# Examples of control structures used in chemical industry

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# Agenda

- Short presentation of the Perstorp group
- General about control applications in the specialty chemicals industry
- Examples of control structure improvements

# The Perstorp group – short facts

- Specialty chemicals company with focus on organic chemistry
  - Turn-over 2011: 11.3 GSEK
  - Owned by PAI Partners – a French equity company.
  - 1500 employees
- Products: Mainly raw materials for other chemical industries
  - Additives in paints and coatings, plastic-processing and automotive industries
  - Thermoplastics, plasticizers, food and feed, solvents, bleaching agents, etc

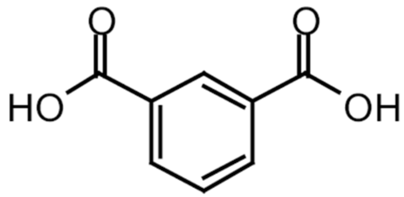
  but also end products like feed additives and bio-diesel.

# Most important product groups

- **Polyols:** pentaerythritol, neopentylglycole, trimethylolpropane, di-penta, di-TMP
- **Esters:** caprolactone, di-propylheptylphthalate, RME (”bio diesel”)
- **Acids:** octanoic acid, isophthalic acid, formic acid, propanoic acid, DMBA
- **Special polymers:** polycaprolactonepolyols, thermoplastics
- **Formates:** sodium-, potassium- and calcium formate, propionates
- **Aldehydes:** formaldehyde, iso- and normal butyric aldehyde
- **Alcohols:** butanol, octanol

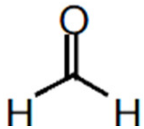
More than 100 different chemicals,
several thousand products

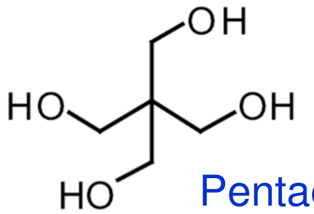
# Some molecules



Isophthalic acid

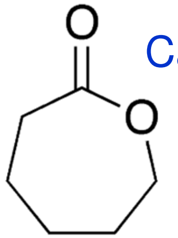
Formaldehyde



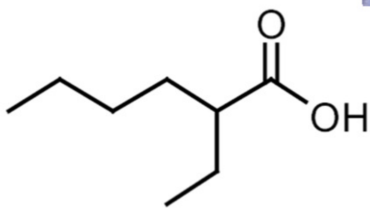


Pentaerythritol


n-butanol

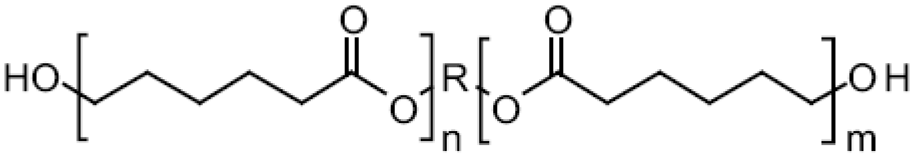


Caprolactone

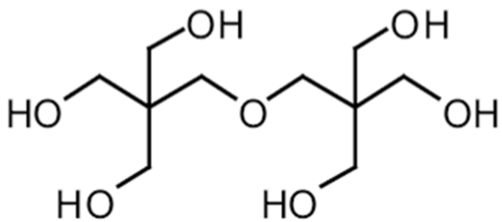


2-ethyl-hexanoic acid


Neopentylglycol



Poly-caprolactone



Di-penta

# Eleven production sites


In total: ~40 plants ~50 000 variables ~8 000 controllers

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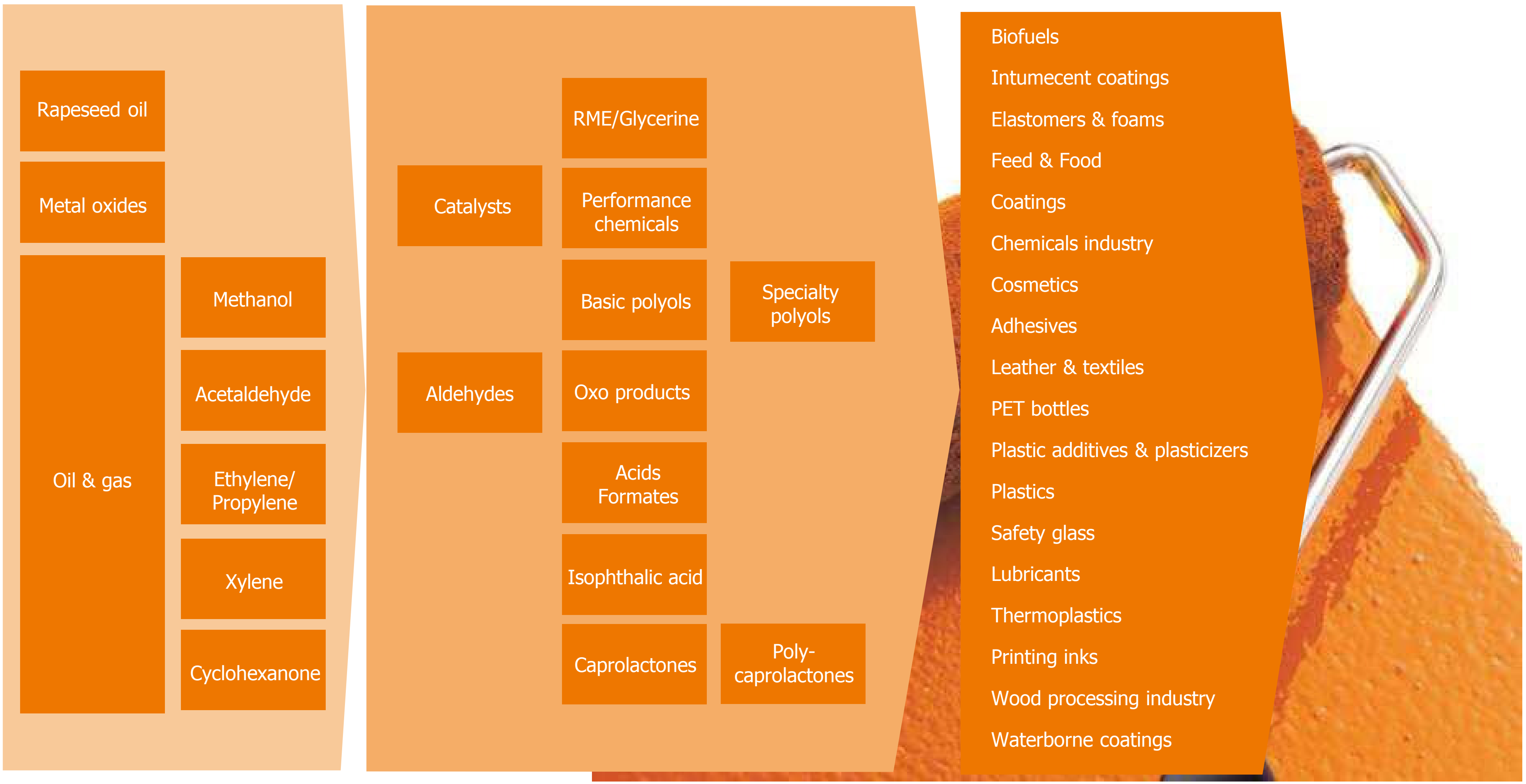
# Perstorp in the value chain

| Raw materials | Products | Application areas |
|---|---|---|
| 75% based on oil or natural gas. | In some cases used as raw materials internally or sold to other companies which produce intermediates for industry or products to end-users. | Such as car coatings, household paint, animal feed and plastics. |



# Many different types of processes to control

- Reactors
  - Tube reactors, tank reactors
  - Continuous, batch, semi-batch
- Heat exchangers
  - Liquid – liquid, steam - liquid
- Distillation columns
  - Continuous, batch
- Evaporators
- Stripper columns
- Crystallisers
- Centrifuges
- Filters
- Thickeners
- Dryers
- Scrubbers
- Boilers

In order to optimize controls you need to have good process knowledge!

# Control: What differs between different industries?

- For mature processes, with a large installed base, there are standard solutions; often quite advanced.
- Examples:
  - Oil refineries (~800 world wide)
  - Paper machines (2000)
- For specialized processes, that are fairly unique, local expertise has to design the controls.
- Examples:
  - Caprolactone (<5 plants world-wide)
  - Butyric aldehyde (<25)
  - Pentaerythritol (<50)

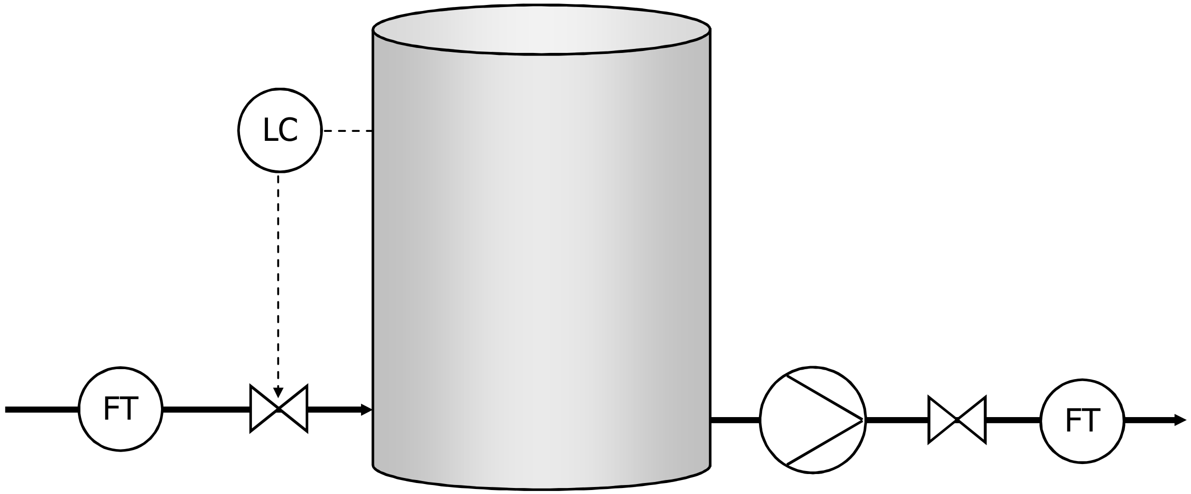
# Typical tasks for the control group

- Improve productivity by decreased variation and increased automation.
  - Smarter control structures, e.g. feedforwards, mid-ranging, cascades, maximizing control, ratio-in-cascade
  - PID control parameter tuning
  - Introduce new controllers
  - Support in commissioning of new plants
  - Automatic control of buffer levels
  - Alarm management

- Process historian (database) ownership; applications and development

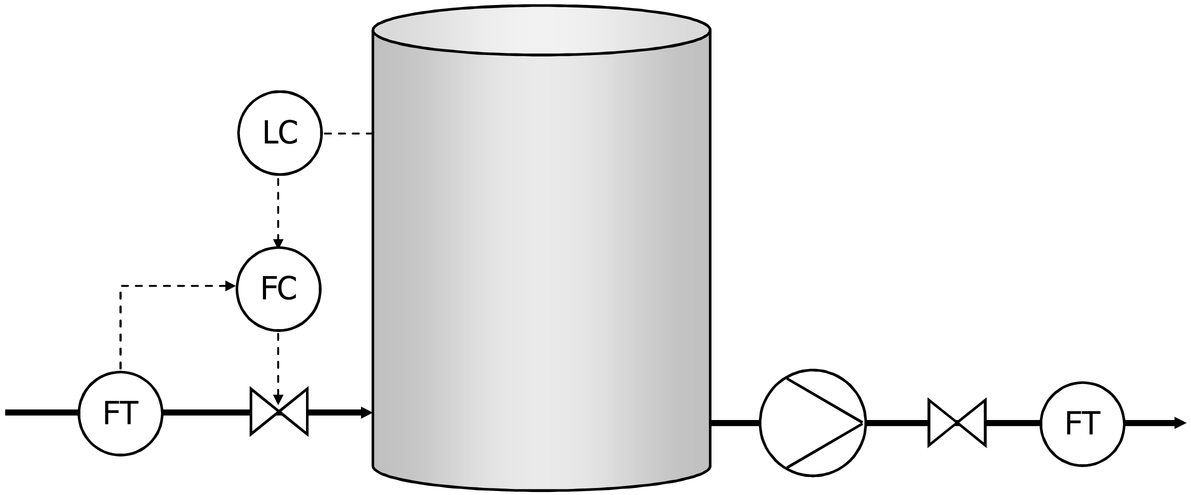
- Training seminars

# Ex: Level control with improvement opportunity




- The level in the tank varies too much, because there are pressure variations in the line for the incoming flow.
- We can't tune the controller more aggressively - then it becomes unstable.
- Can we still improve control performance?

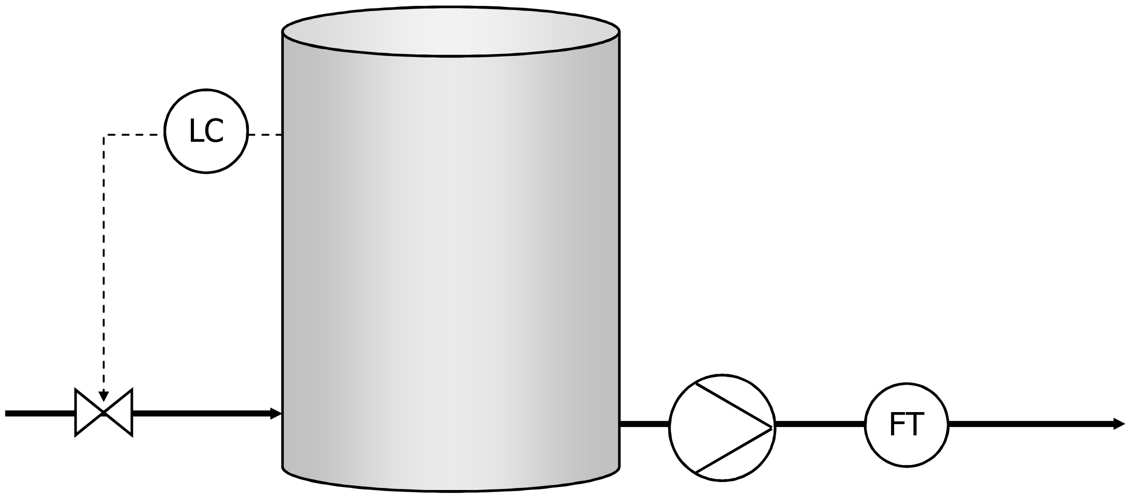
# Solution: Control the flow too!




= Cascade control

# Ex: Level control with improvement opportunity

- The level in the tank varies too much because of variations in output flow.
- We can't tune the controller more aggressively - then it becomes unstable.
- Can we still improve control performance?



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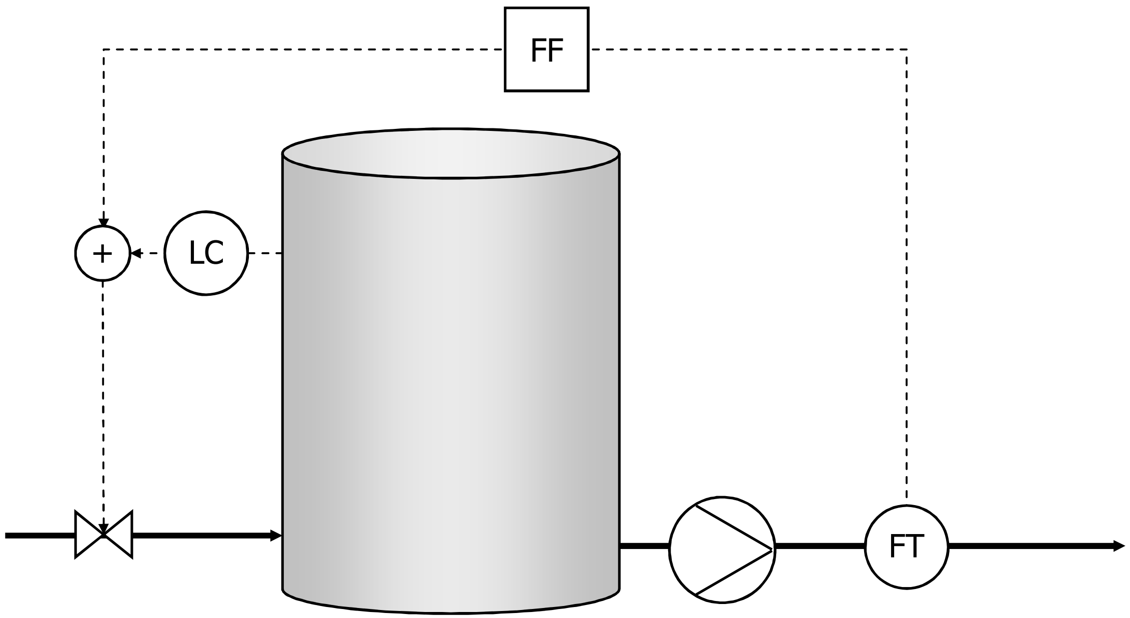
# **Feedforward**: Give early information to the controller




$$u(t) = \underbrace{K_c\left(e(t)+\frac{1}{T_i}\int_0^t e\right)}_{\text{Feedback term}} + u_{FF}$$

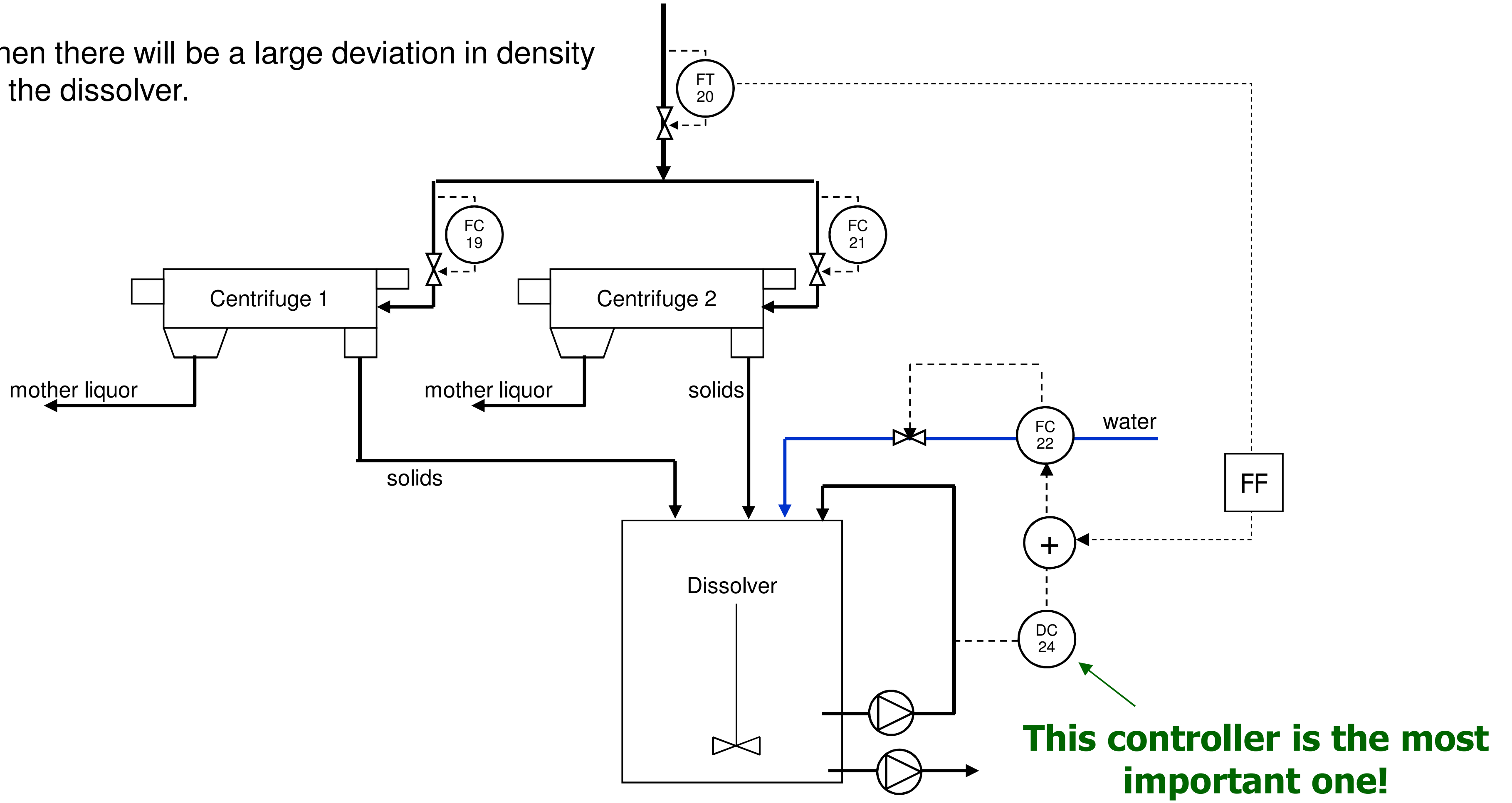
Feedback term

Feedforward term

# Density control in a dissolver

Issue: Sometimes one of the flows FC-19 or FC-21 is closed for cleaning the centrifuge.

