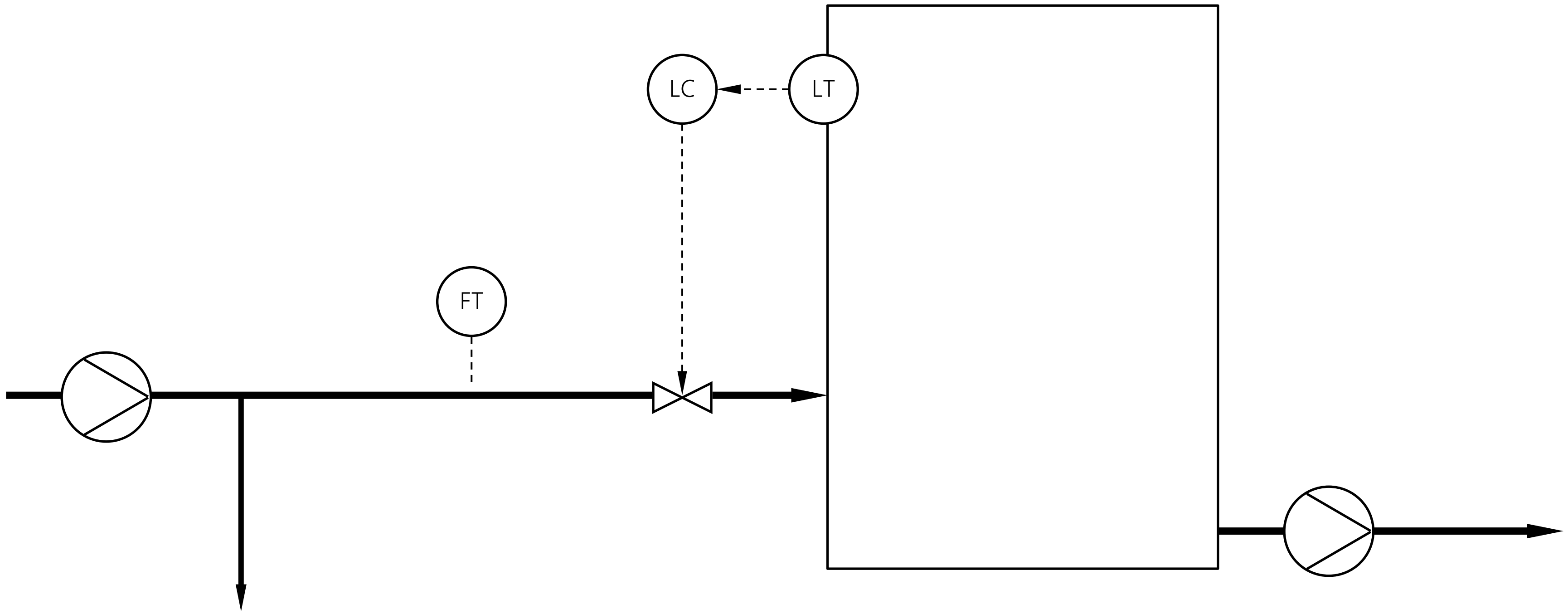
Slides Krister Forsman

DYCOPS workshop
16 June 205

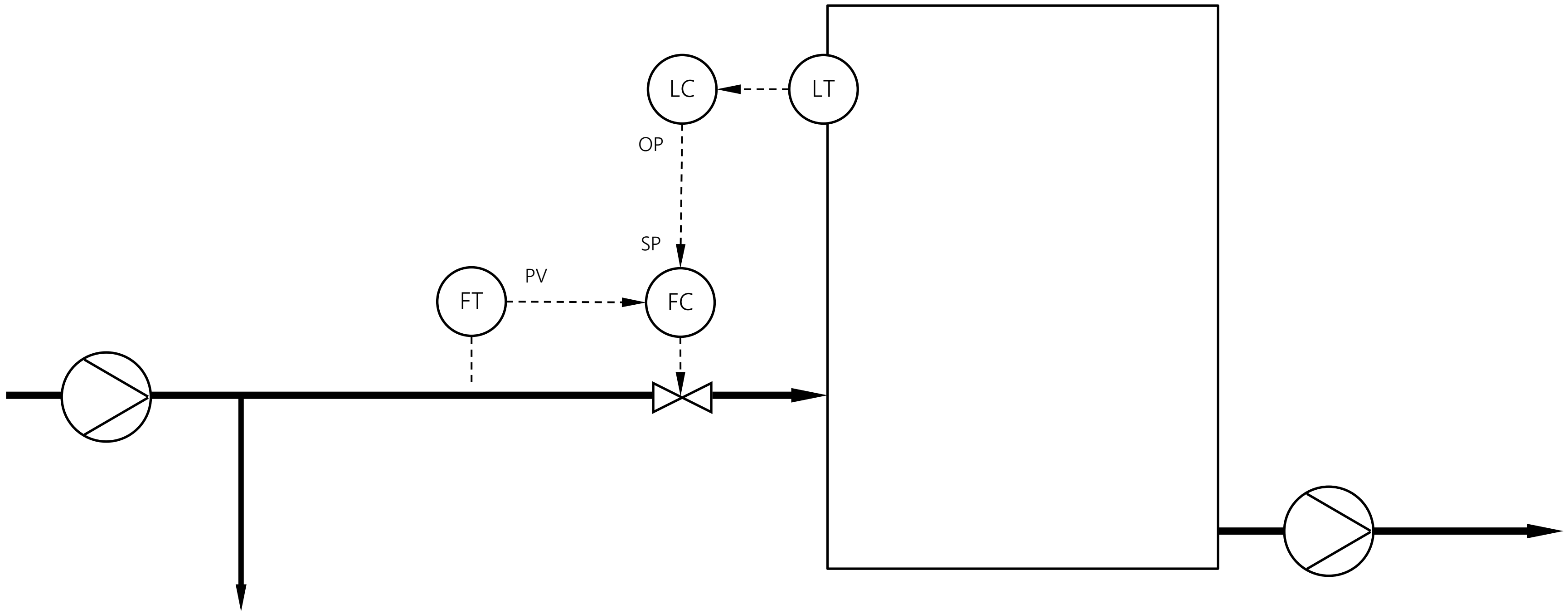
Cascade control

Example: Level control with improvement potential

- It is important to keep this level close to its setpoint.
- The tank level varies too much, due to large pressure variations in the feed line.
- We can’t make the controller more aggressive, because then it becomes unstable.
- Can we still improve control performance?

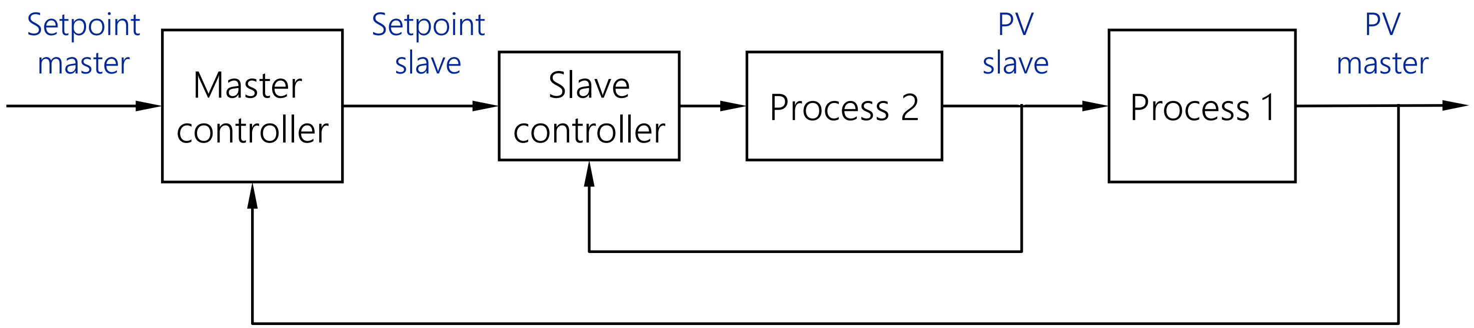


Solution: Control the flow too!



