startup procedure, e.g. dominating phases (e.g. semi-continous phase in distillation column startup).

- Part of the process, e.g. reactor or distillation column.

Discrete event driven system

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