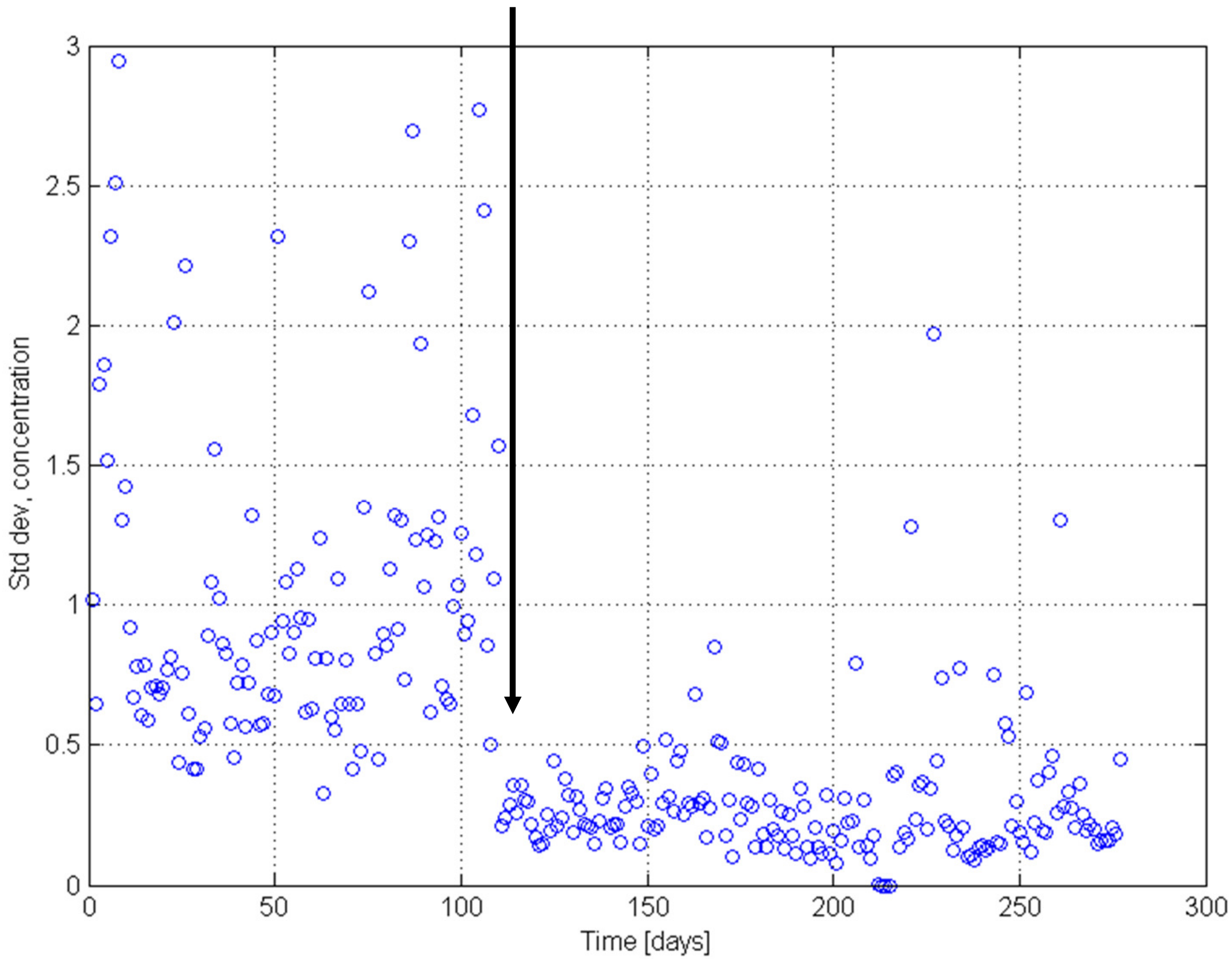
Then there will be a large deviation in density in the dissolver.




**This controller is the most important one!**

# Feedforward and PI tuning reduces variations

Feedforward, and some other improvements, introduced



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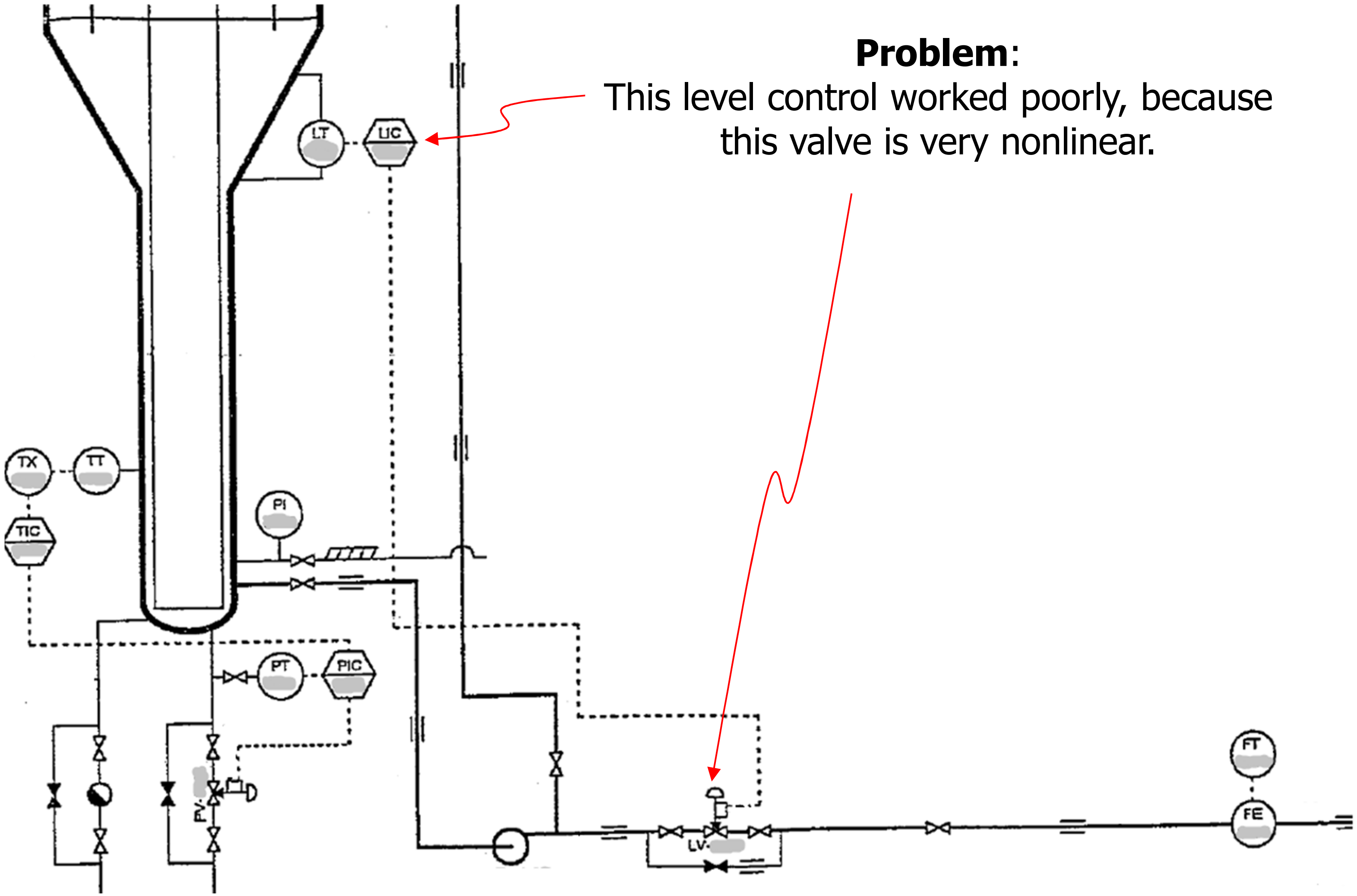
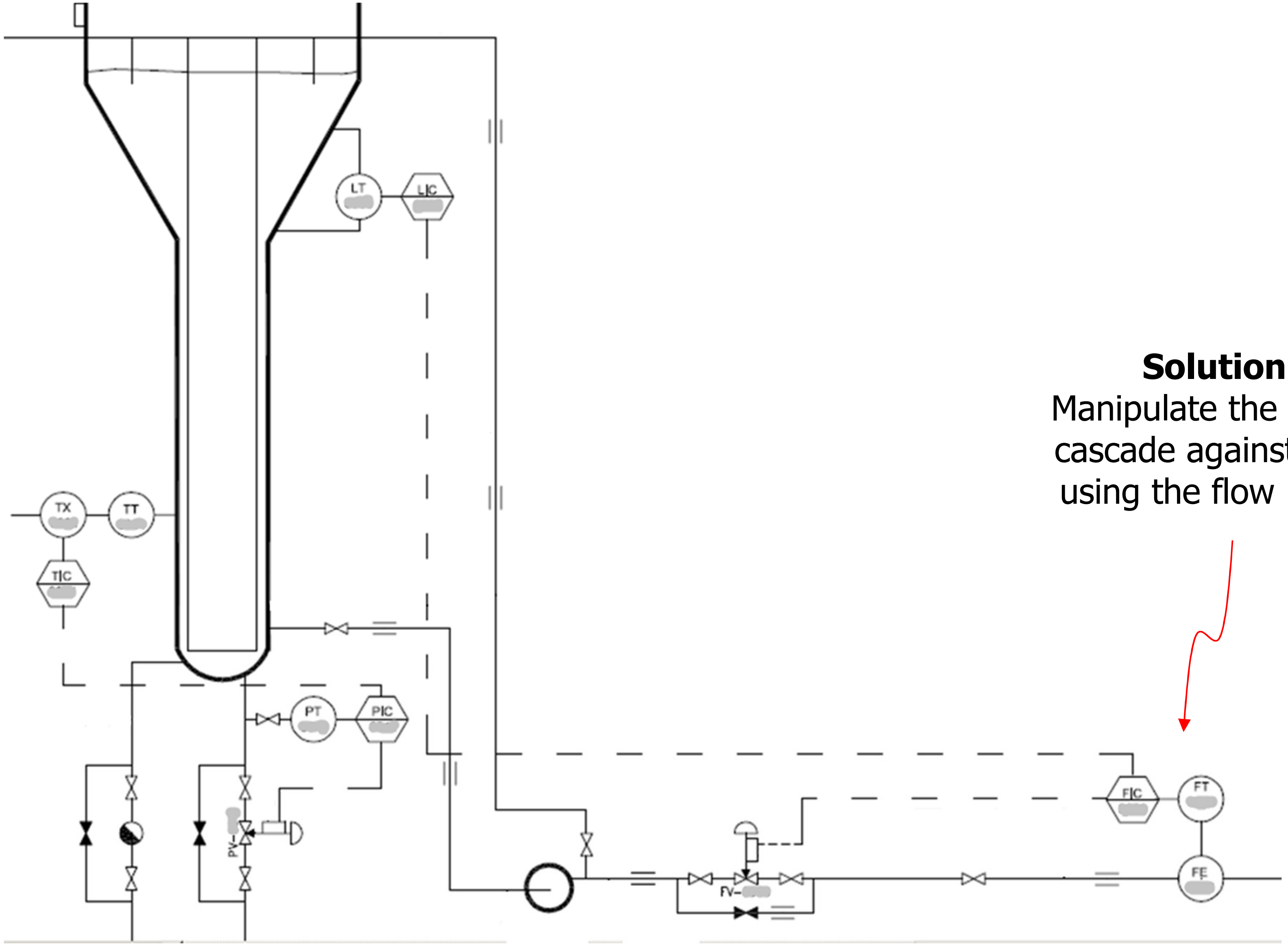
# Evaporator with poor level control

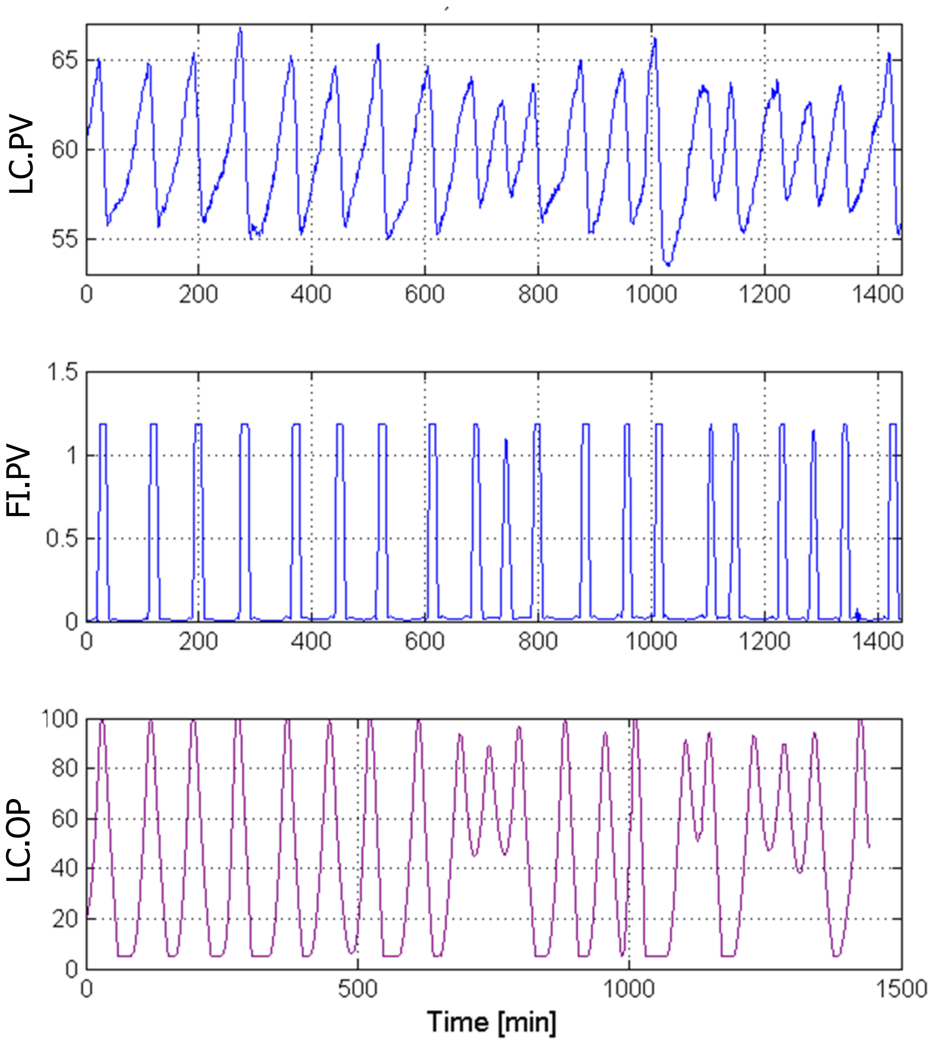


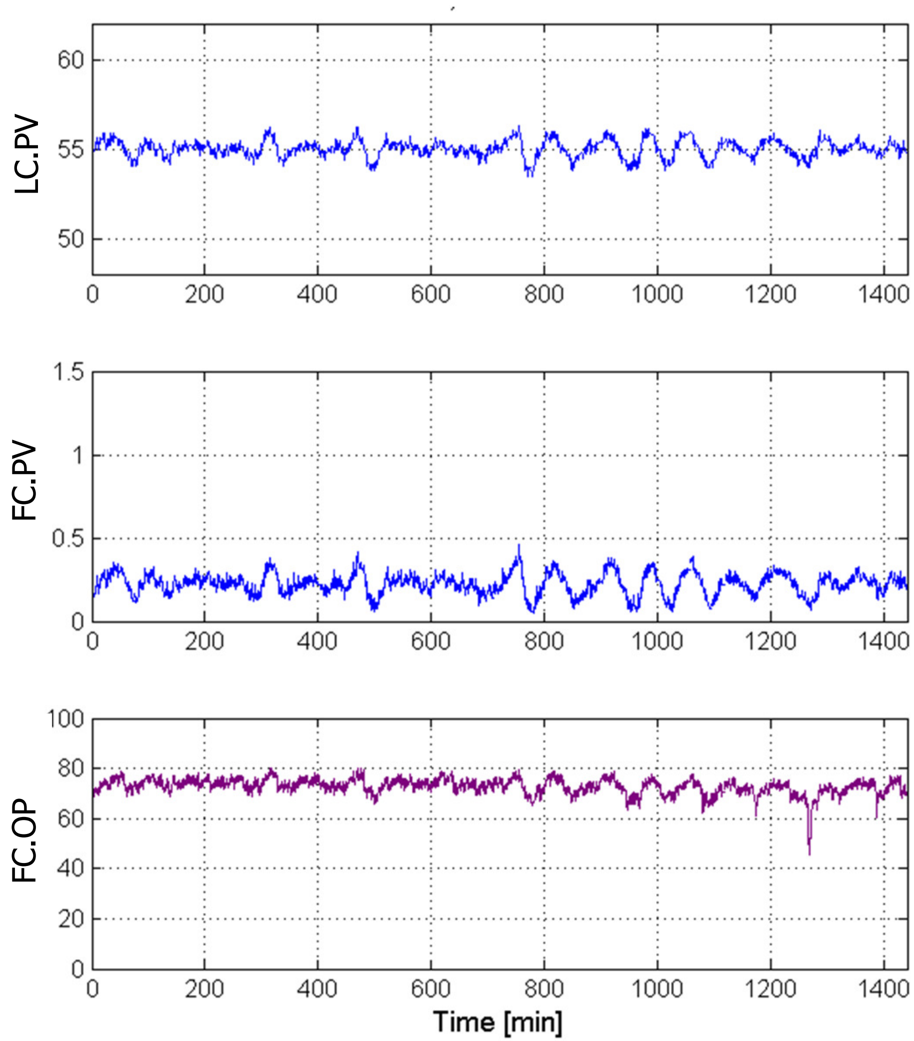

**Problem**:
This level control worked poorly, because this valve is very nonlinear.




**Solution**:
Manipulate the flow in cascade against level, using the flow meter.

# More stable level and smoother flow using cascade control

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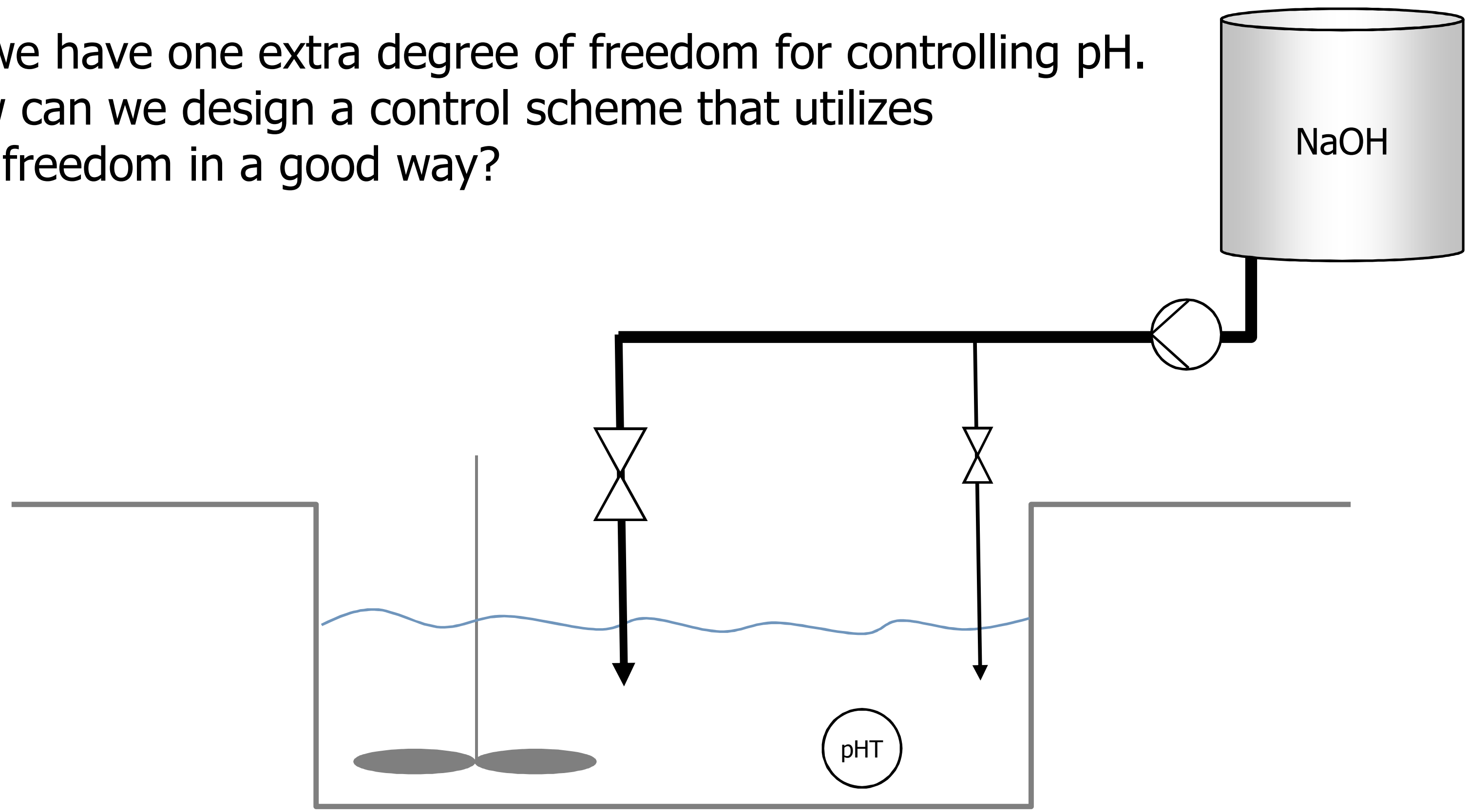
# Sewer pH-control process

There are two valves for feeding caustic to the pit:
a small, accurate one, and a larger coarse valve.

So we have one extra degree of freedom for controlling pH.
How can we design a control scheme that utilizes
this freedom in a good way?