Cascade control: One controller provides the SP to another

- The control signal of the master controller (u_1) is the setpoint of the slave controller (r_2).

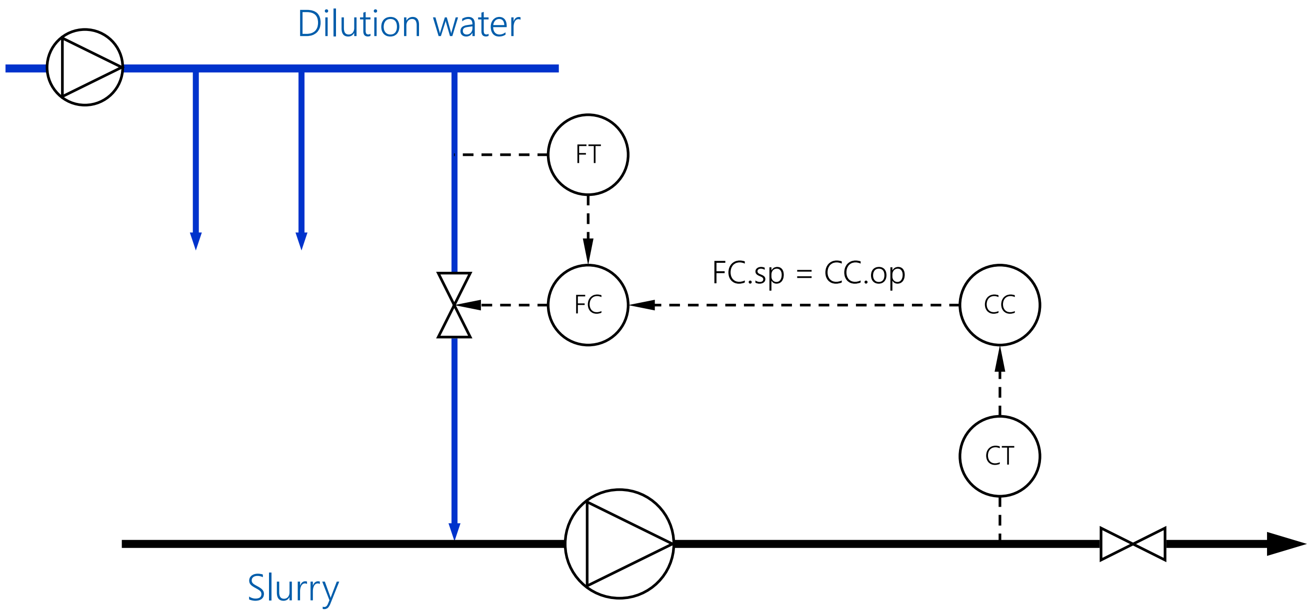


Typical examples for cascade control

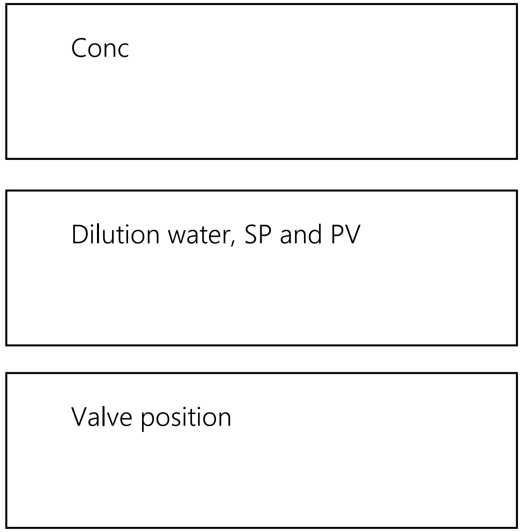
Master	Slave
Level	Incoming or outgoing flow
Temperature HEX	Cooling water flow
Concentration	Dilution water flow
Column temperature	Reflux flow
Evaporator density	Steam flow
pH	Caustic addition

Cascade control: Slave disturbance "paradox"

- Scenario: Concentration control (master) with dilution water as slave
  - A decrease in water header pressure causes the flow of dilution water to decrease.
  - Outline how dilution water flow and concentration vary over time.
    The master and slave controller are both active.
- Hint: Start by drawing in the middle graph



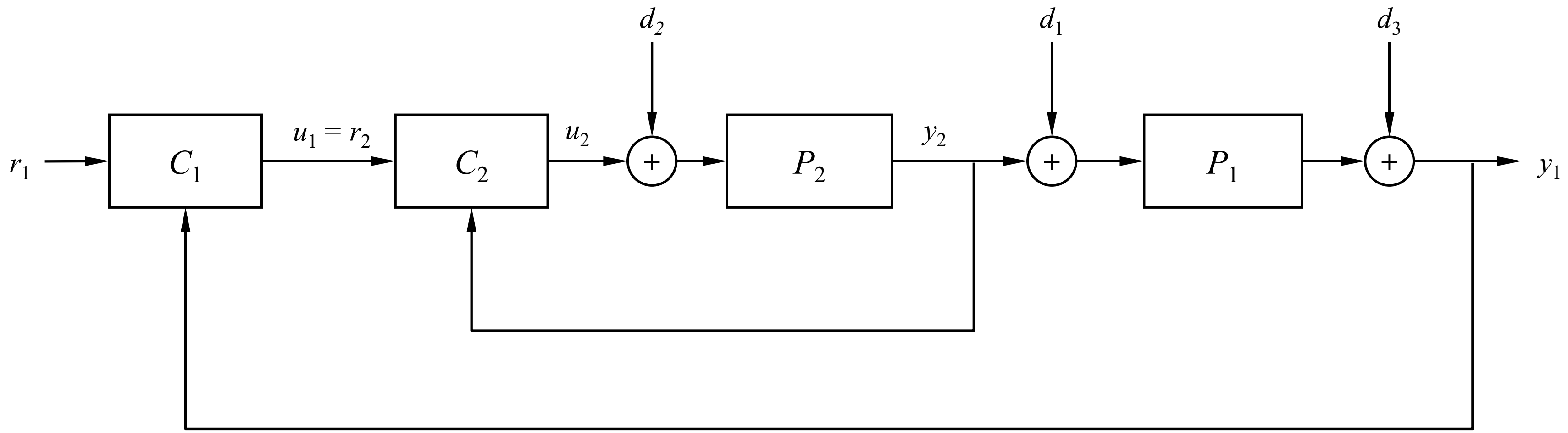
Give the answer as 4 trends



Question: Where does the disturbance enter?