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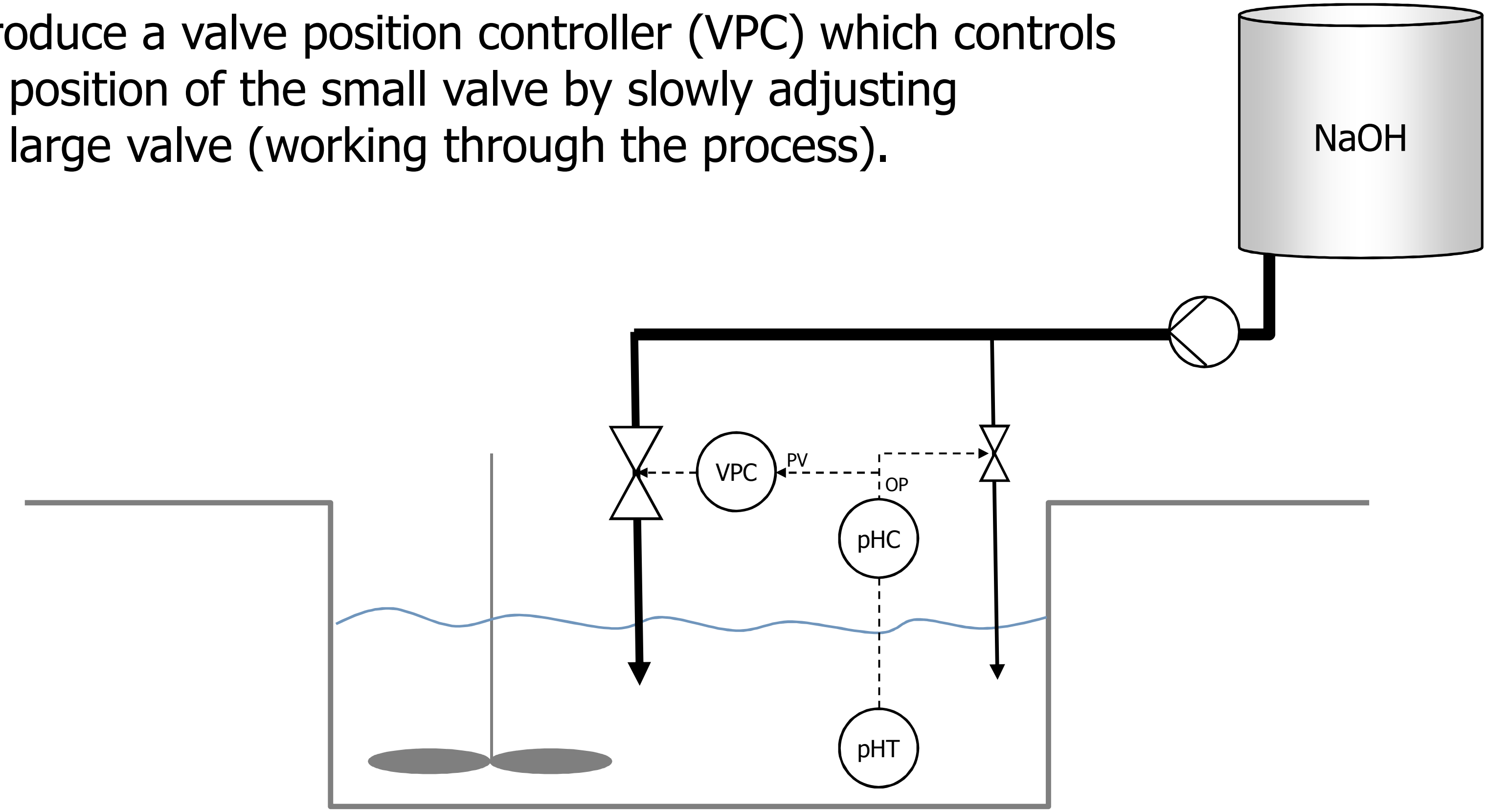
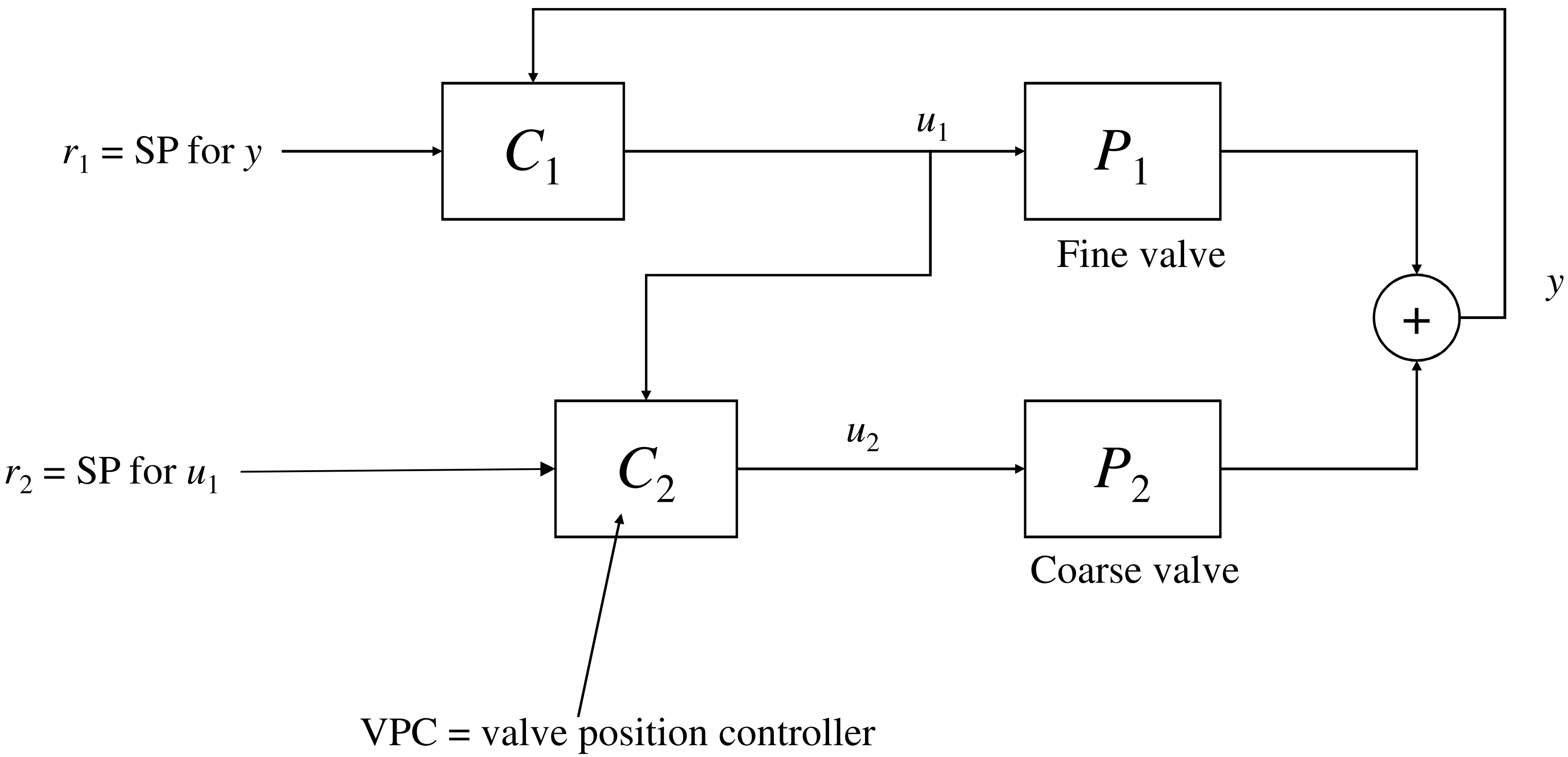
# Solution: Mid-ranging (valve position control)

Let a pH-controller manipulate the small valve

Introduce a valve position controller (VPC) which controls the position of the small valve by slowly adjusting the large valve (working through the process).

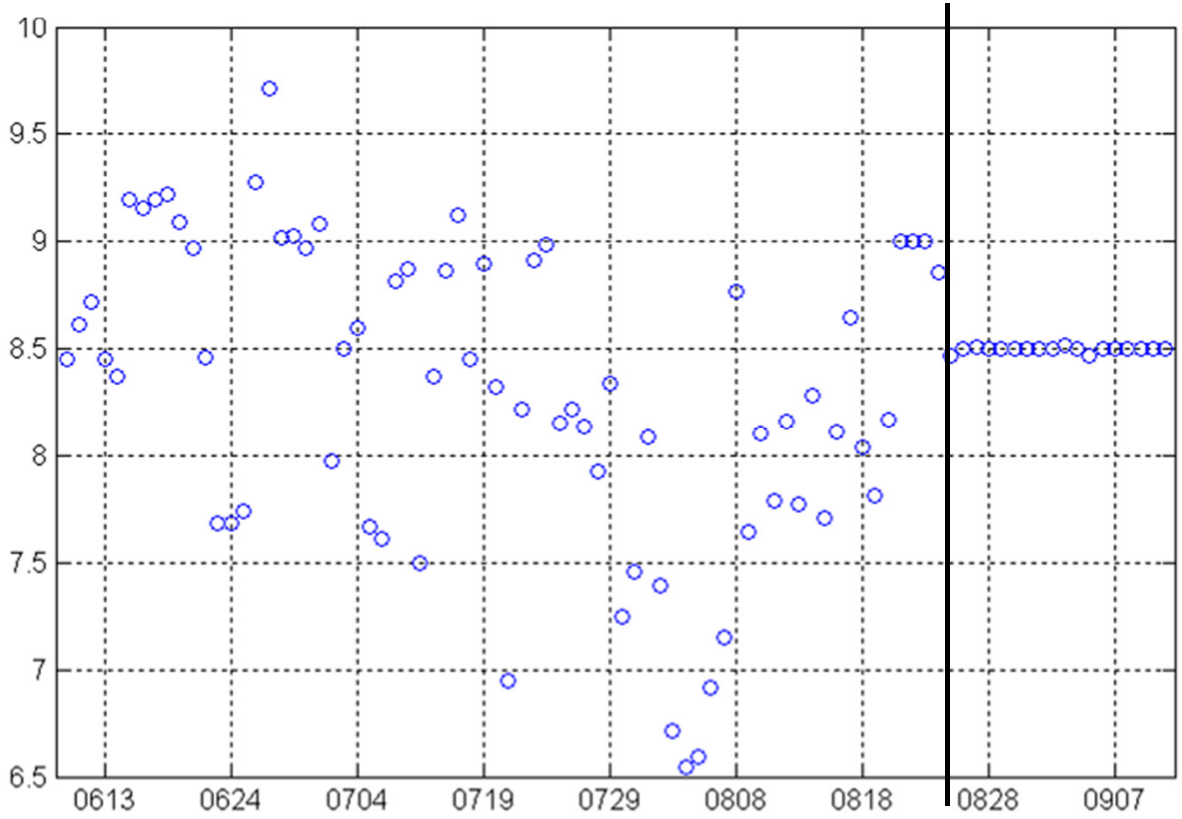


# Block schedule for MR-control: Give setpoint for $u_1$



# pH control; Results

pH: Daily averages before and after new control structure

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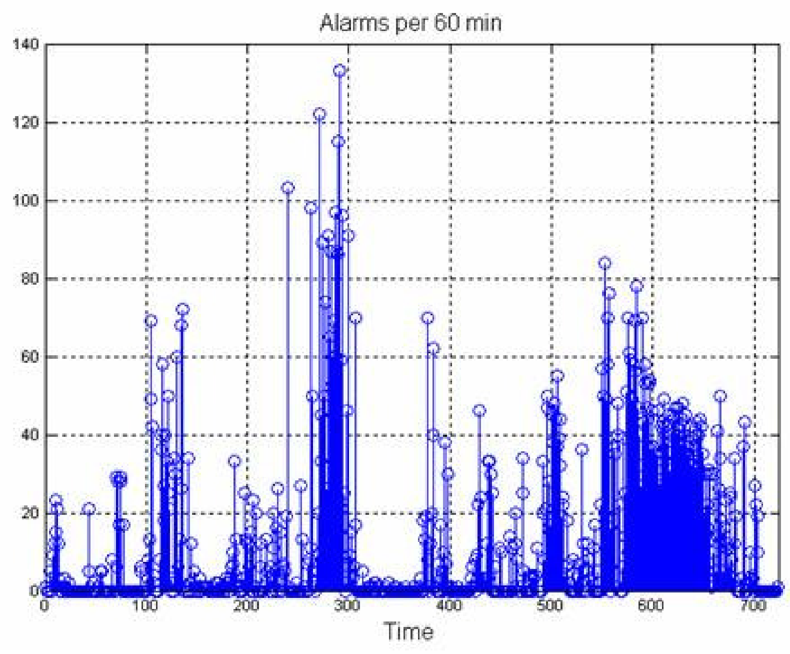
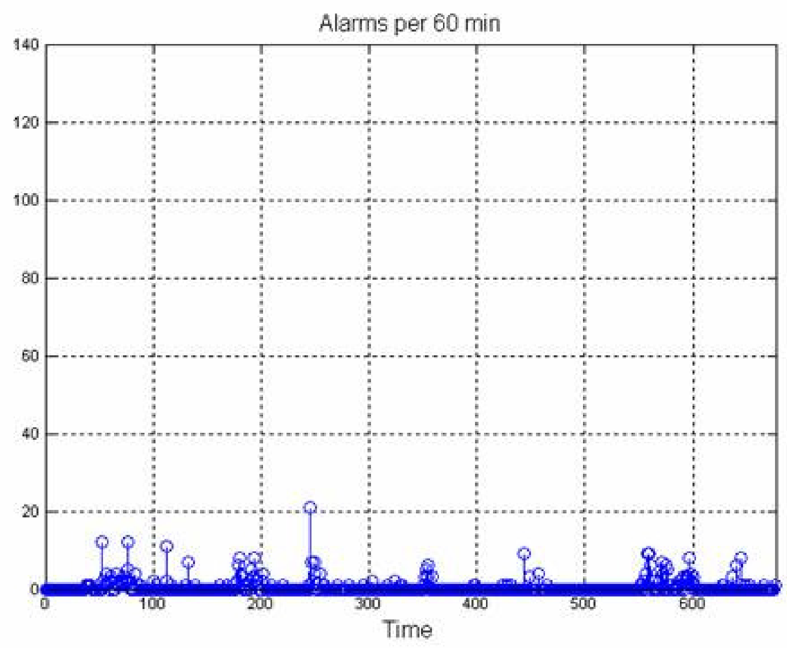
# Improved pH-control gives fewer alarms




Before: 10 488 alarms in one month

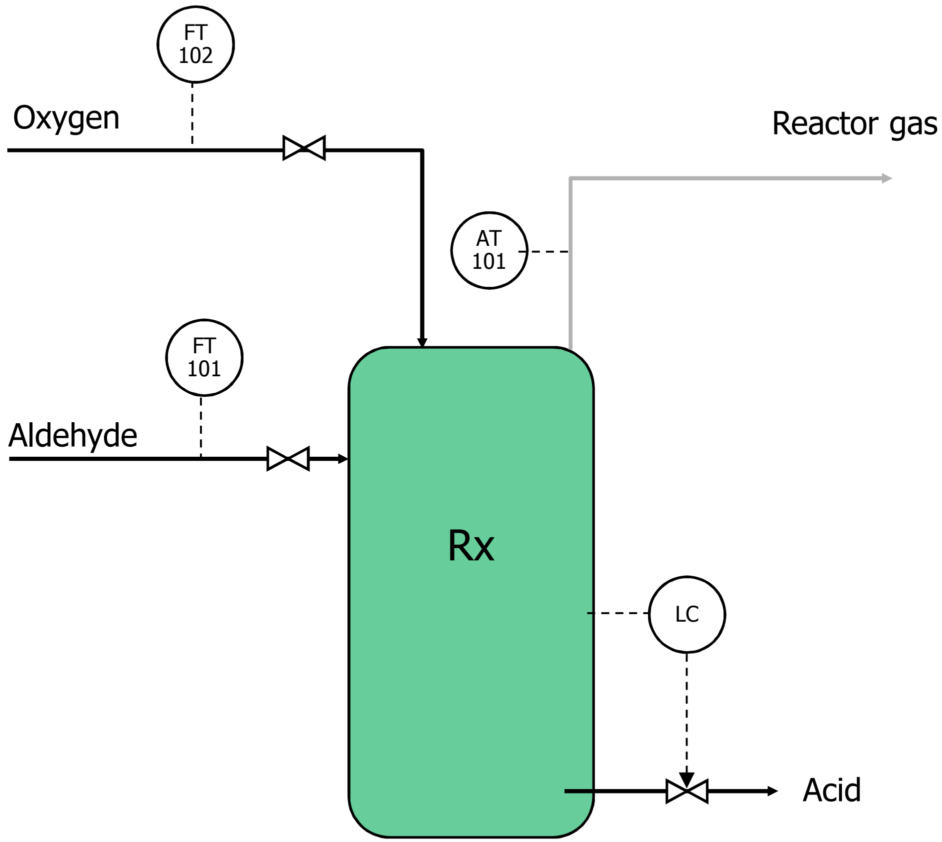
After: 418 alarms in one month

96% fewer alarms from this object.

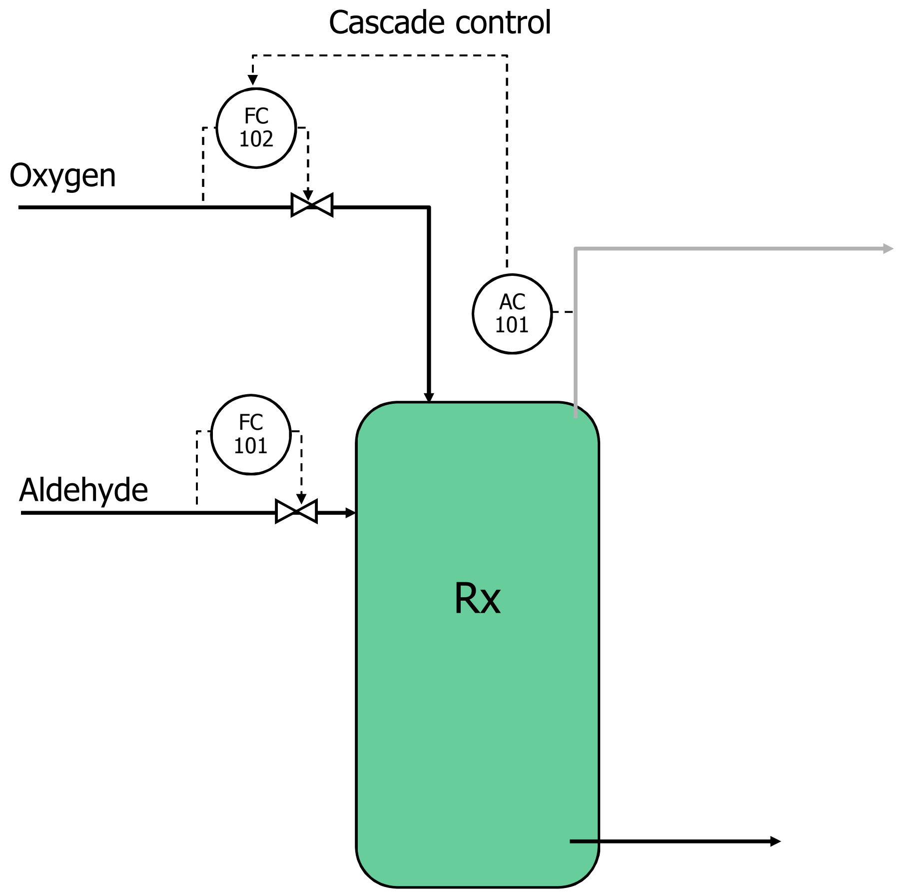
# Reactor control; Residual oxygen

- Process: Two phase oxidation reactor (aldehyde-oxygen)
  - On-line measurement of residual $O_2$ in reactor gas; must be controlled.
  - Which different control structures are feasible, and what are their respective advantages?

Aldehyde feed is master = TPM
(determines production rate)

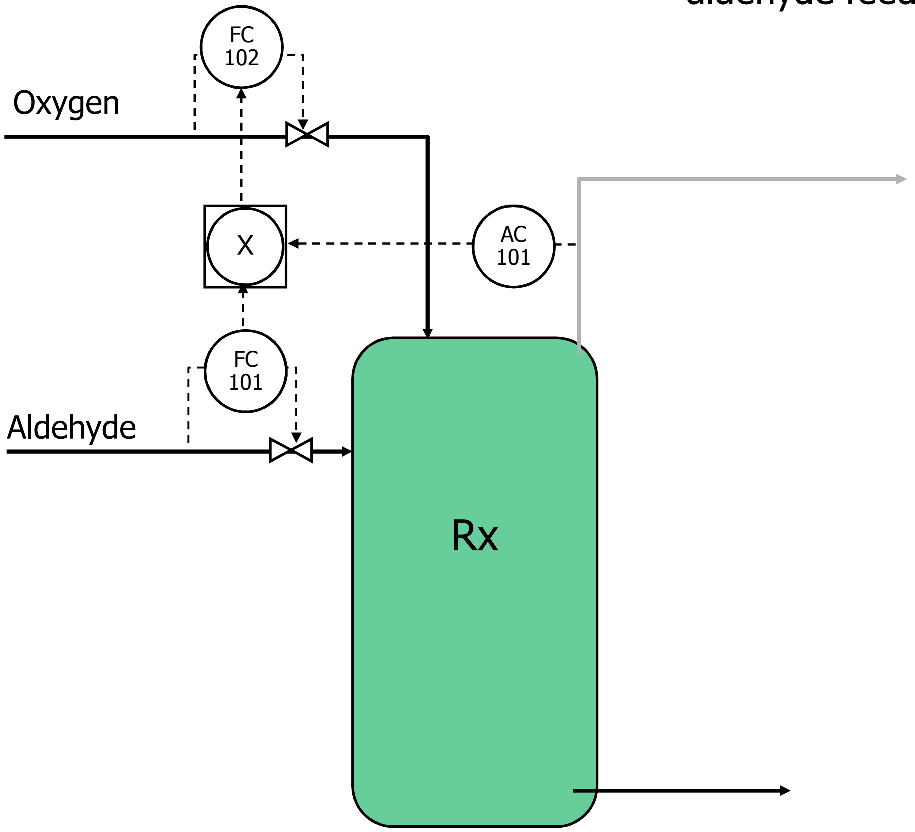


# Alternative 1: $O_2$-feed in cascade against residual-$O_2$



# Alt 2: $O_2$ feed in ratio against aldehyde, cascade against residual $O_2$

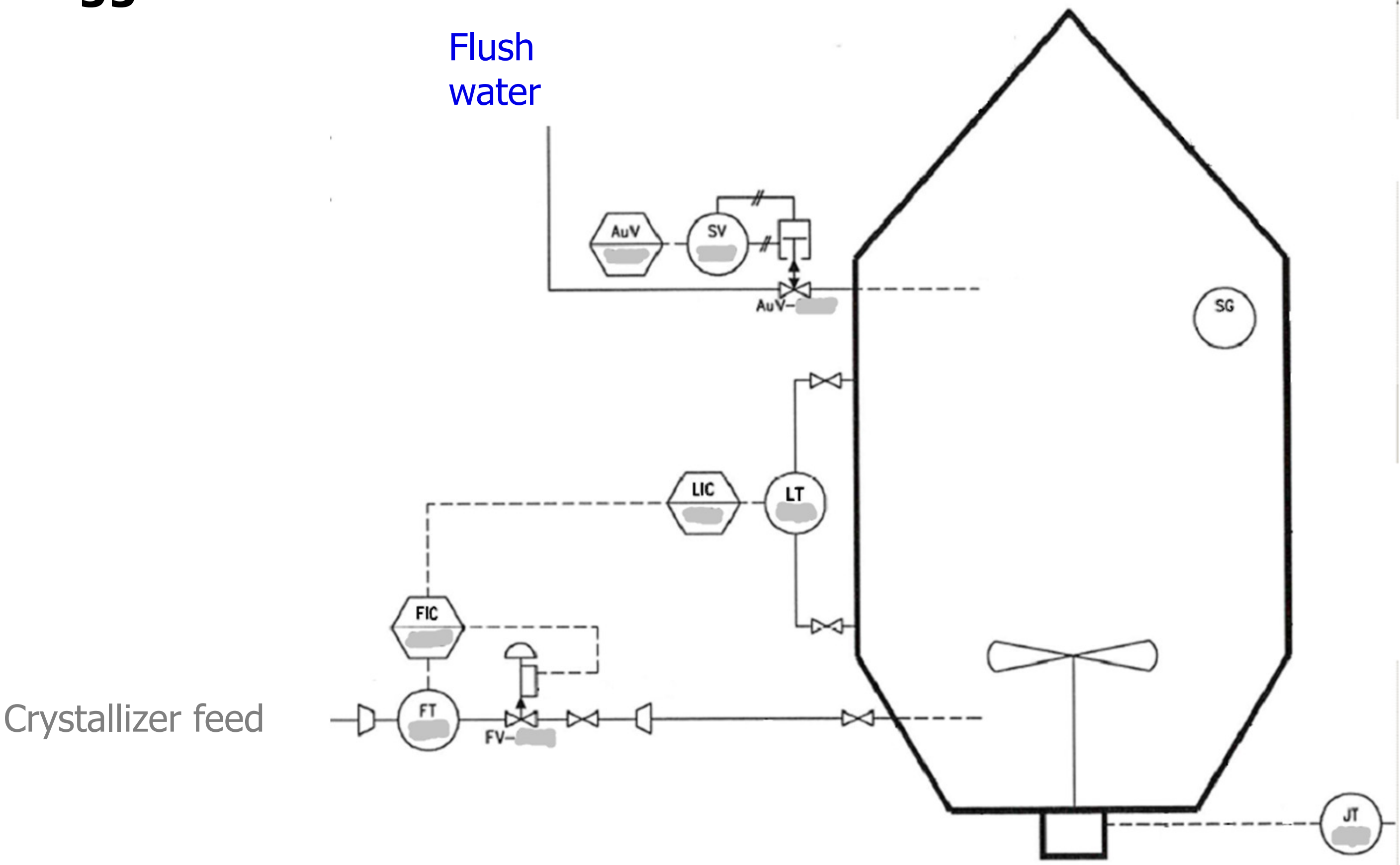
This structure is superior if aldehyde feed varies, e.g. during a start-up.



# Crystallizer: Level disturbance from wash sequence

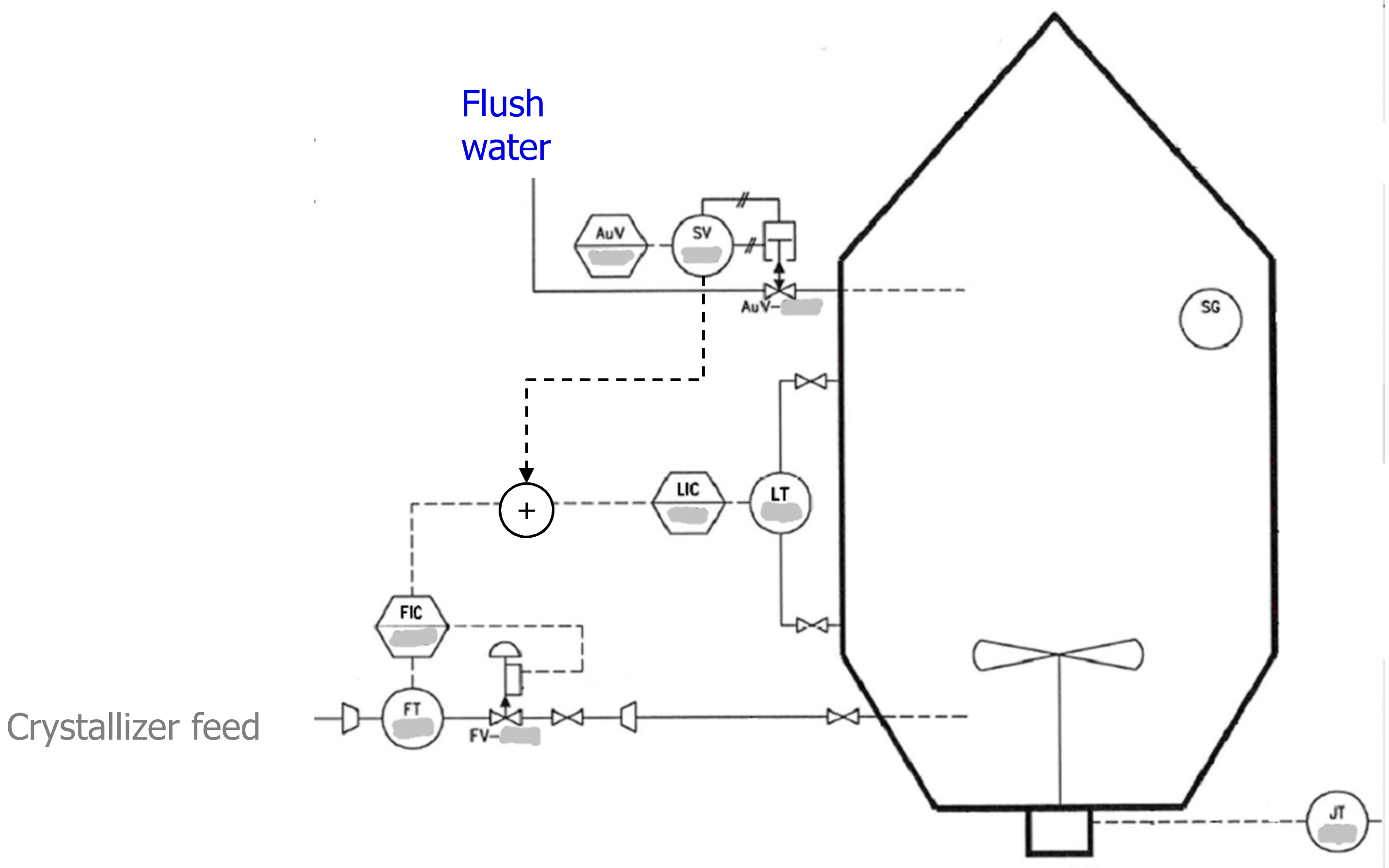
- A crystallizer is automatically flushed with water, once every second hour. The water flow is large enough to affect level.
- It's important to keep a steady level.

**Improvement suggestions?**



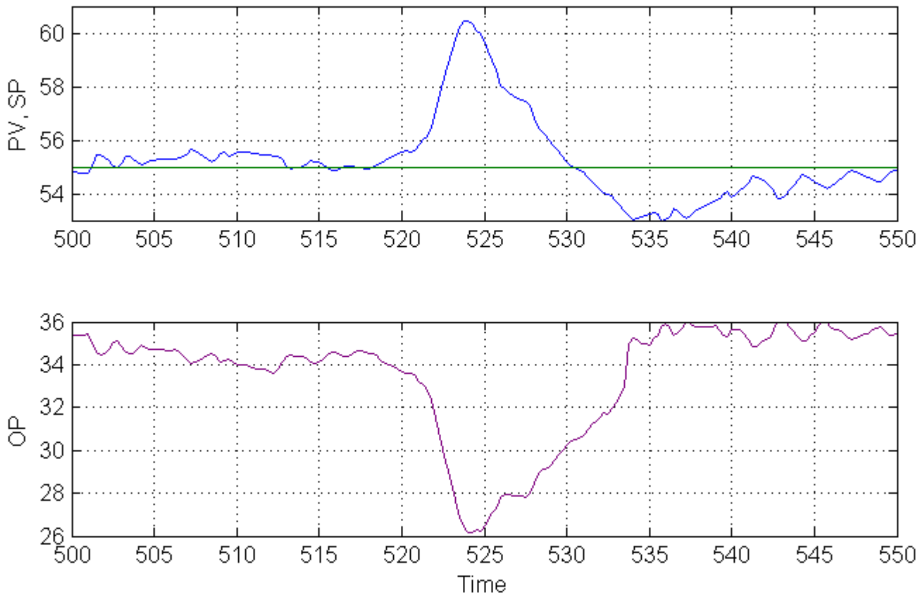
# Crystallizer: Feedforward reduces level variations

- A FF from on-off-valve to level controller reduces level variations.
- Thus, it's ok to make a FF from a discrete (binary) variable.

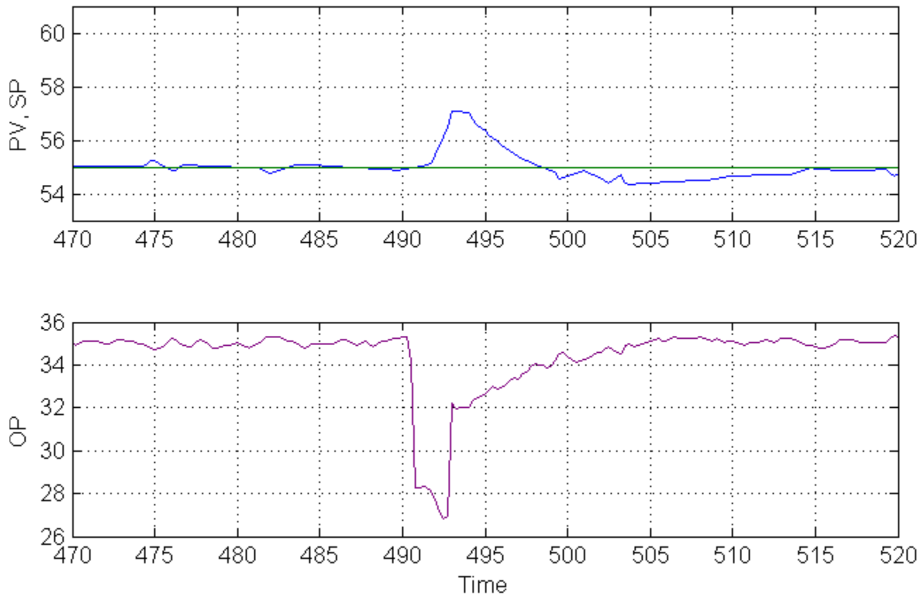


# Smaller level variations in crystallizer

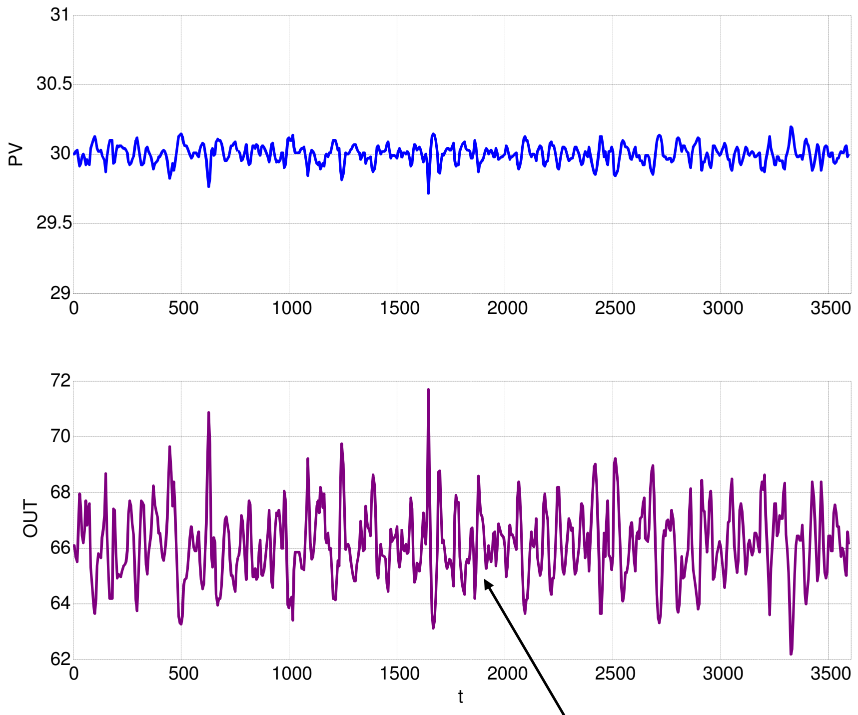
Without feedforward from on-off valve

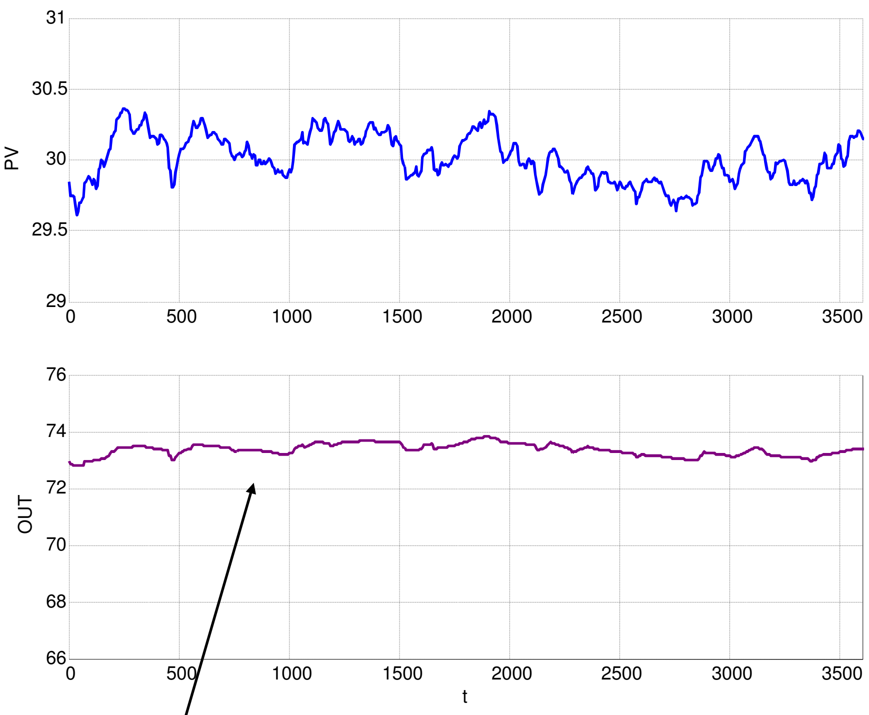



With feedforward



# Level in distillation column: What is the overall control objective?

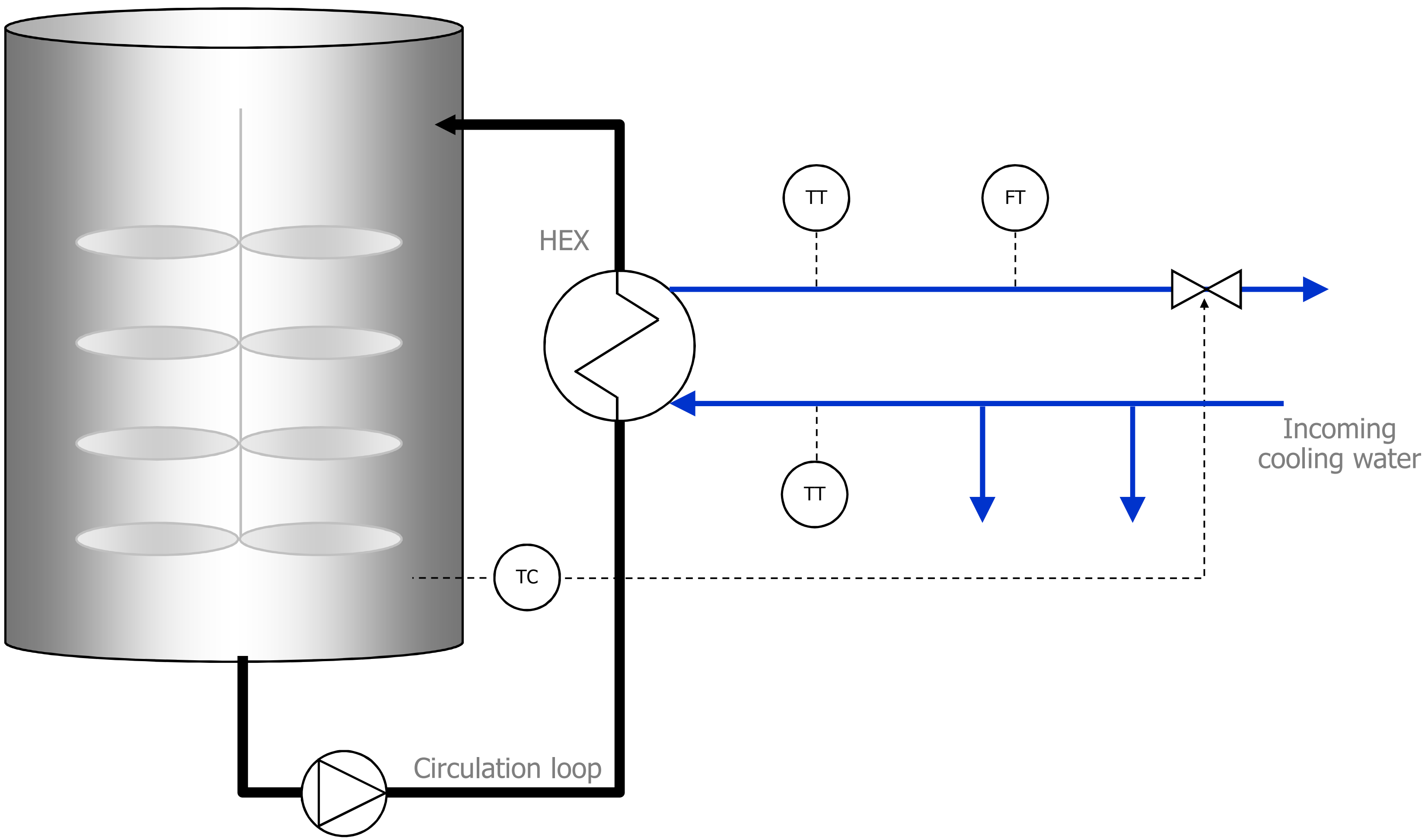


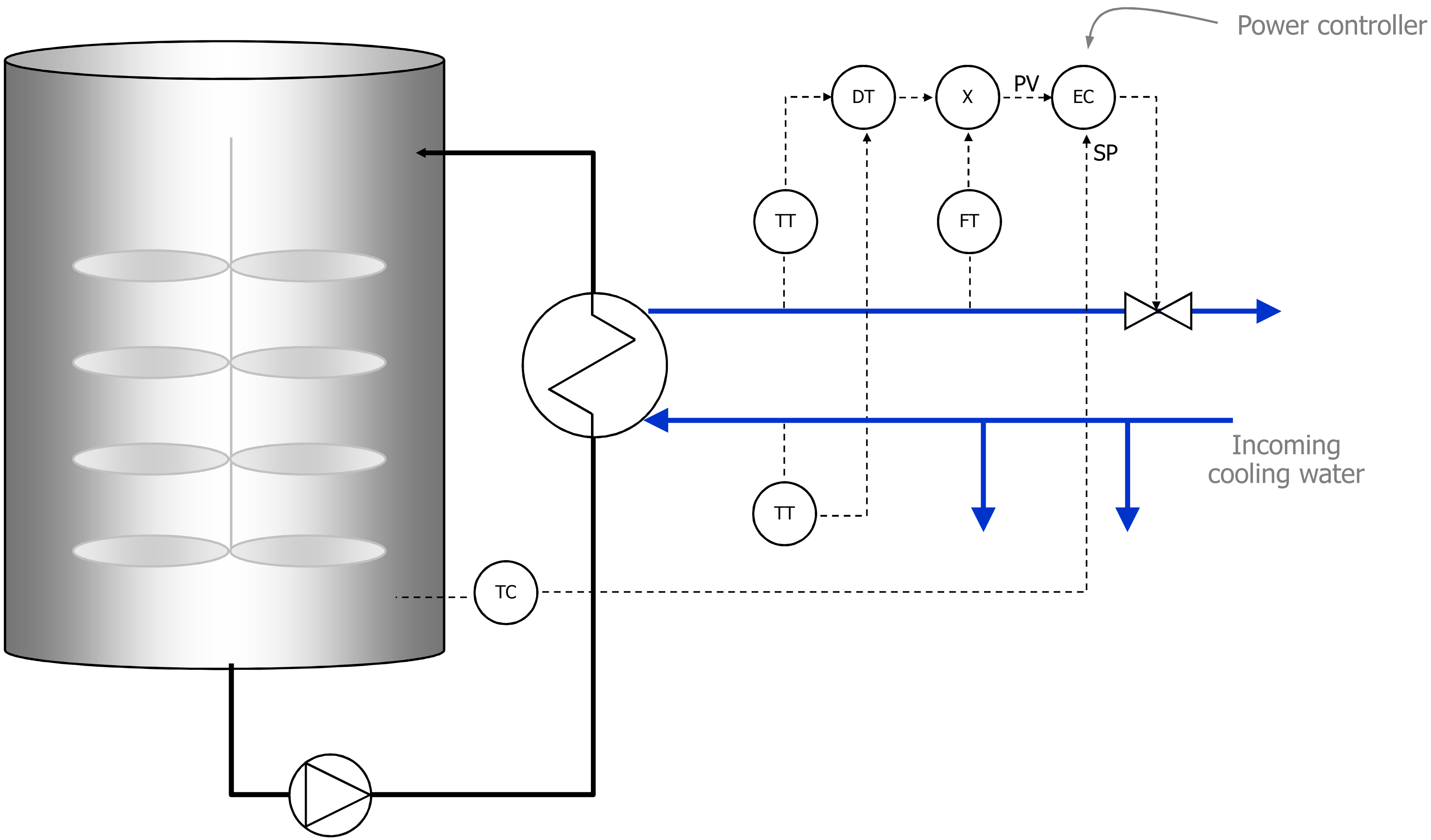

This flow is the feed to the next column (product column)