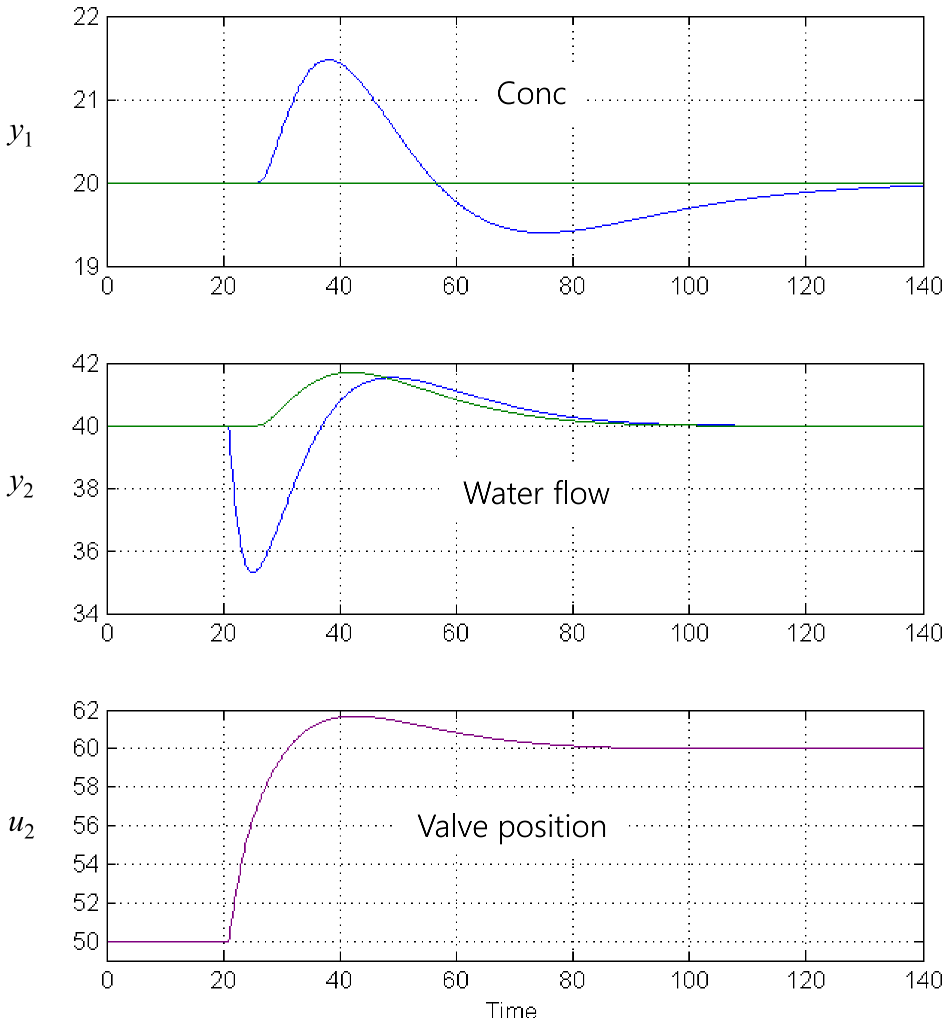
In the above scenario: Which is the best model for pressure disturbances?

Do they enter as d_1, d_2 or d_3 in the block diagram below?



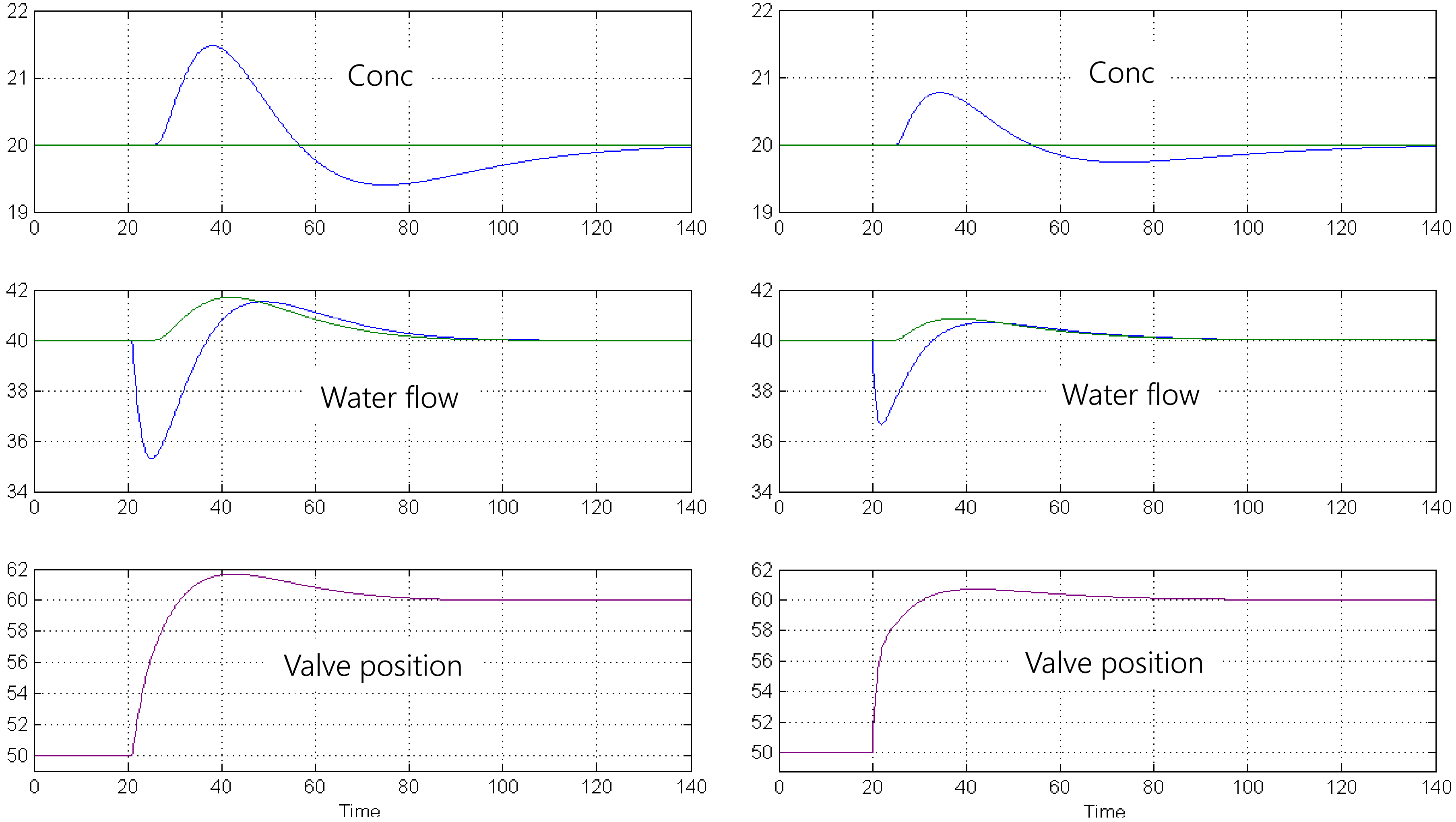
Answer: d_2

Solution: S-shaped disturbance in concentration



- The decrease in water flow affects concentration, even though the flow controller is active.
- The master controller acts on this, changing the water flow SP.
- In some sense this is unnecessary. This disturbance should be handled by the slave controller.