# Exothermic reactor temperature control

The reactor solution is circulated through a heat exchanger (cooler).
The reaction is very exothermic: it is important to control the temperature.
Typical variations/disturbances: Cooling water header pressure, CW temperature



# New control structure: Power control

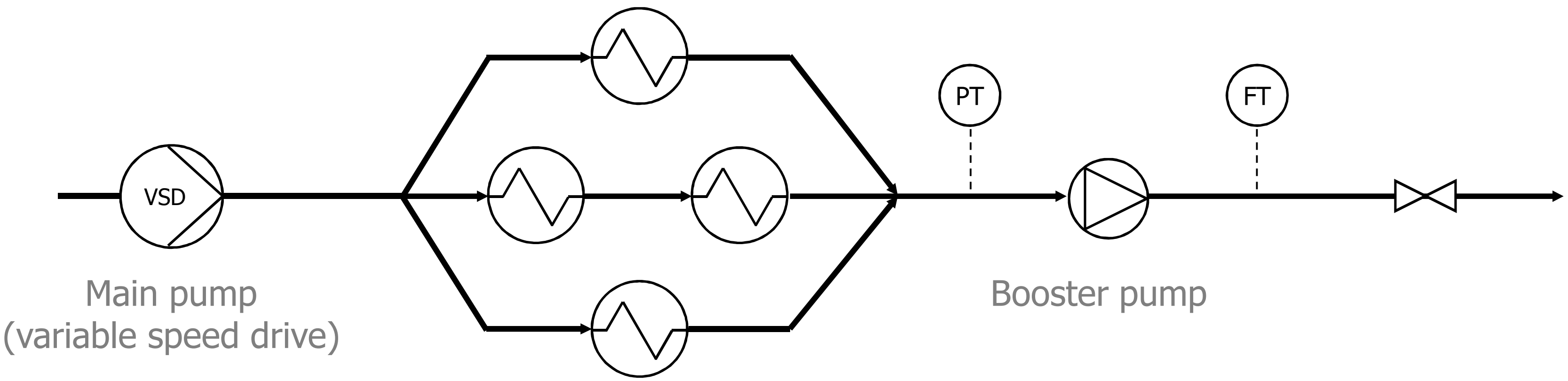


# Pressure and flow control in feed line

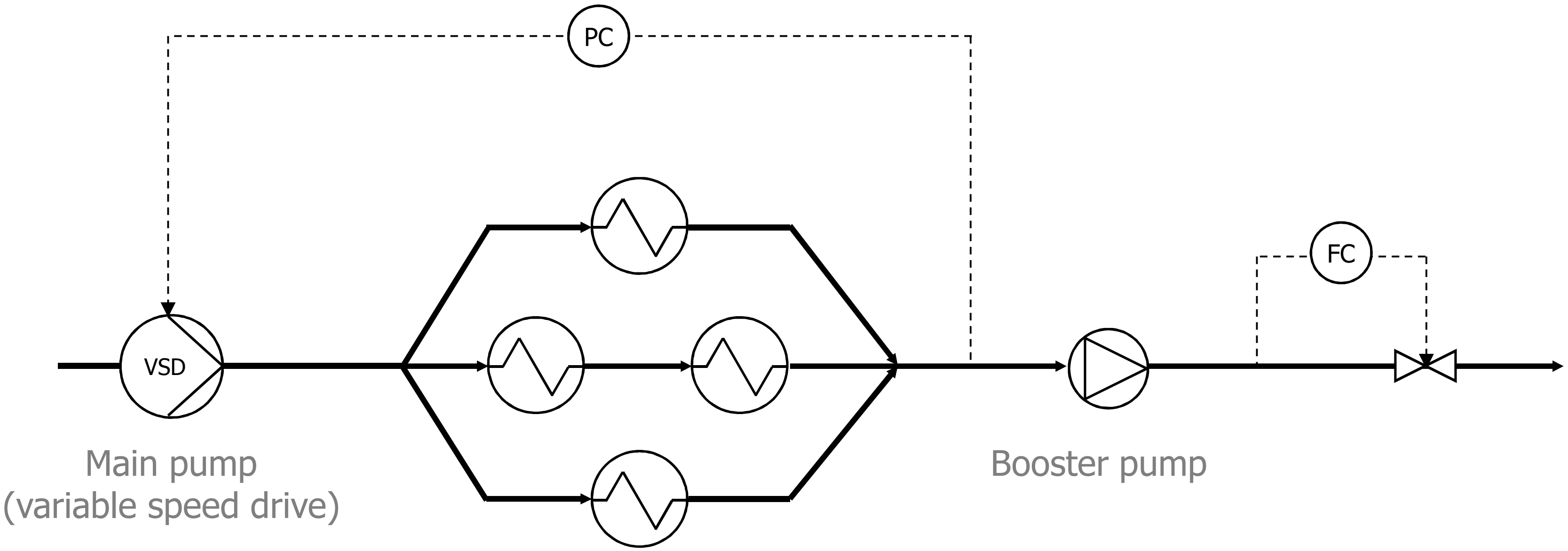
Requirements: Control flow and make sure that the booster pump does not cavitate

If possible: Minimize pumping energy losses

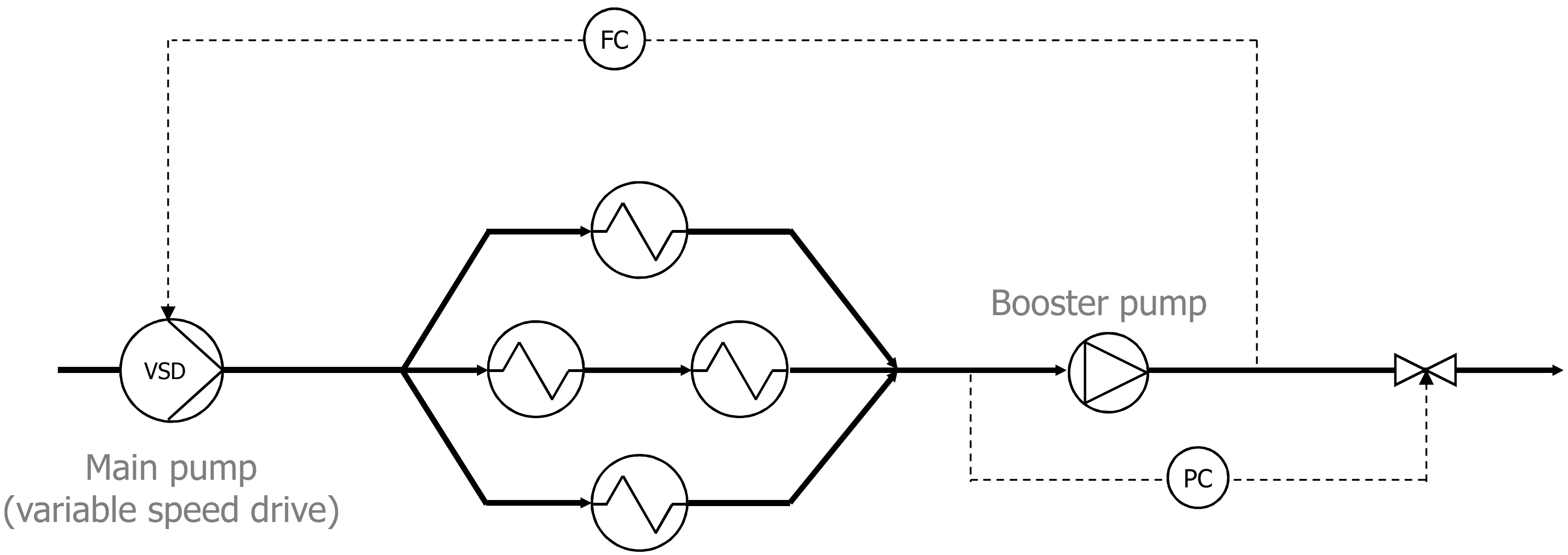
Manipulated variables: Valve position and main pump speed



# Is this possible?



# Is this possible?

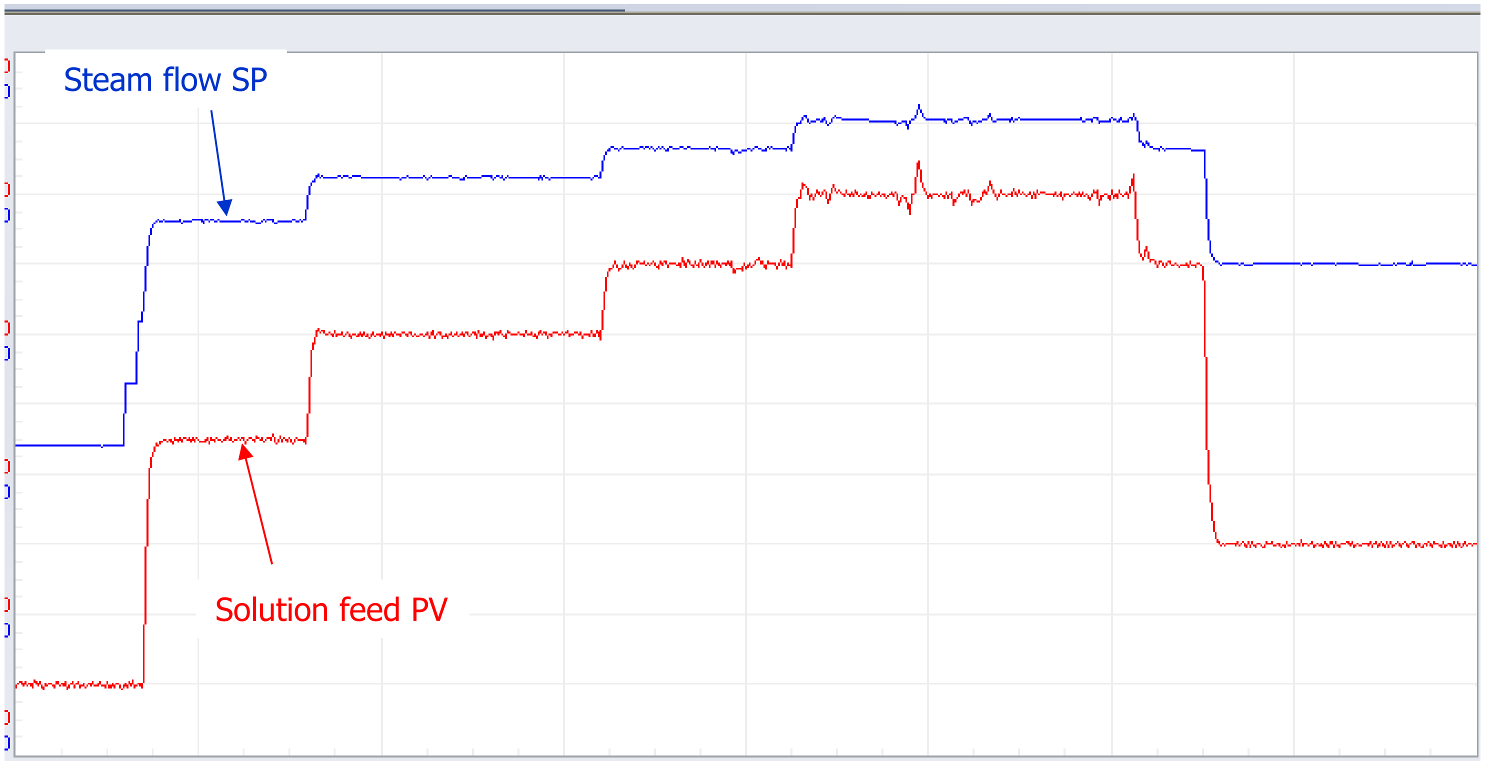



Minimize pumping energy losses by choosing PC setpoint
as low as possible, without risking cavitation.
Then the PV will be 100% open when possible.

# Evaporation proces: Ratio control issue

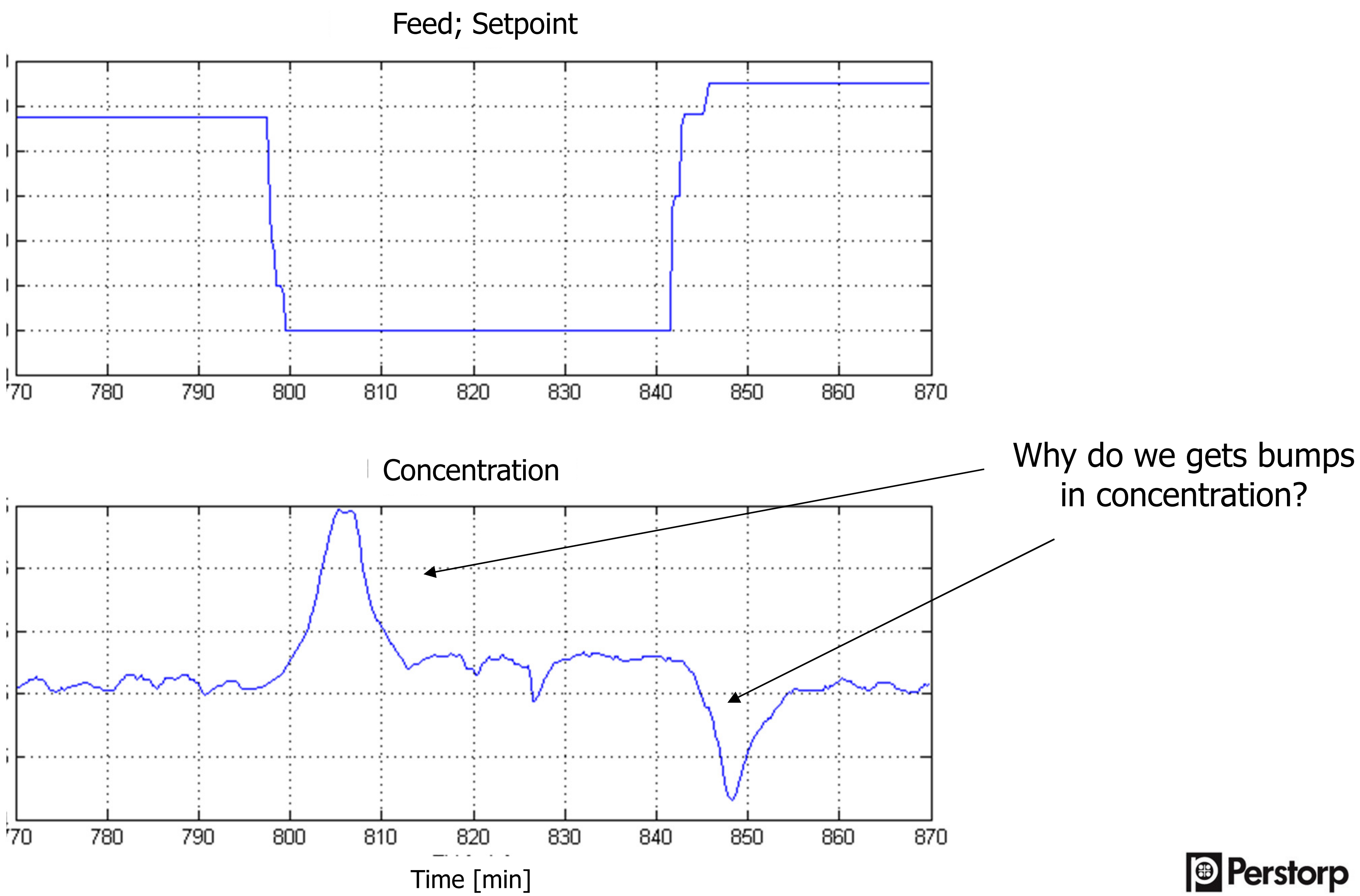
Process: Evaporation. Increase solution concentration by evaporating water.

Solution feed flow is master, steam flow in ratio against feed.



Steam flow SP
Solution feed PV

# Disadvantage with this structure



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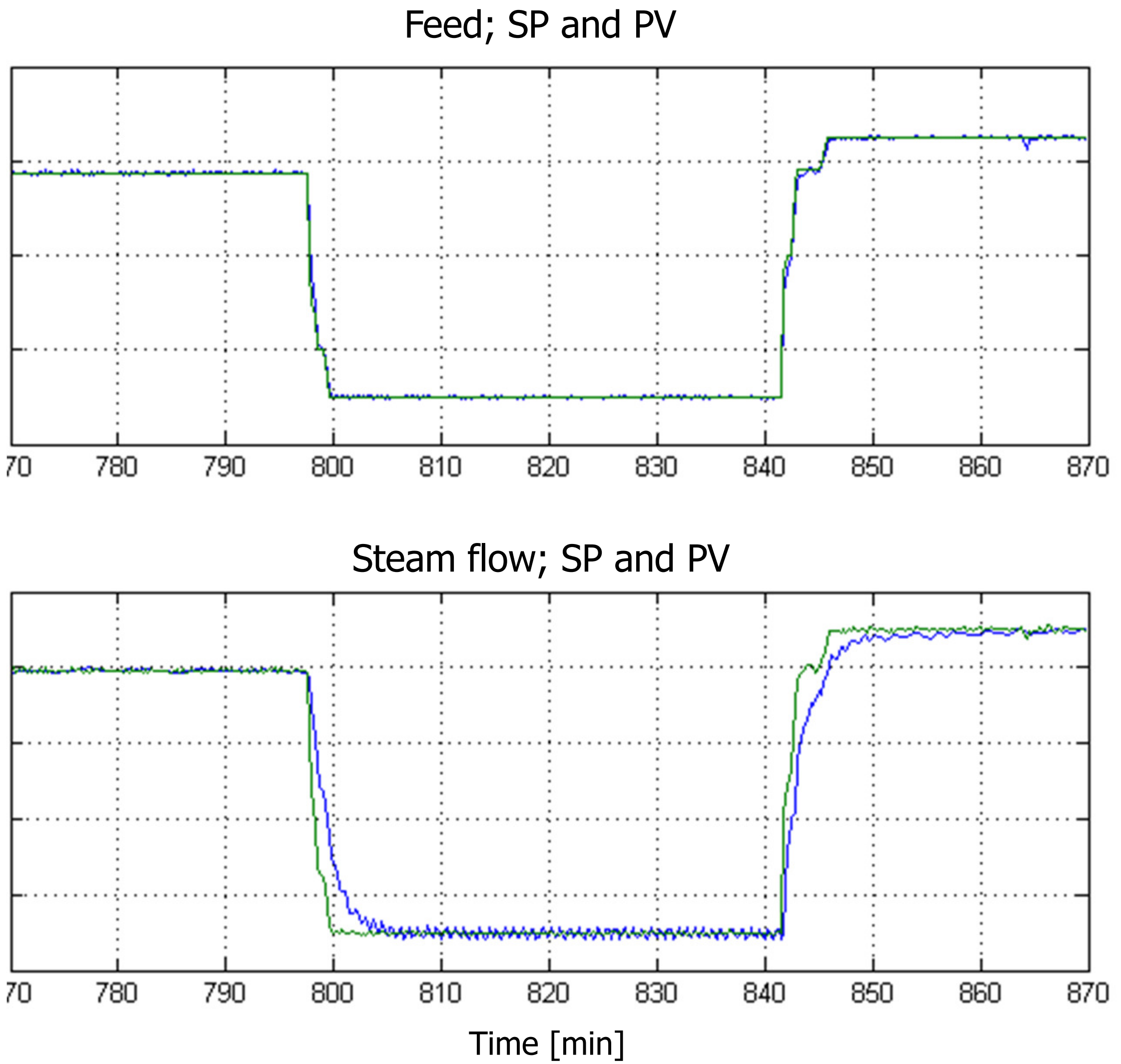
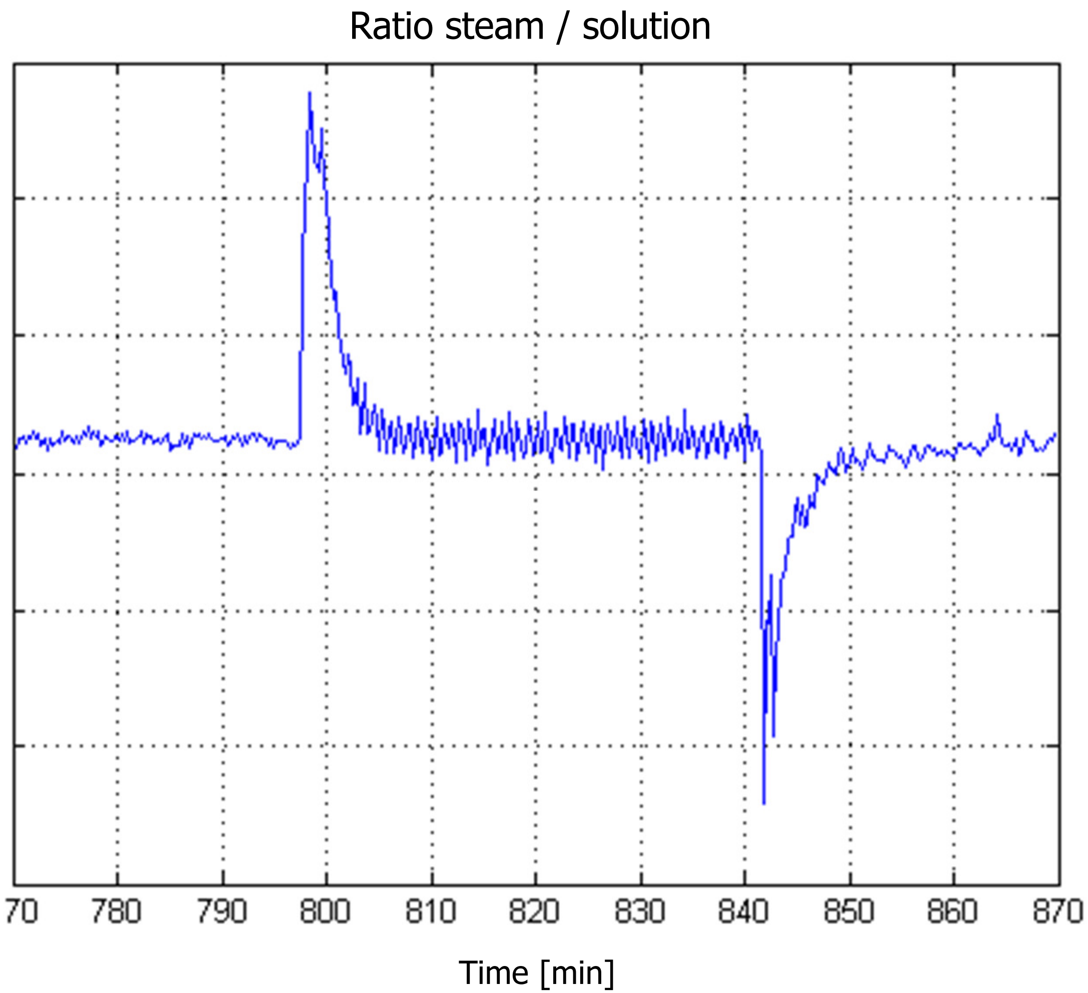
# Same scenario: comparison feed – steam