Conclusion: Important that the slave is much faster than the master

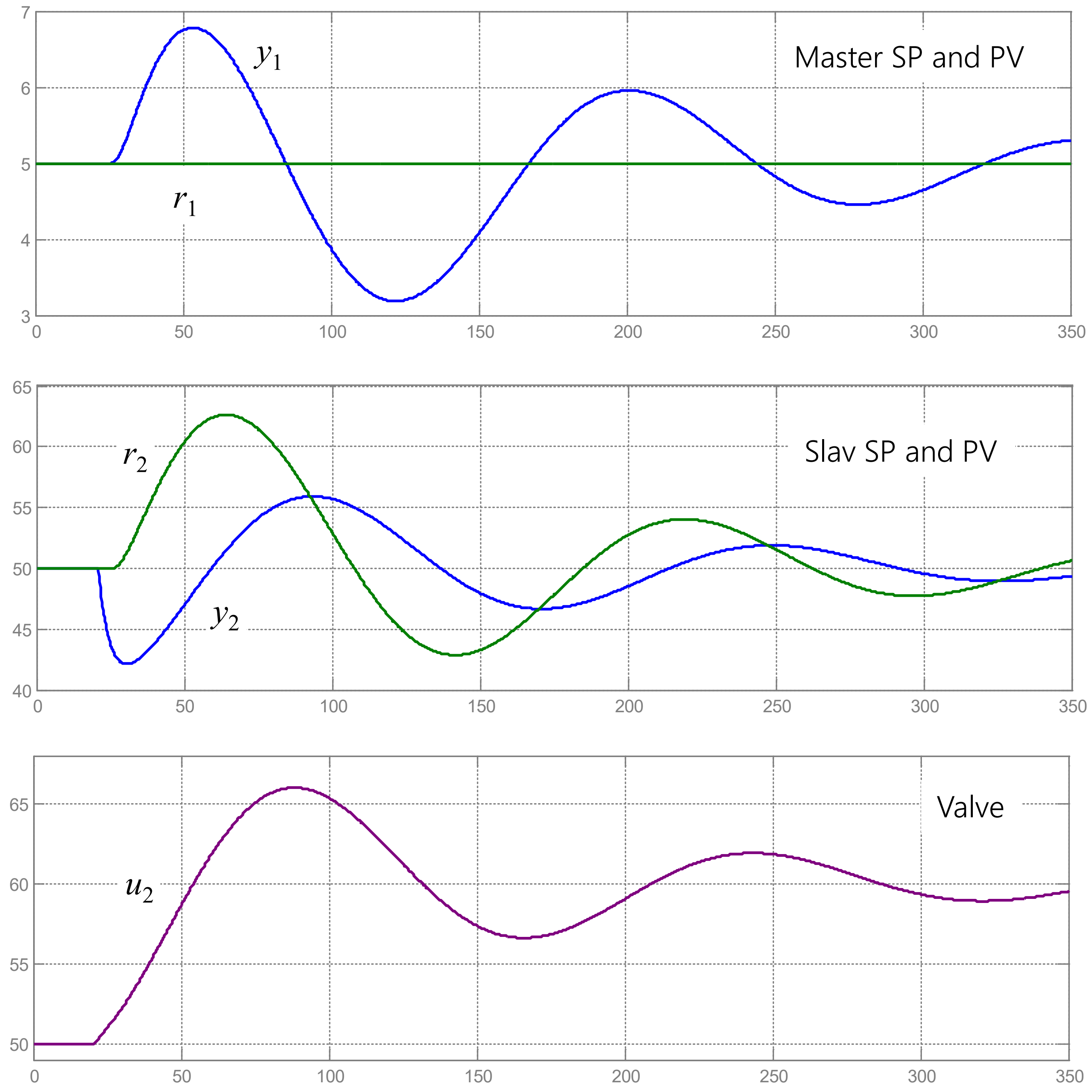


Here somebody tried to make the master faster than the slave.

The disturbance is thrown back and forth between master and slave

On the borderline to unstable.

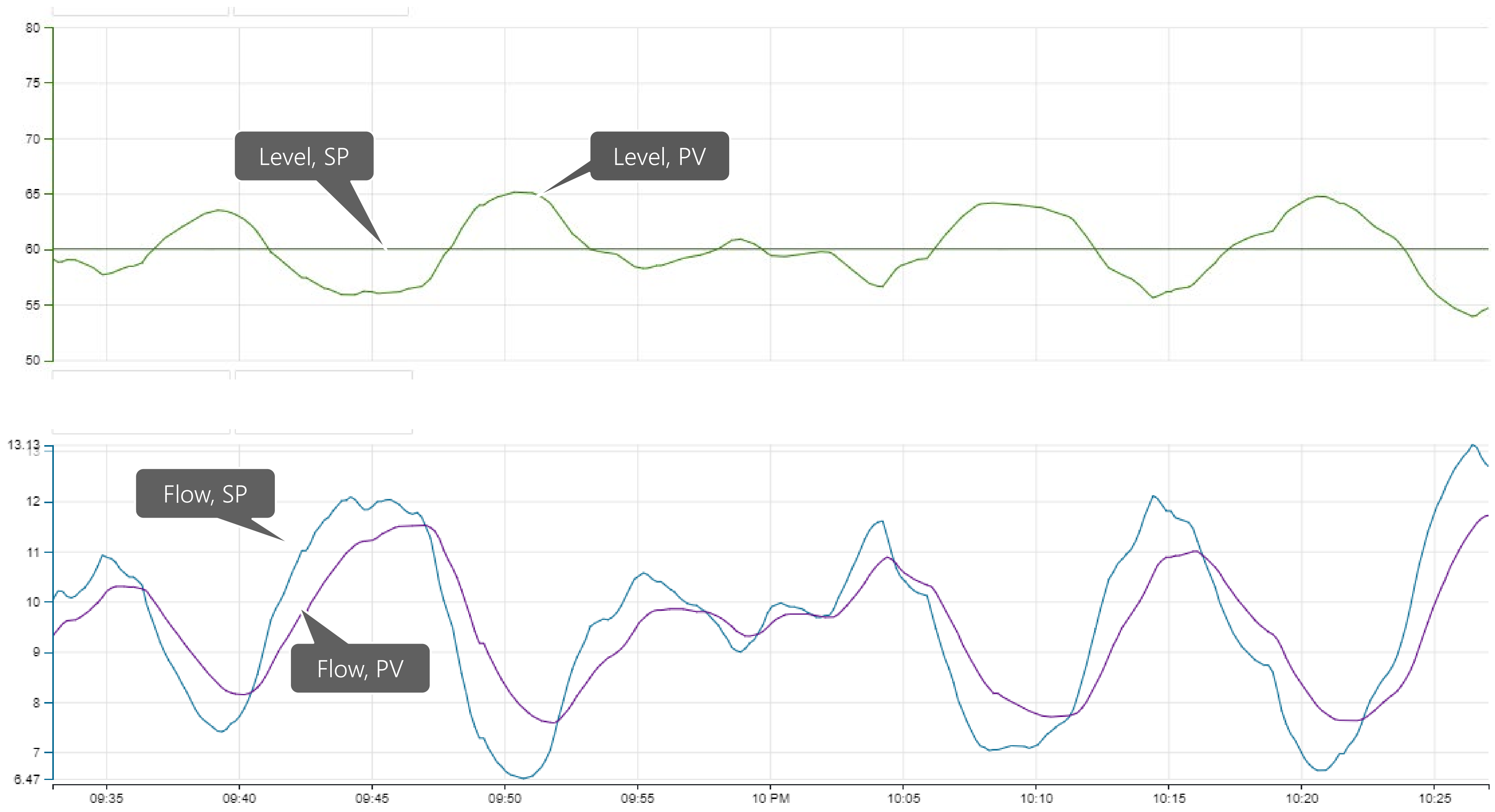
" λ_1=15 λ_2=40 "



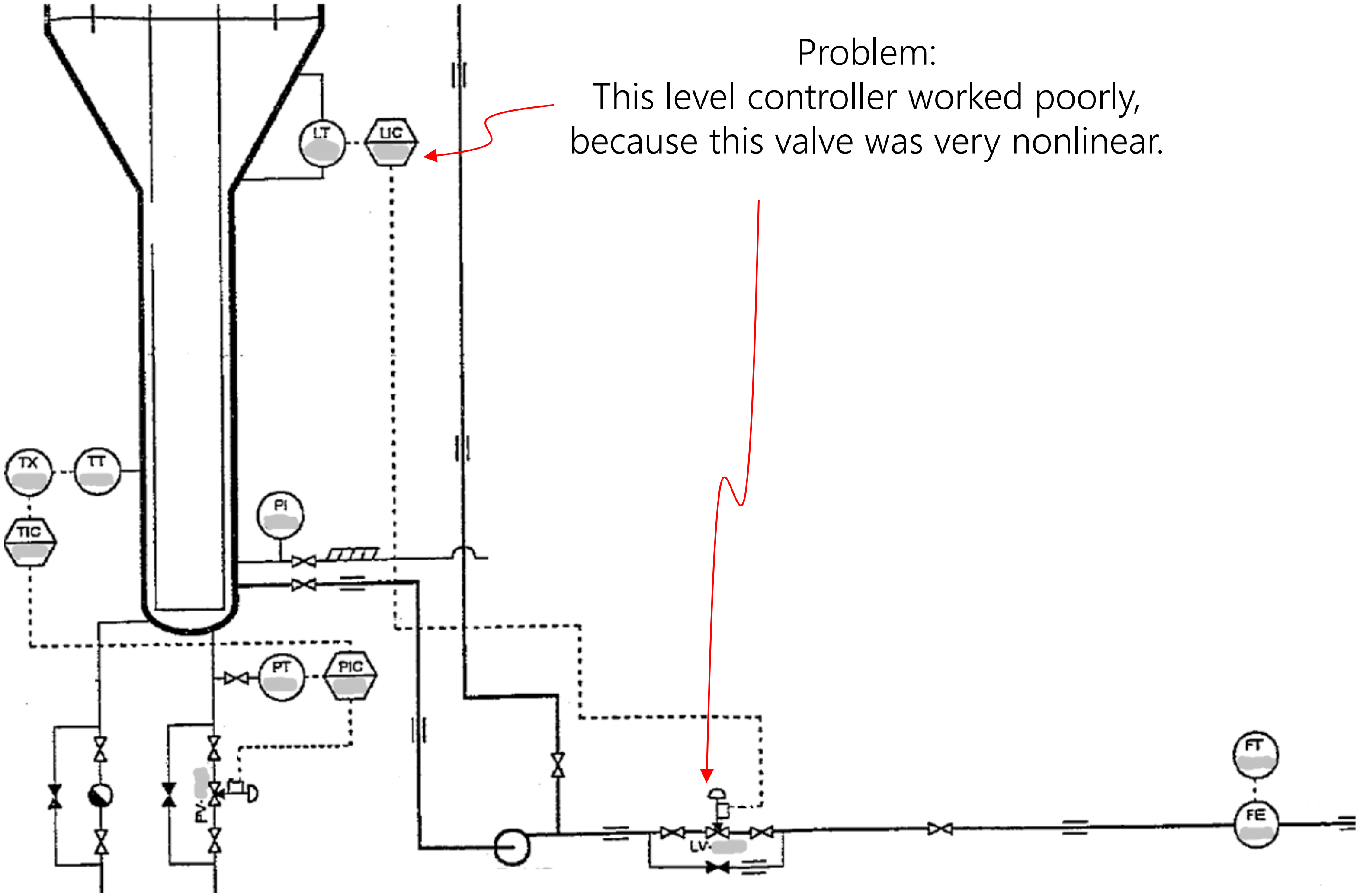
When to use cascade control?

- The slave loop dynamics should be considerably faster than the master loop dynamics.
- The motivation for cascade control is in the slave controller.
  It should handle disturbances entering before process 2.
- Can cascade control be motivated even if you don’t have disturbances at the lowest level?
  - The answer is ”yes”, and we will soon see why.