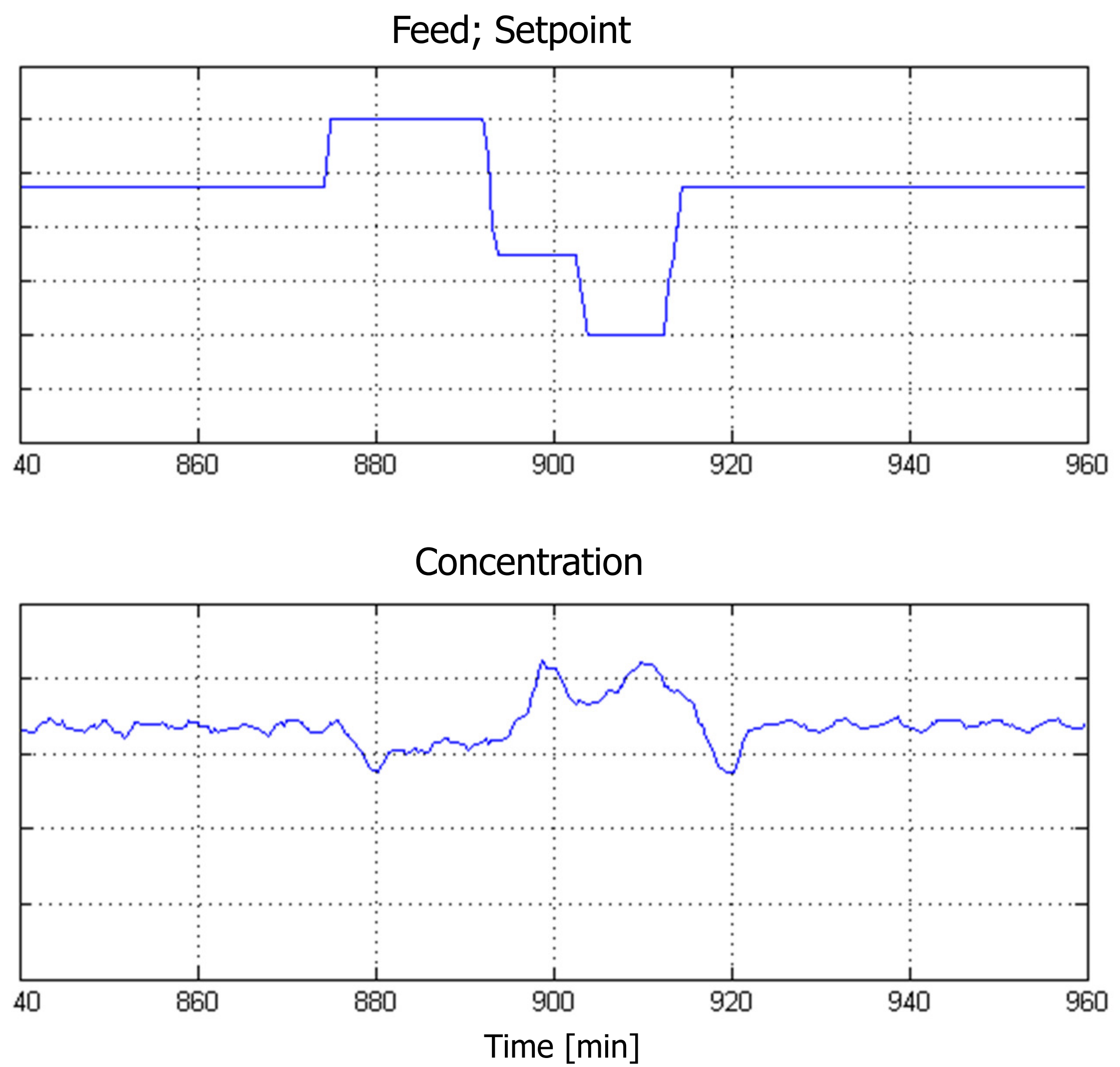


# The ratio steam / solution not constant during SP-changes

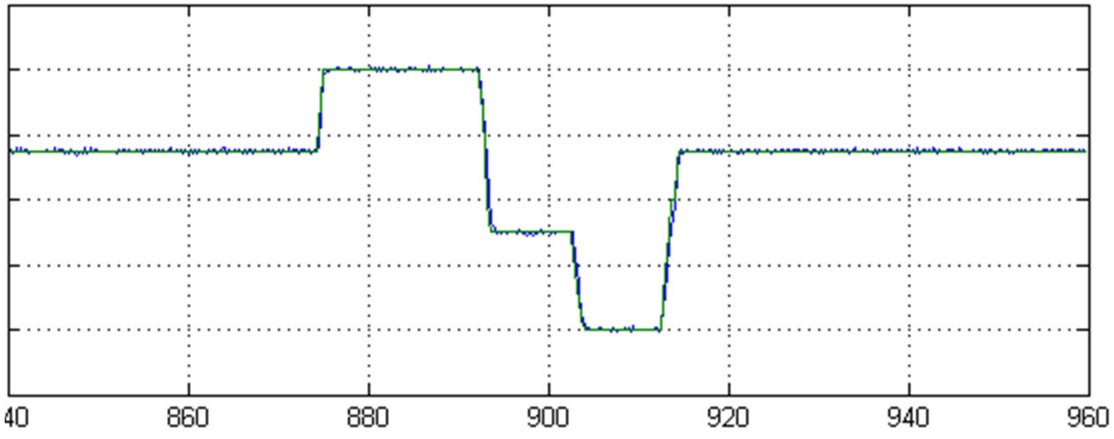



**What should we do in order to reduce the deviation in ratio?**

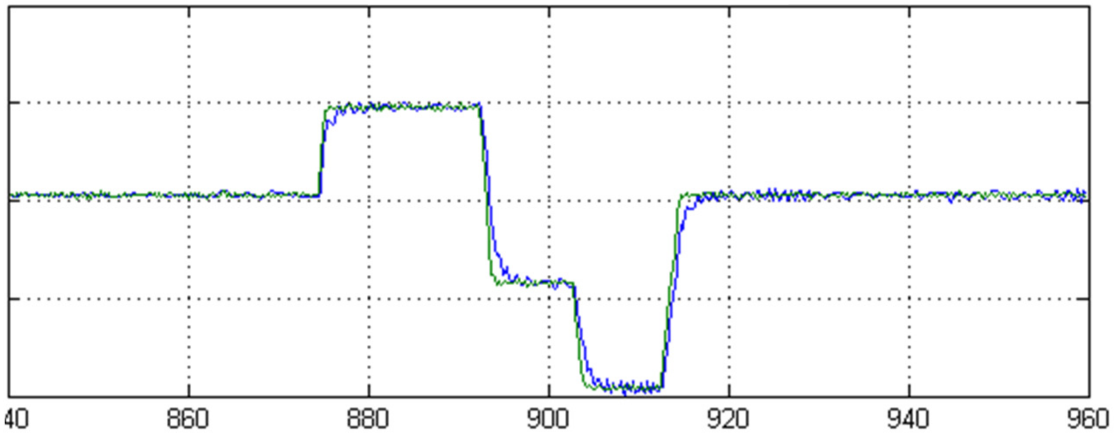
# Speed up steam flow controller! Result:



Feed; SP and PV

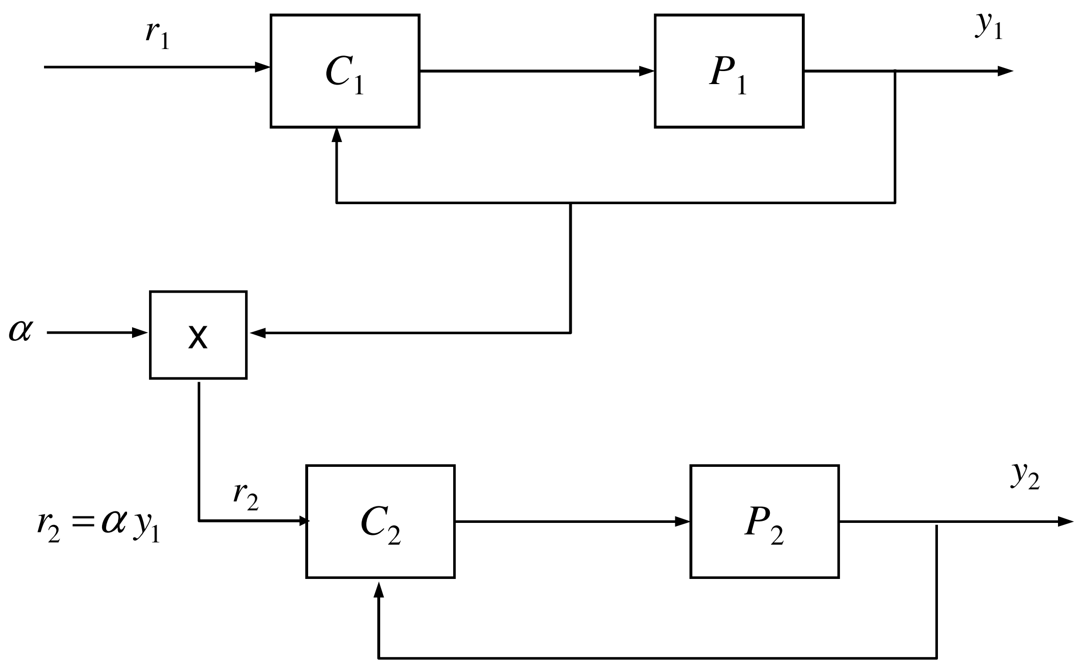



Steam flow; SP and PV



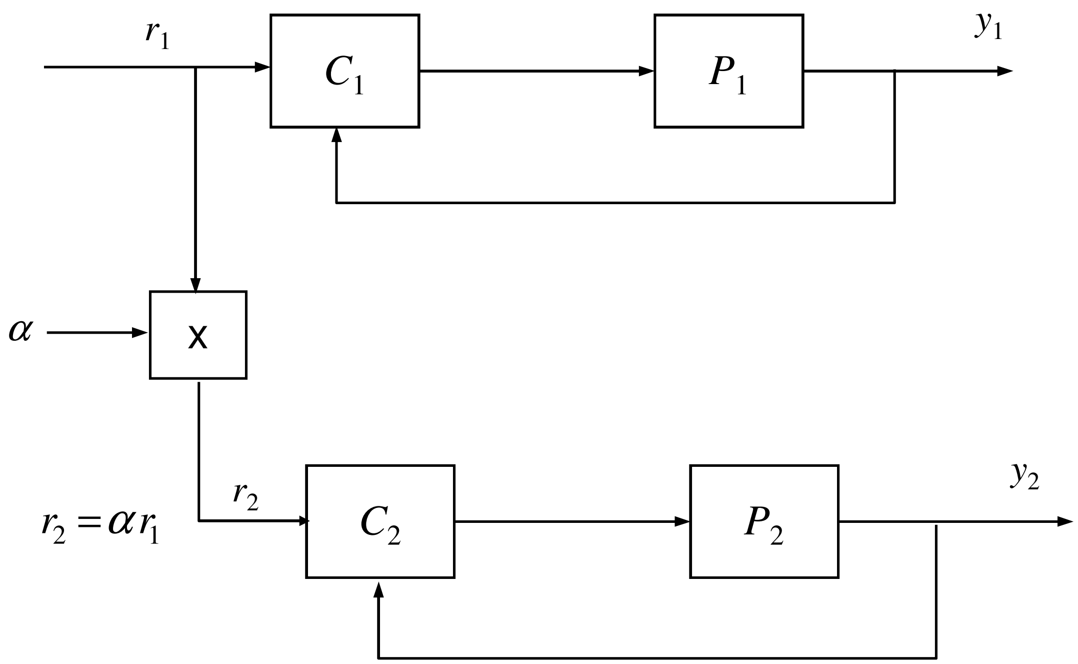
# Current control structure

Is there a different structure that would reduce deviations even more?



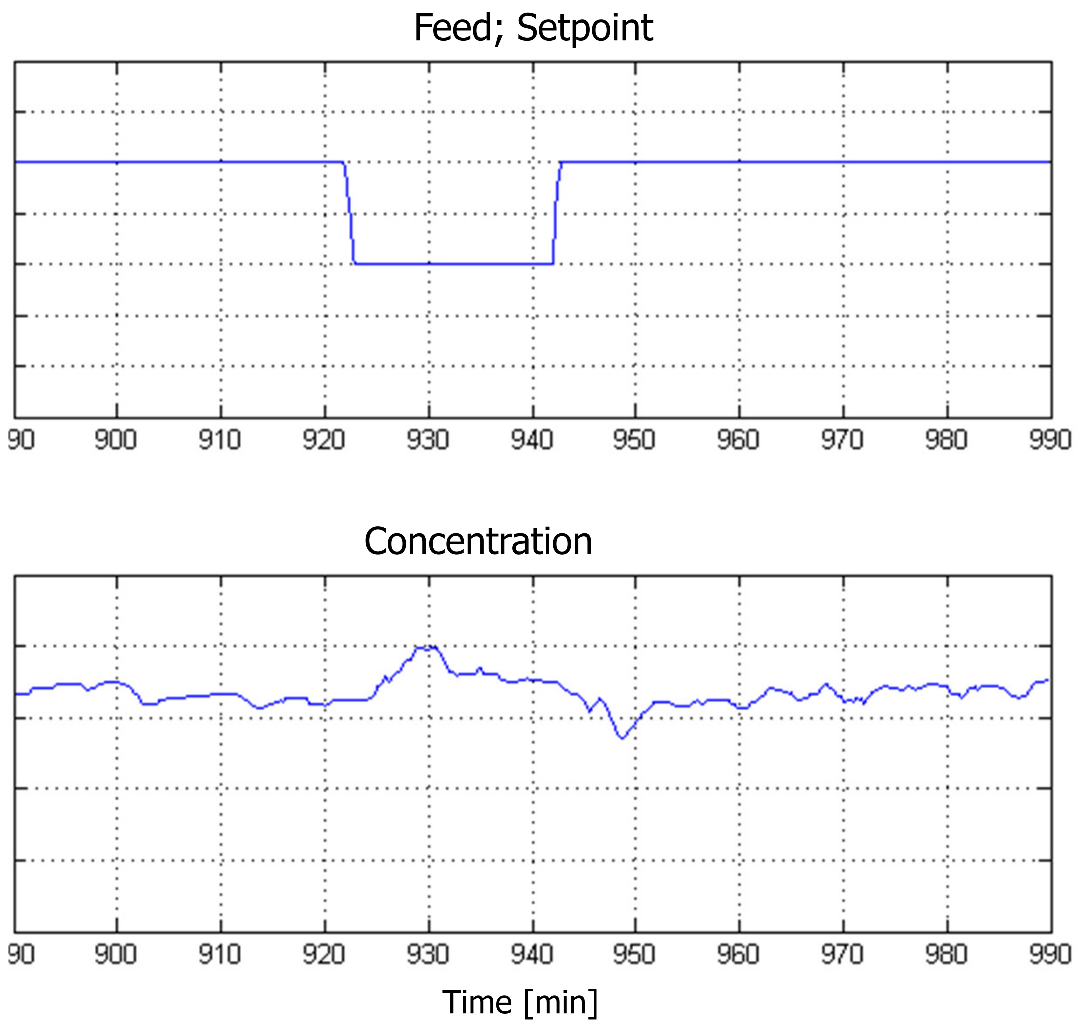
# Alternative structure

The SP of the slave is calculated from master **SP** rather than master **PV**!




What are the disadvantages of this scheme?

# Test run: Ratio against SP



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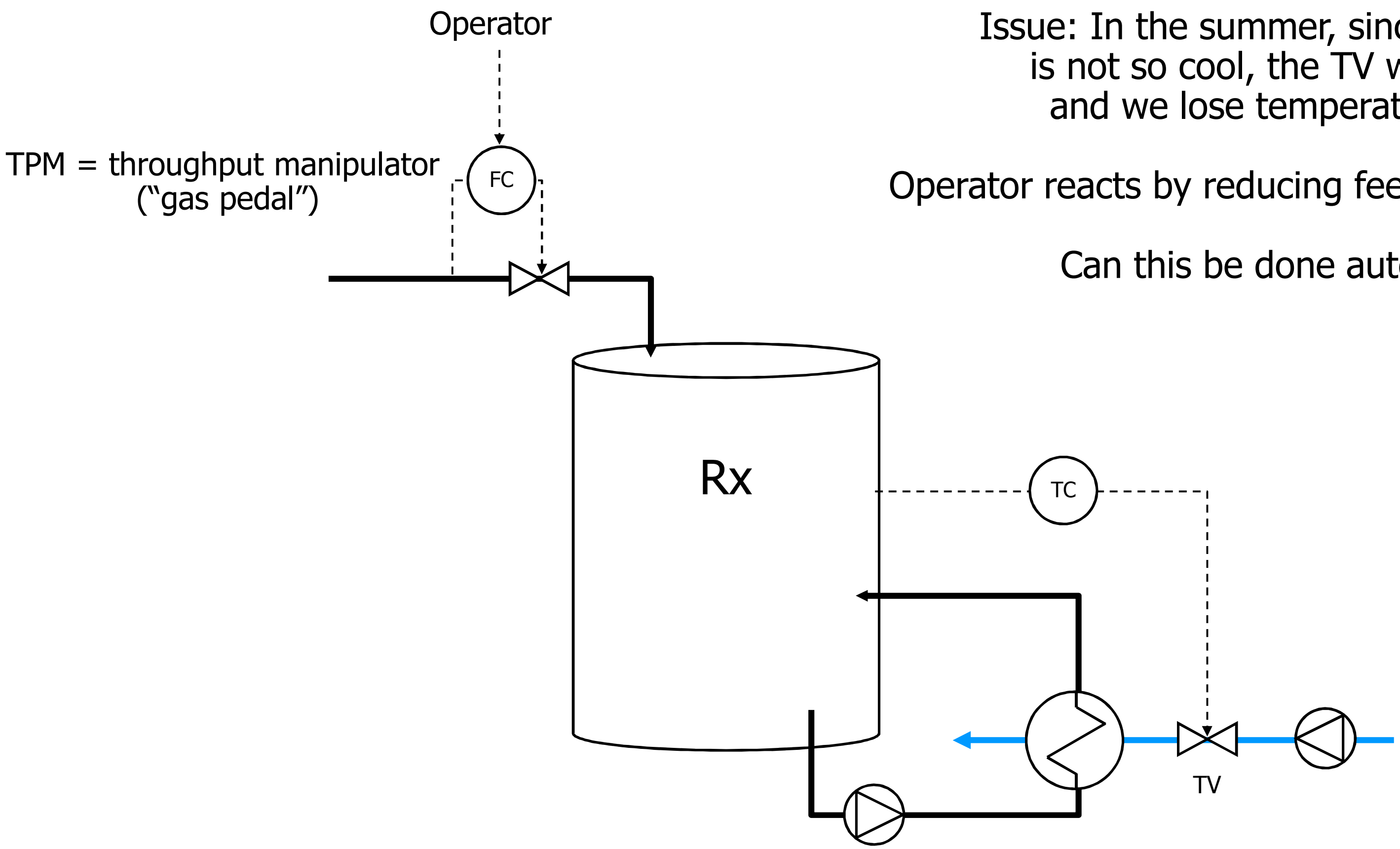
# Example: Cooling capacity limited Rx; Normal operation



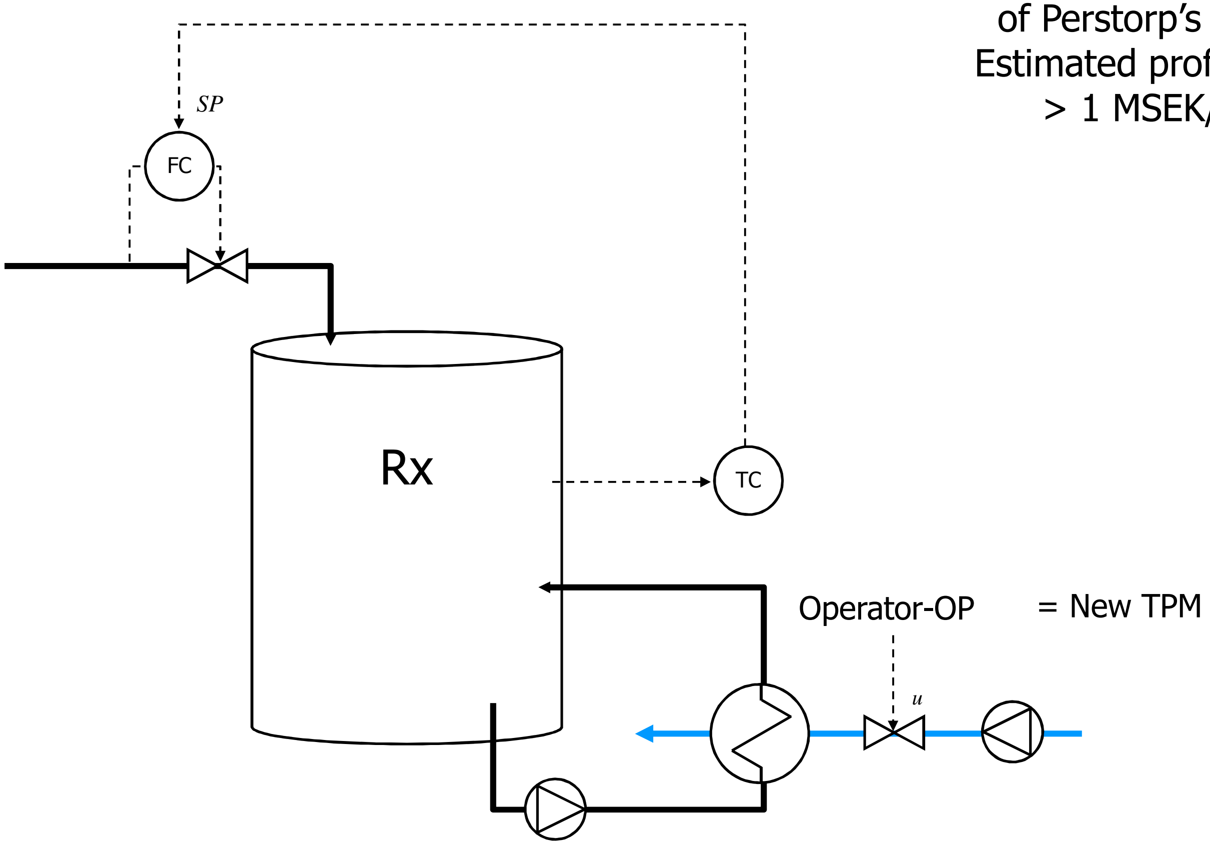

Issue: In the summer, since cooling water is not so cool, the TV will open fully and we lose temperature control.