Real example of poor frequency separation

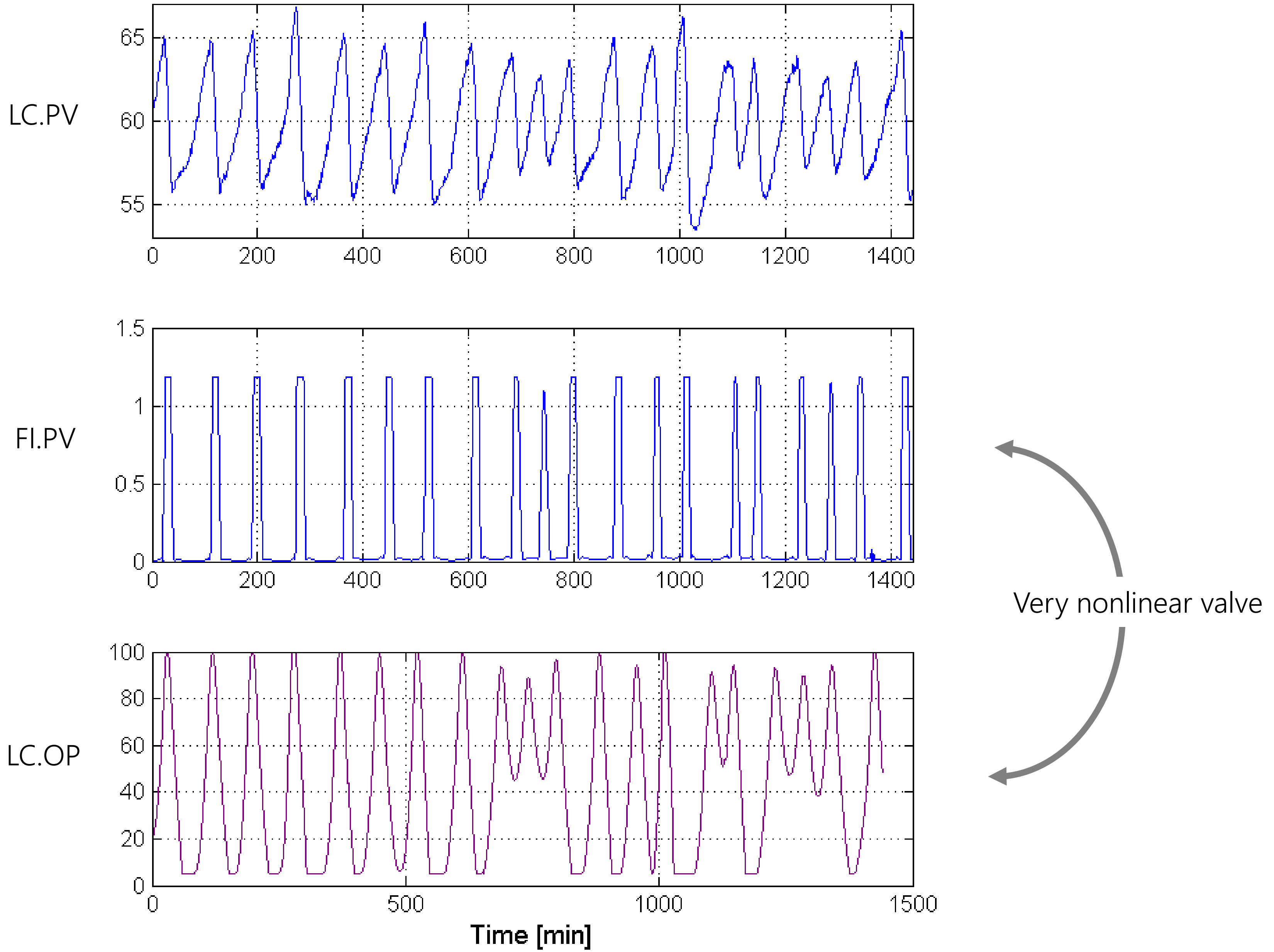


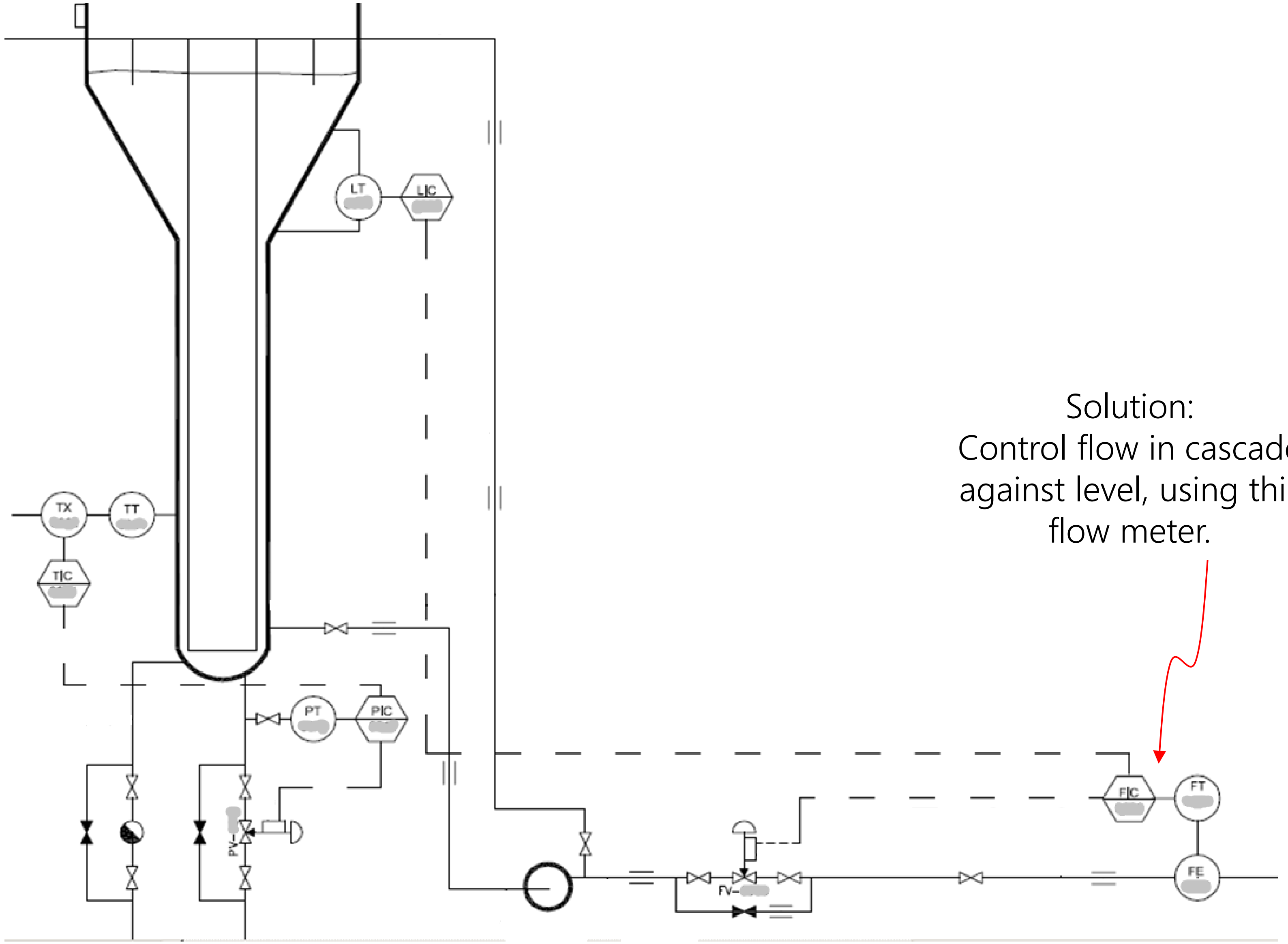
Evaporator: Improvement opportunity



Problem:
This level controller worked poorly,
because this valve was very nonlinear.

Before improvement:

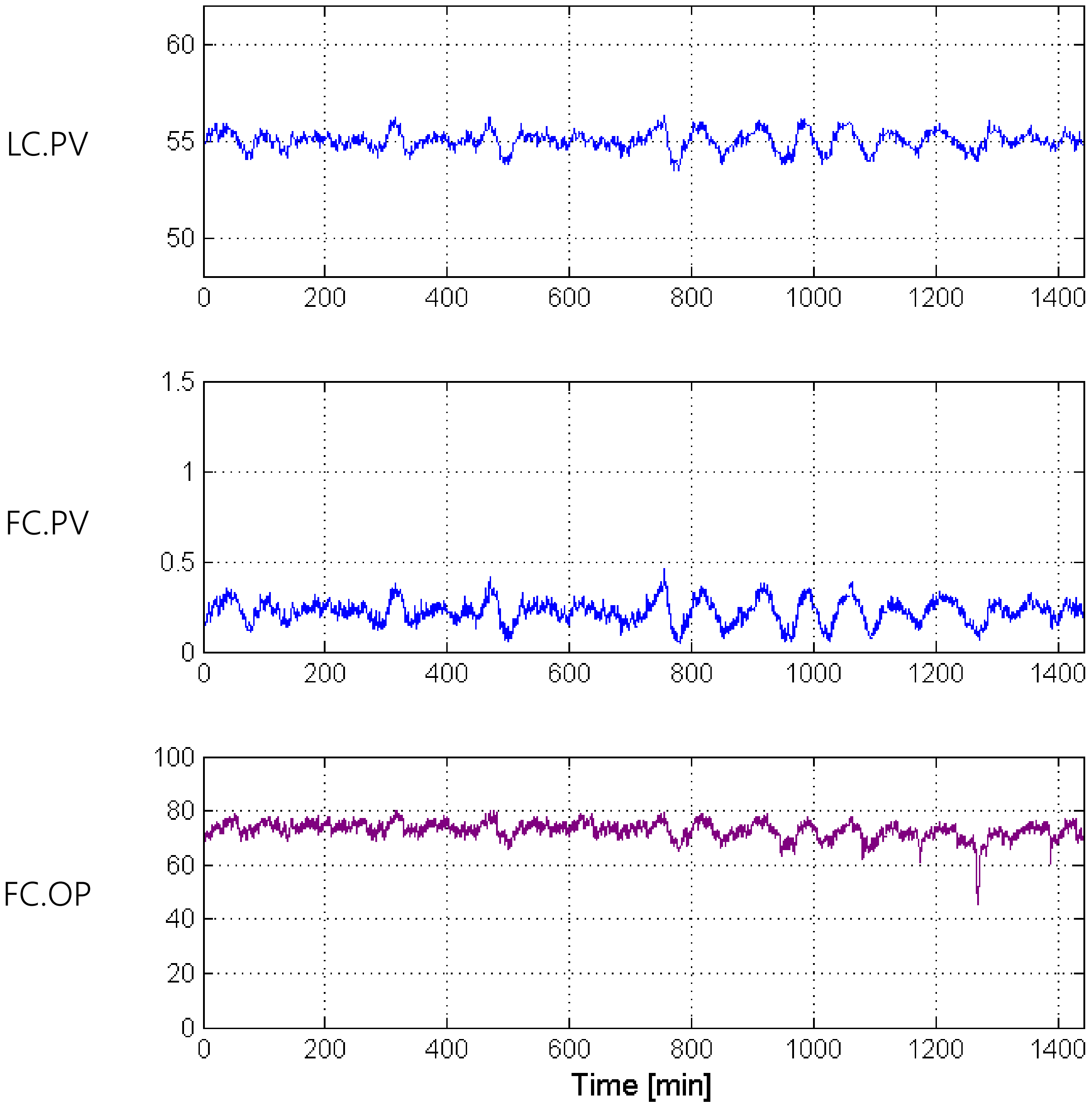




Solution:
Control flow in cascade against level, using this flow meter.