Operator reacts by reducing feed (=production rate).

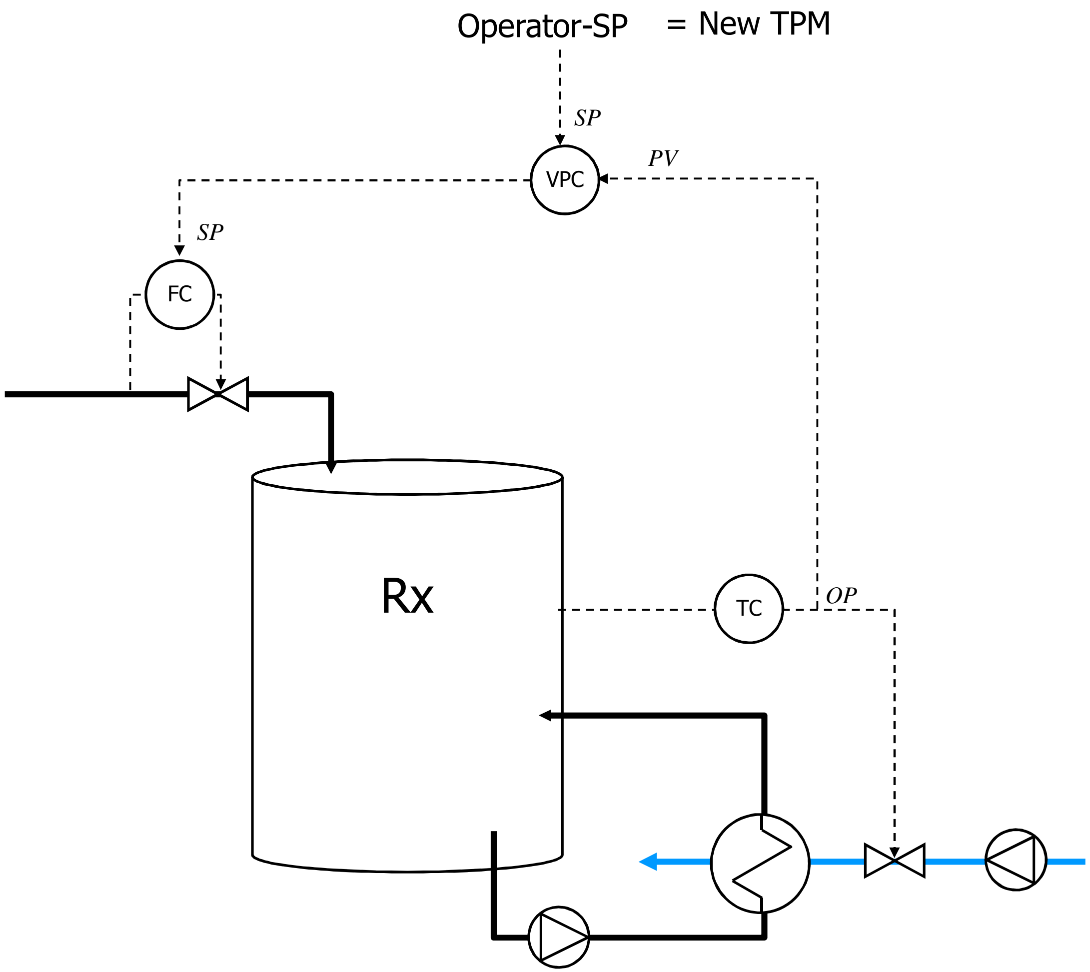
Can this be done automatically?

# Cooling capacity limitation; Maximizing control 1

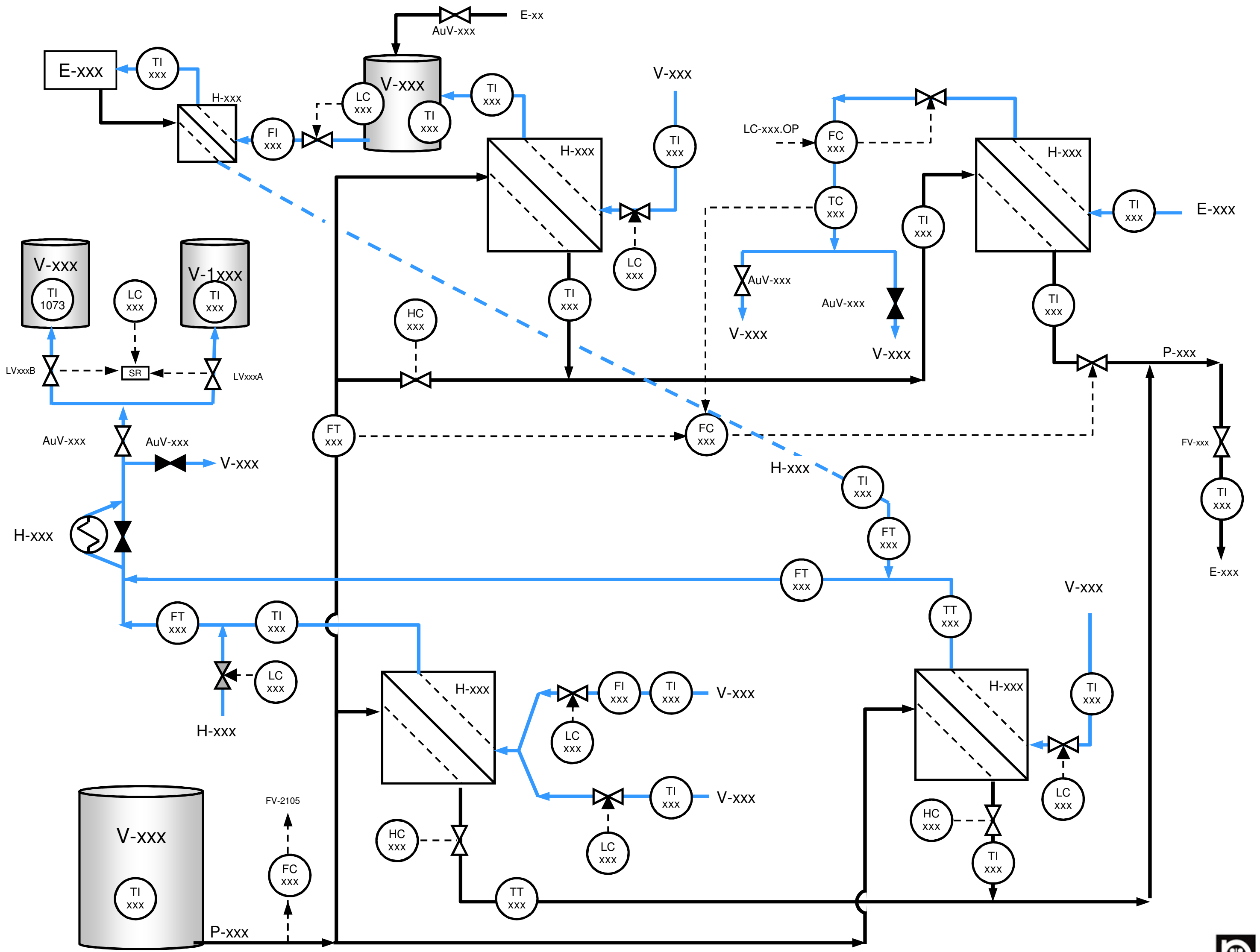
Implemented in one of Perstorp's plants. Estimated profitability: > 1 MSEK/year



# Cooling capacity limitation; Maximizing control 2



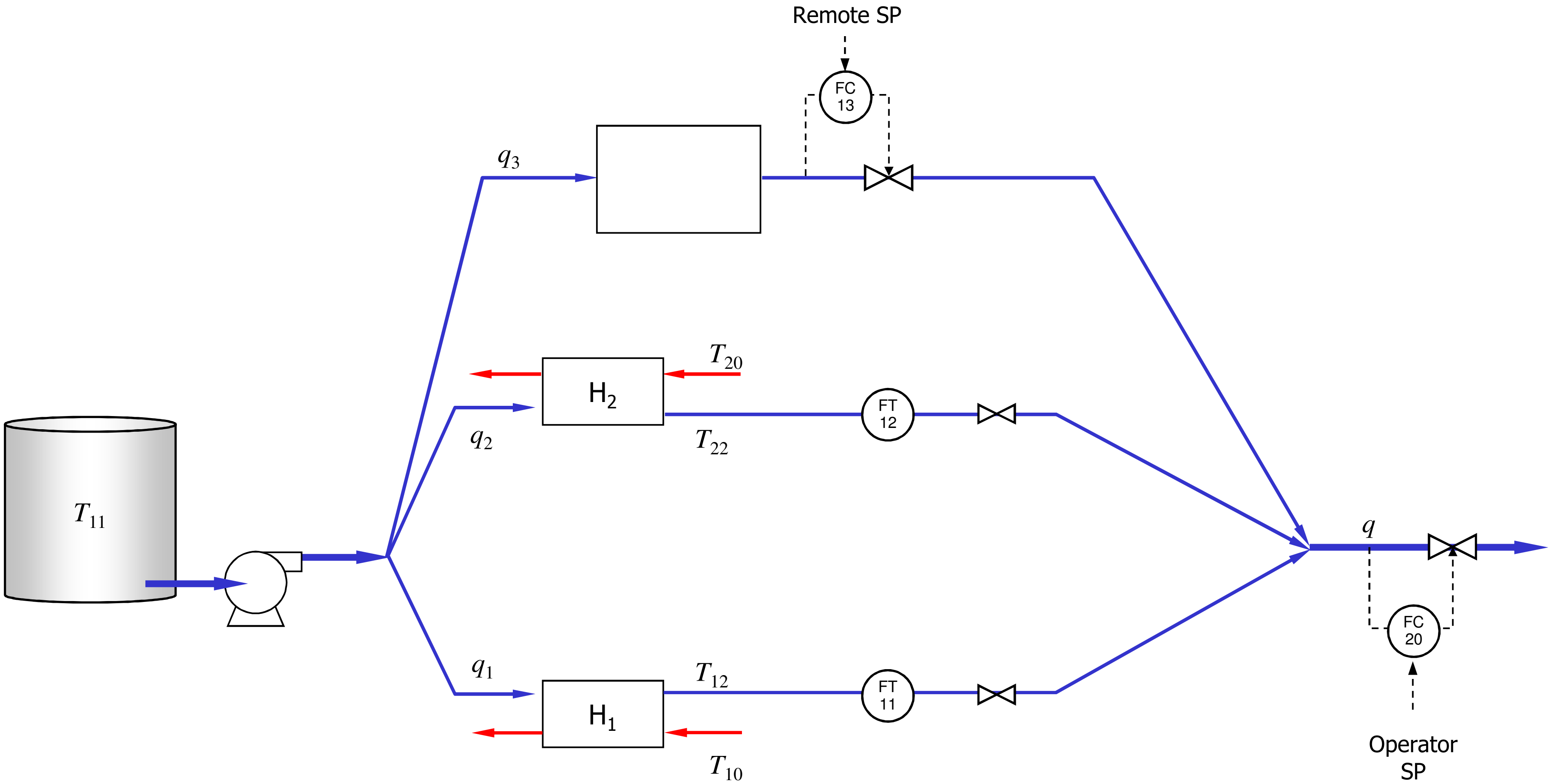
# Example: HEX network optimization



Self-optimizing control solves the problem

Controlled variable (invariant): $(T_{12}-T_{11})^2(T_{20}-T_{11})-(T_{22}-T_{11})^2(T_{10}-T_{11})$

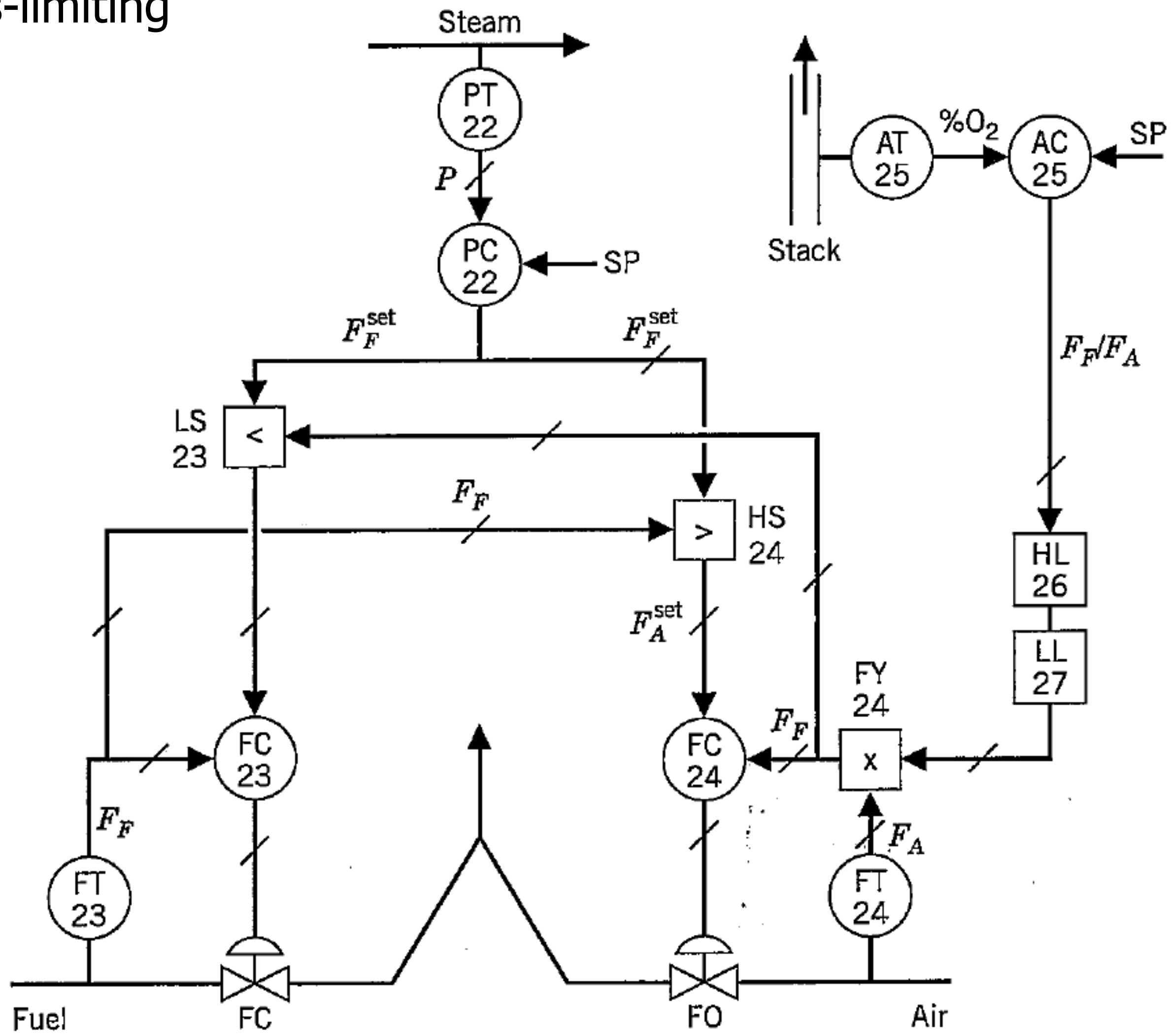
Manipulated variable: The ratio of flow $q_2$ to $q_1+q_2$



## Fuel – air – control for a steam boiler

Make sure that air is always in excess, both when load increases and decreases, while controlling both the steam pressure and the residual oxygen.

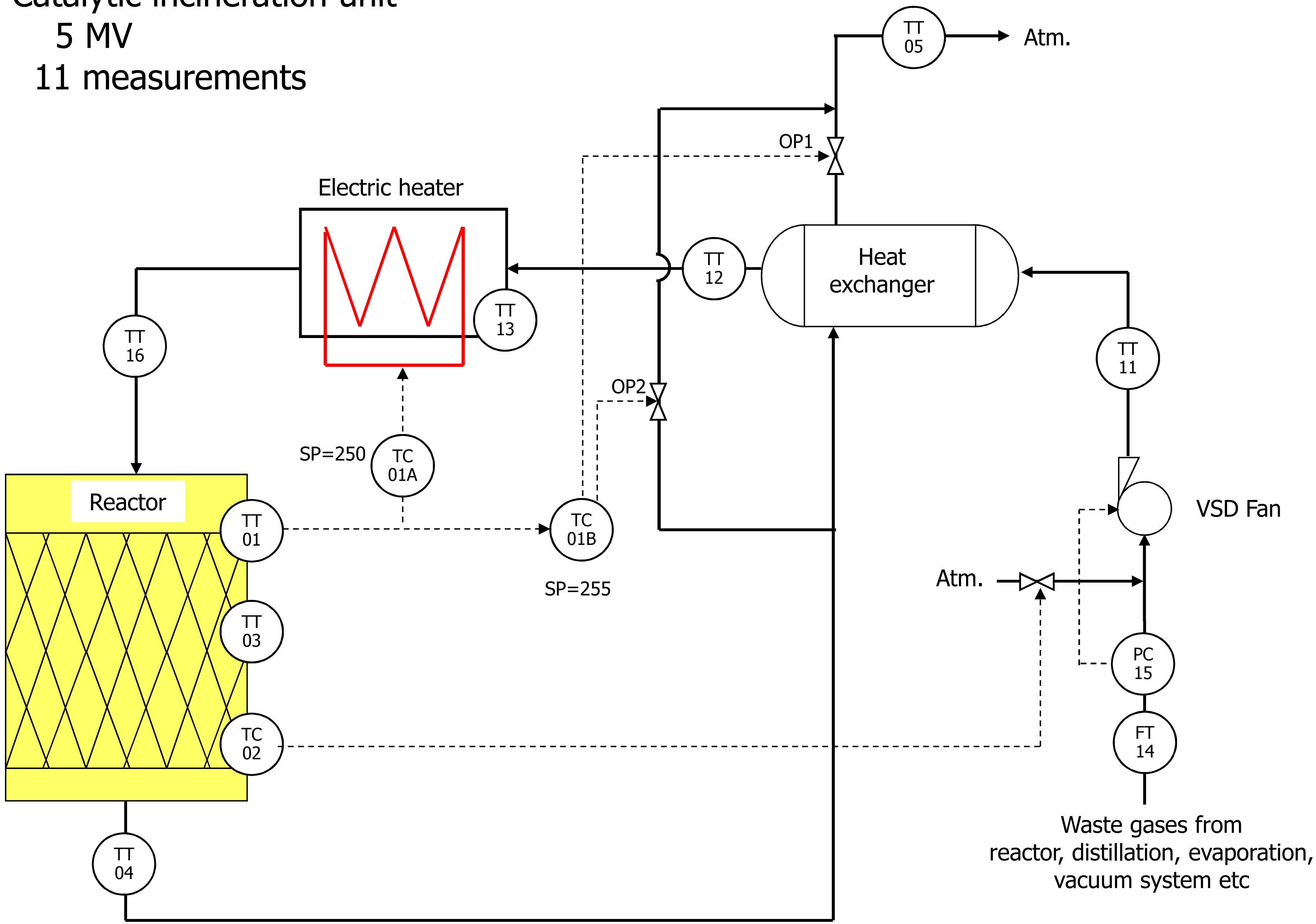
”Classical solution” = cross-limiting



# Catalytic incineration unit

5 MV

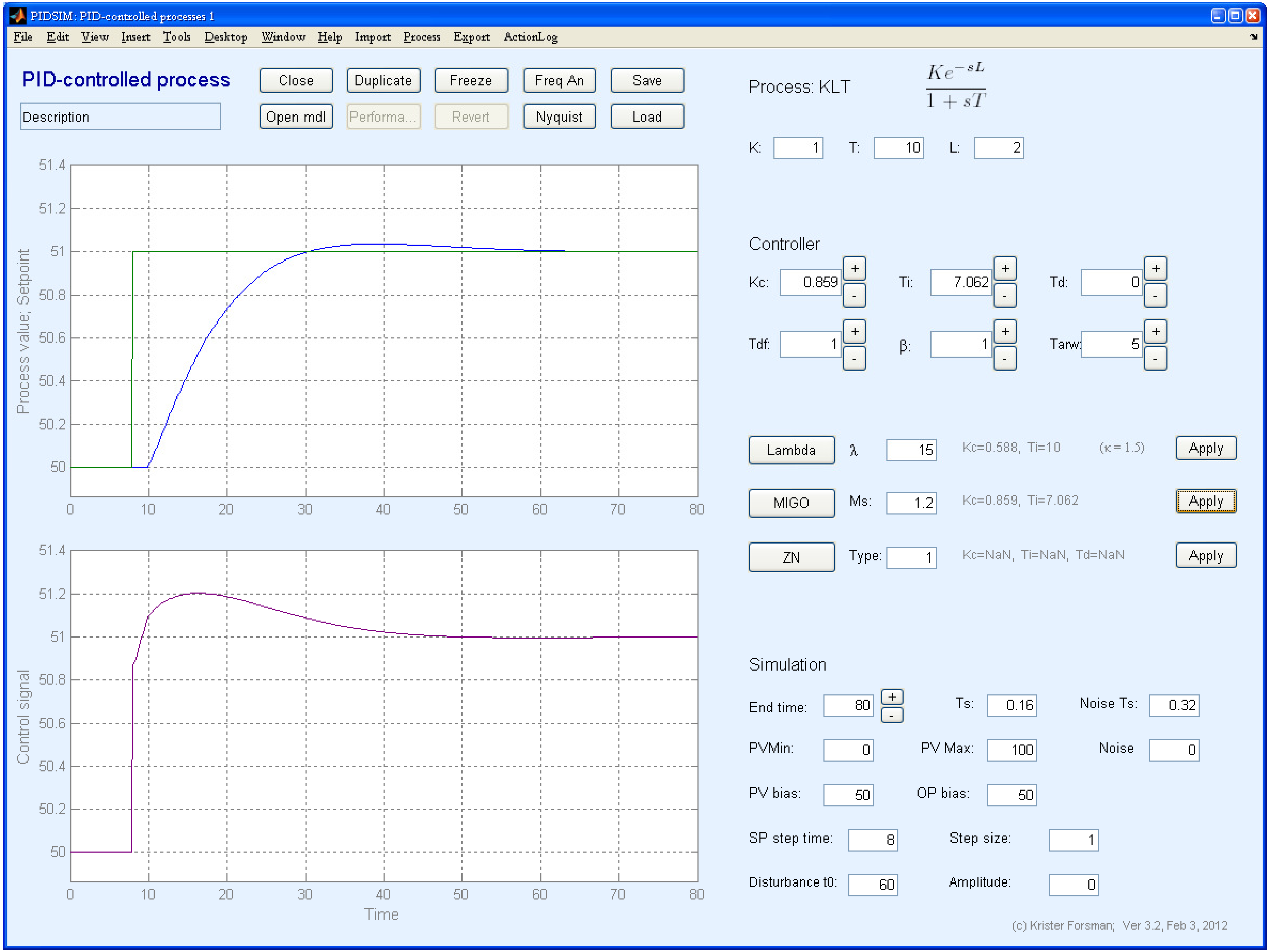
11 measurements



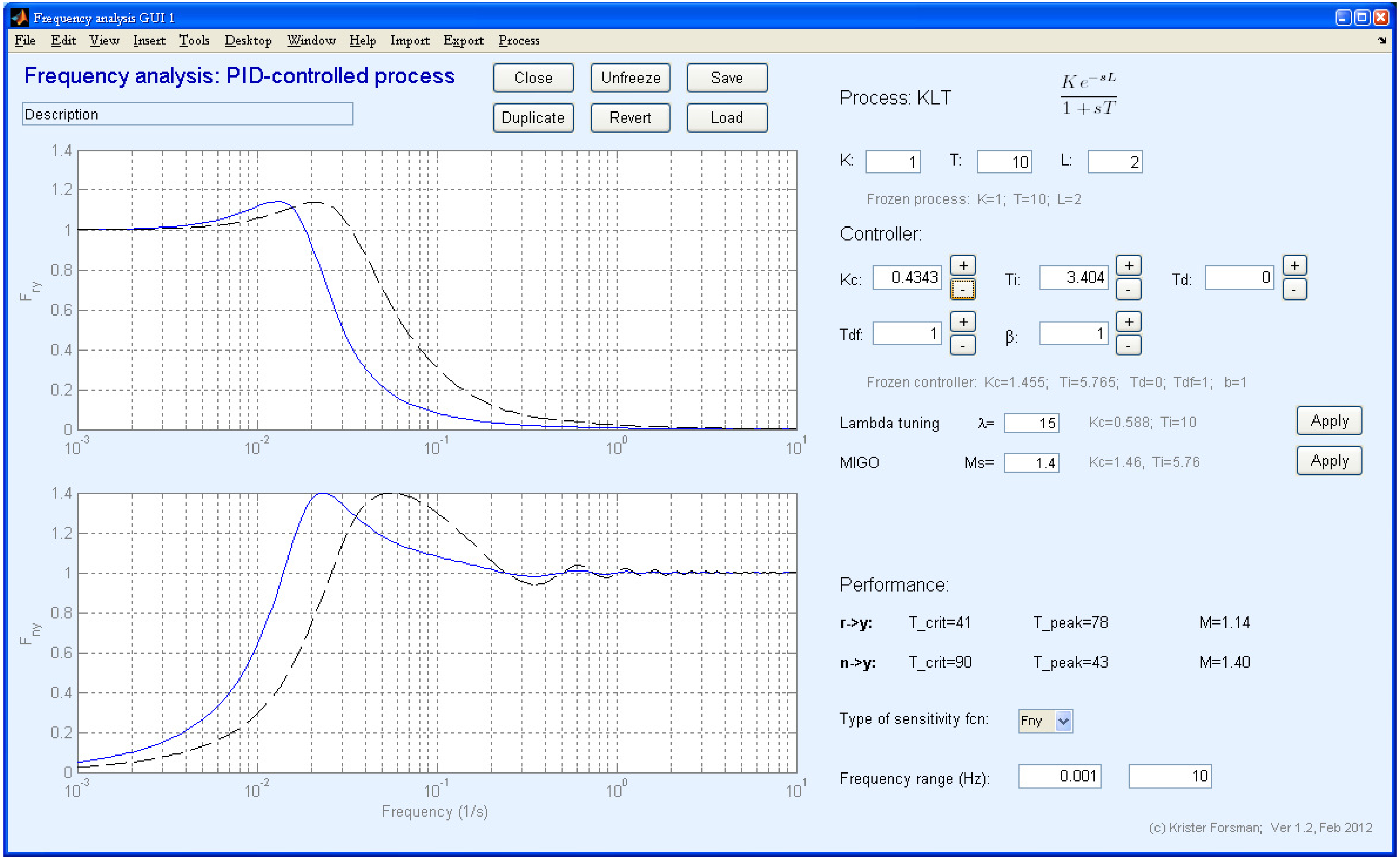
# Tools are centered around Matlab

- The most important tools in the optimization work is Matlab and IP21.
- We have developed our own libraries for
  - Data collection from IP21
  - Data analysis
  - Simulation of controllers and control structures
  - Identification
  - Assessment of control performance

- Some examples of tools below

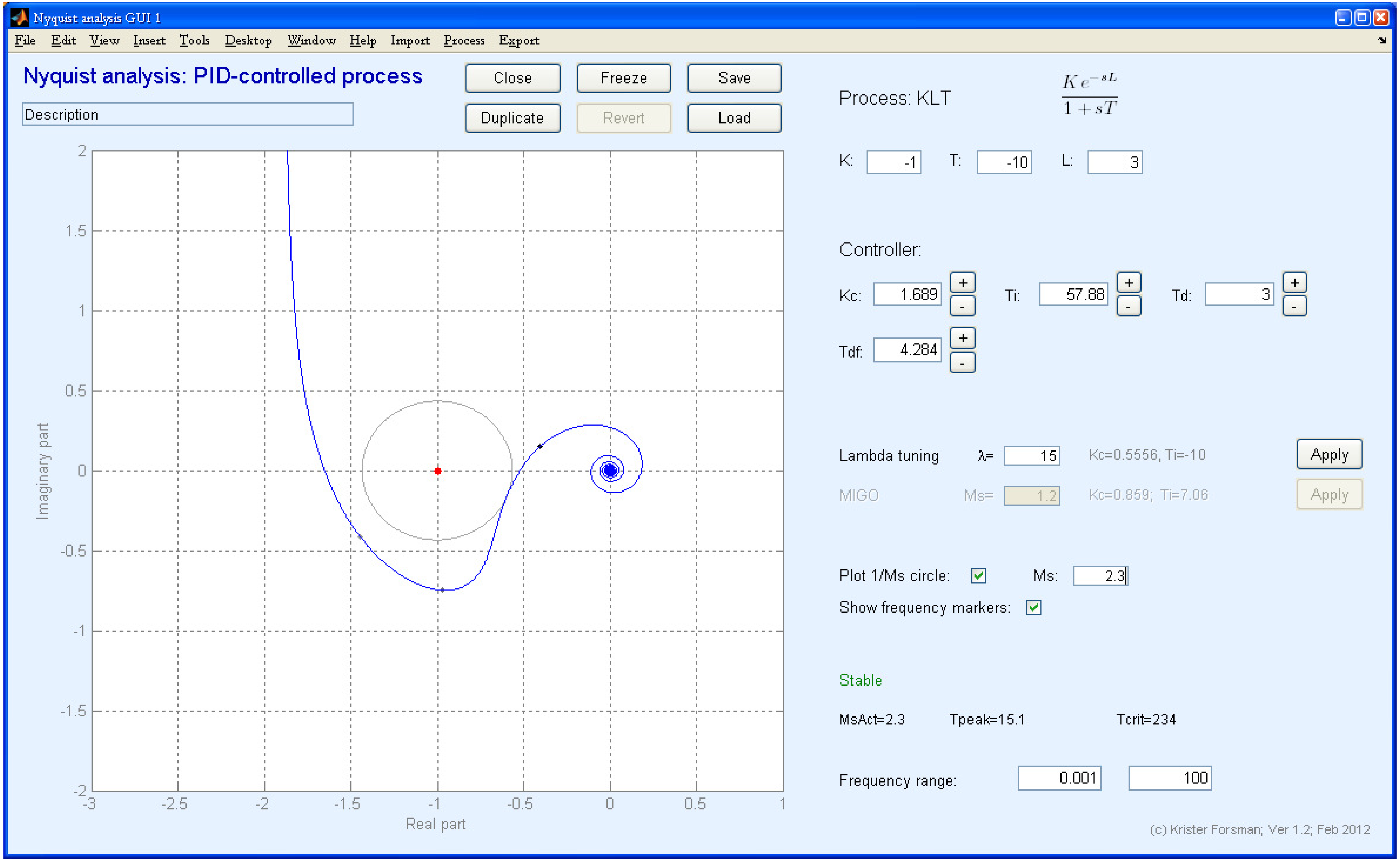


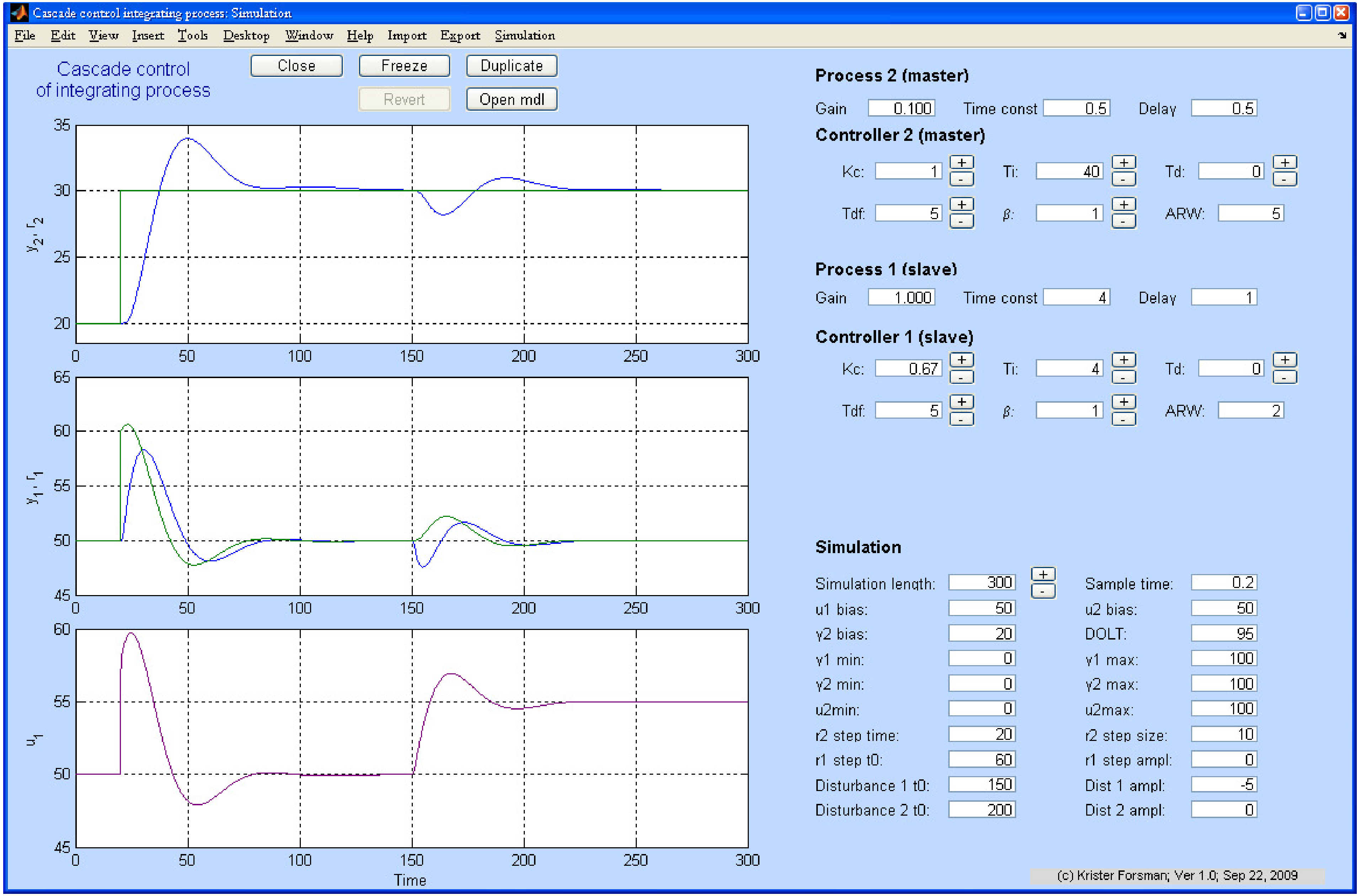
PIDSIM: PID-controlled processes 1
File Edit View Insert Tools Desktop Window Help Import Process Export ActionLog
PID-controlled process
Description
Close
Duplicate
Freeze
Freq An
Save
Open mdl
Performa...
Revert
Nyquist
Load
Process value; Setpoint
Control signal
Time
Process: KLT
$\frac{Ke^{-sL}}{1+sT}$
K: 1
T: 10
L: 2
Controller
Kc: 0.859
Ti: 7.062
Td: 0
Tdf: 1
β: 1
Tarw: 5
Lambda
λ 15
Kc=0.588, Ti=10
(κ = 1.5)
Apply
MIGO
Ms: 1.2
Kc=0.859, Ti=7.062
Apply
ZN
Type: 1
Kc=NaN, Ti=NaN, Td=NaN
Apply
Simulation
End time: 80
Ts: 0.16
Noise Ts: 0.32
PVMin: 0
PV Max: 100
Noise 0
PV bias: 50
OP bias: 50
SP step time: 8
Step size: 1
Disturbance t0: 60
Amplitude: 0
(c) Krister Forsman; Ver 3.2, Feb 3, 2012

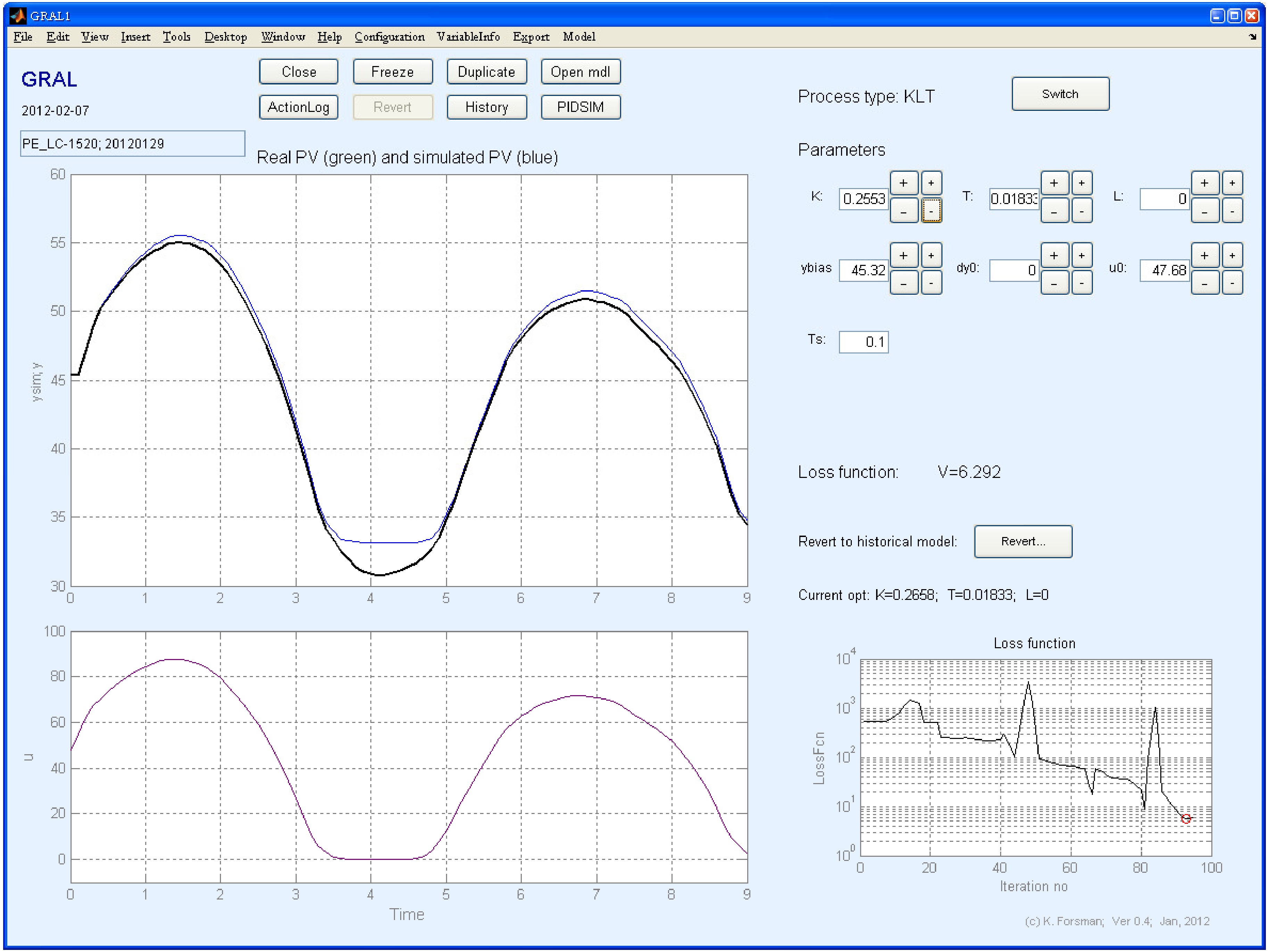





Perstorp
WINNING FORMULAS


Nyquist analysis GUI 1
File Edit View Insert Tools Desktop Window Help Import Process Export
Nyquist analysis: PID-controlled process
Close
Freeze
Save
Description
Duplicate
Revert
Load
Process: KLT
$\frac{Ke^{-sL}}{1+sT}$
K: -1
T: -10
L: 3
Controller:
Kc: 1.689
Ti: 57.88
Td: 3
Tdf: 4.284
Lambda tuning
λ= 15
Kc=0.5556, Ti=-10
Apply
MIGO
Ms= 1.2
Kc=0.859; Ti=7.06
Apply
Plot 1/Ms circle:
Ms: 2.3
Show frequency markers:
Stable
MsAct=2.3
Tpeak=15.1
Tcrit=234
Frequency range:
0.001
100
(c) Krister Forsman; Ver 1.2; Feb 2012
Imaginary part
Real part
2
1.5
1
0.5
0
-0.5
-1
-1.5
-2
-3
-2.5
-2
-1.5
-1
-0.5
0
0.5
1


Cascade control integrating process: Simulation
File Edit View Insert Tools Desktop Window Help Import Export Simulation
Cascade control
of integrating process
Close
Freeze
Duplicate
Revert
Open mdl
y2, r2
y1, r1
u1
Time
Process 2 (master)
Gain 0.100 Time const 0.5 Delay 0.5
Controller 2 (master)
Kc: 1 Ti: 40 Td: 0
Tdf: 5 β: 1 ARW: 5
Process 1 (slave)
Gain 1.000 Time const 4 Delay 1
Controller 1 (slave)
Kc: 0.67 Ti: 4 Td: 0
Tdf: 5 β: 1 ARW: 2
Simulation
Simulation length: 300
Sample time: 0.2
u1 bias: 50
u2 bias: 50
y2 bias: 20
DOLT: 95
y1 min: 0
y1 max: 100
y2 min: 0
y2 max: 100
u2min: 0
u2max: 100
r2 step time: 20
r2 step size: 10
r1 step t0: 60
r1 step ampl: 0
Disturbance 1 t0: 150
Dist 1 ampl: -5
Disturbance 2 t0: 200
Dist 2 ampl: 0
(c) Krister Forsman; Ver 1.0; Sep 22, 2009



GRAL1
File Edit View Insert Tools Desktop Window Help Configuration VariableInfo Export Model
GRAL
2012-02-07
PE_LC-1520; 20120129
Close
Freeze
Duplicate
Open mdl
ActionLog
Revert
History
PIDSIM
Real PV (green) and simulated PV (blue)
ysim, y
u
Time
Process type: KLT
Switch
Parameters
K: 0.2553
T: 0.01833
L: 0
ybias 45.32
dy0: 0
u0: 47.68
Ts: 0.1
Loss function: V=6.292
Revert to historical model:
Revert...
Current opt: K=0.2658; T=0.01833; L=0
Loss function
LossFcn
Iteration no
(c) K. Forsman; Ver 0.4; Jan, 2012

# Process model structures supported by our Matlab library

KLT $$\frac{Ke^{-sL}}{1+sT}$$

Integrating $$\frac{k_v e^{-sL}}{s(1+sT)}$$

IPZ $$\frac{k_v (1+sT_2)\, e^{-sL}}{s(1+sT_1)}$$

PPZ $$\frac{K(1+sT_2)\, e^{-sL}}{(1+sT_1)(1+sT_3)}$$

P3 $$\frac{Ke^{-sL}}{(1+sT)^3}$$