Cascade control eliminates the effect of the valve nonlinearity.

No more level cycling.



Cascade control linearizes

- Why did the slave process become linear in the example above?

- The slave process is a control loop. From SP to PV.
- In a closed loop the gain (from SP to PV) is always 1.
  - In steady state SP = PV, if the controller has integral action.

Additional advantage:

- In some applications it's convenient for the operator to be able to run the flow in Auto, with setpoint, instead of manipulating the control valve.

Disadvantage

- If the flow sensor malfunctions, it is not possible to control the level any more.
  The operator cannot "switch back to the old ways".

Policy: Motivate all additional complexity

General principle:

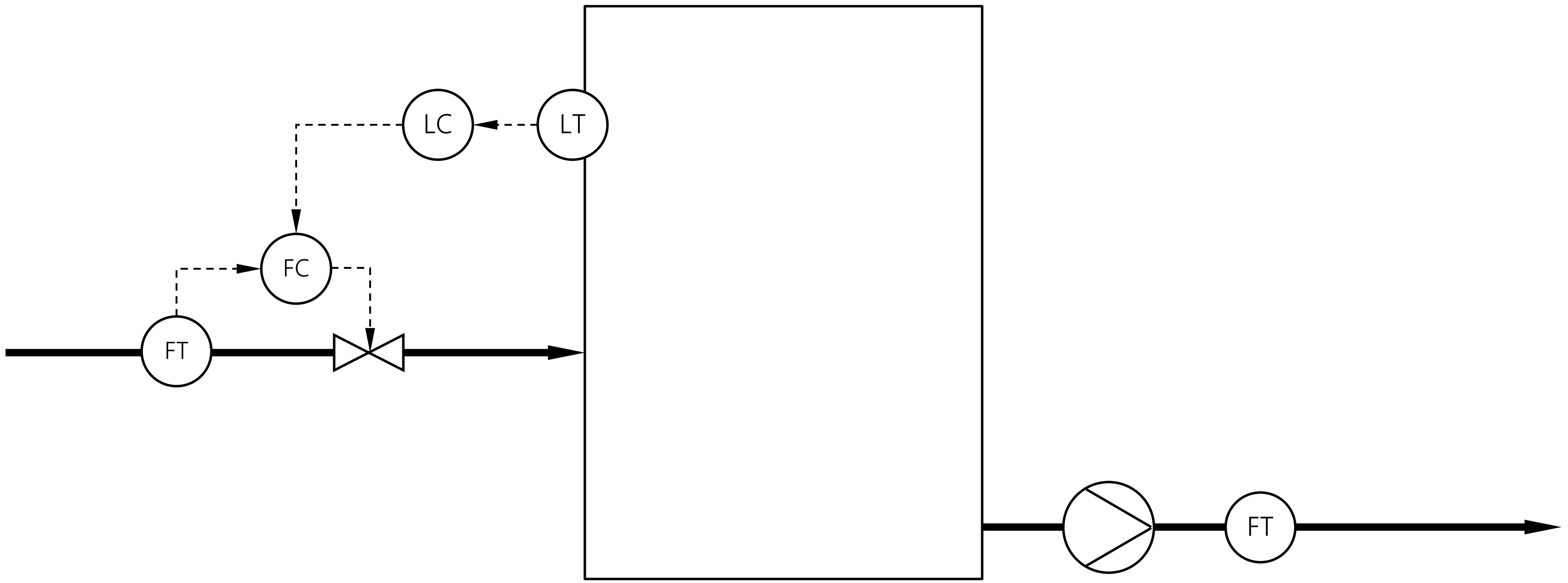
- If you are using something else than a simple feedback you should motivate why SISO is not enough.

- In the first case above, pressure disturbances were that motivation.

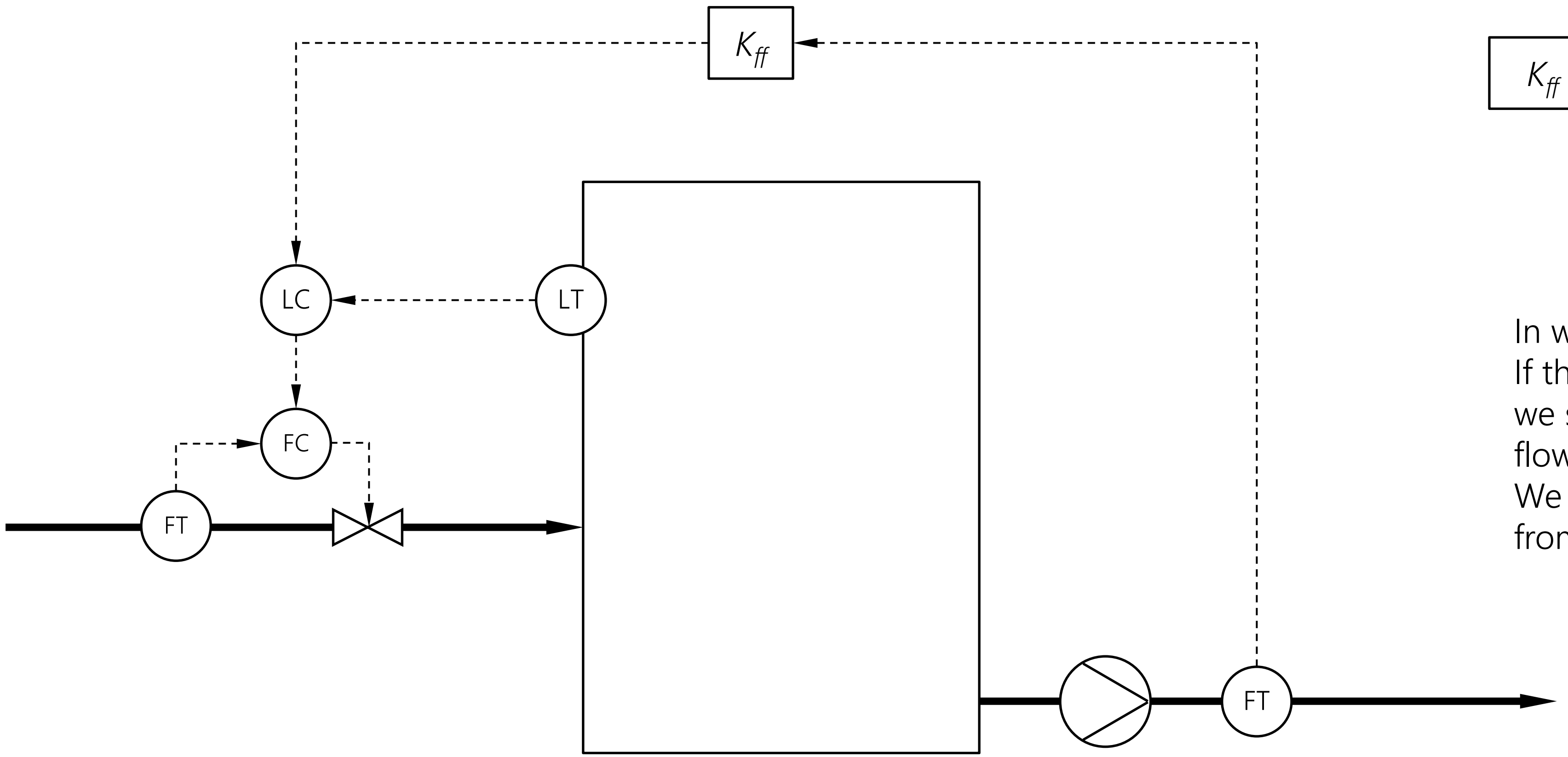
Feedforward

Ex: Level control with improvement potential

- It is important to keep this level close to its setpoint.
- The tank level varies too much, because of large variations in the outgoing flow.
- The LC cannot be tuned more aggresively: then it becomes unstable.
- Can we still improve control performance?



Feedforward: Warn the controller in advance



K_{ff} = multiplication with a factor K_{ff}

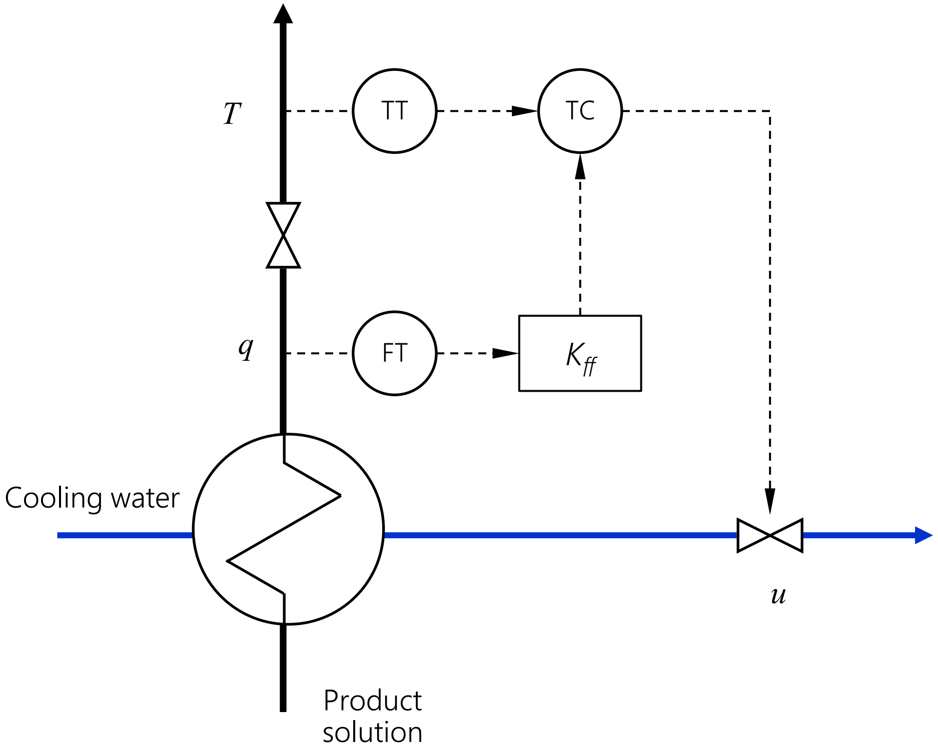
In this case probably K_{ff} =1

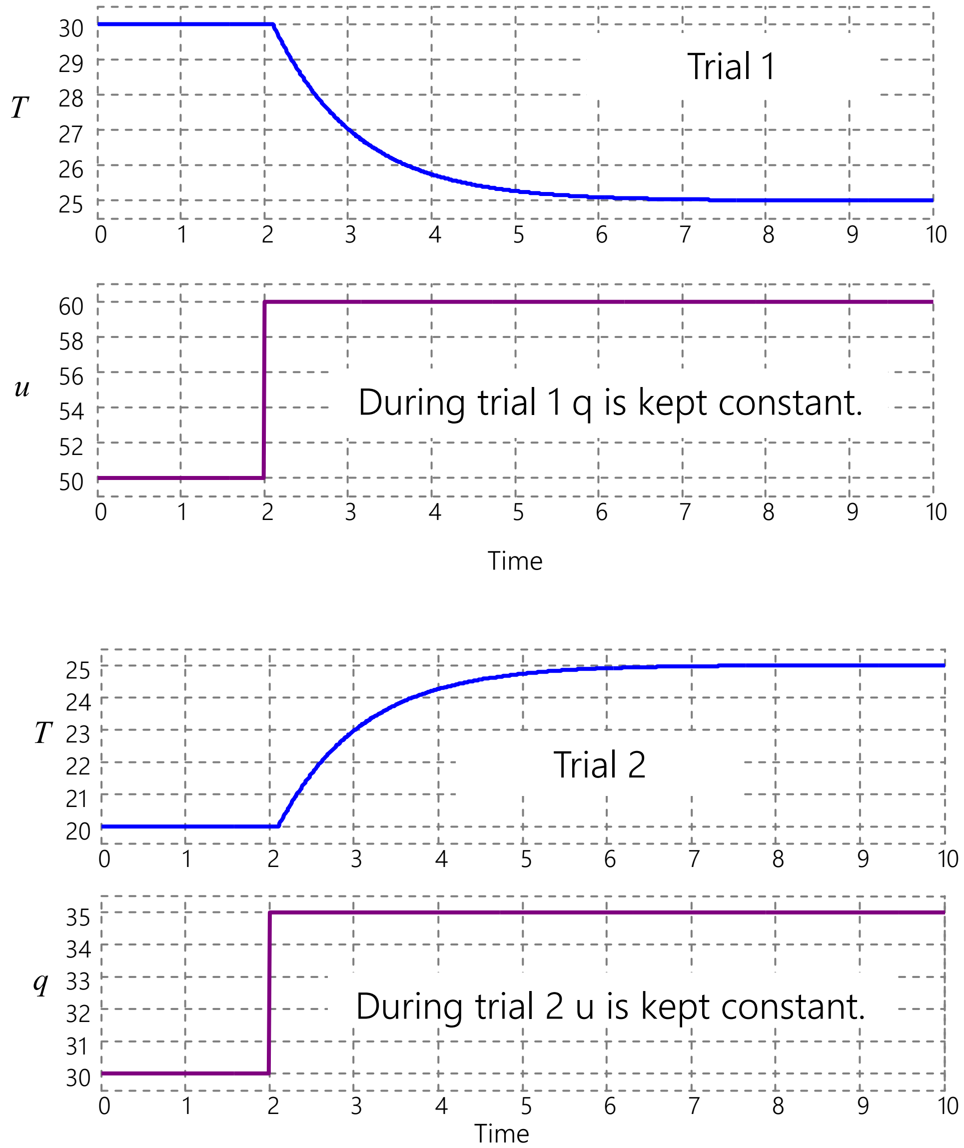
In words:
If the outgoing flow increases by 5 m^3/h
we should immediately increase the incoming flow by the same amount.
We don't need to wait for the level to deviate from its setpoint.

Exercise: FF gain for HEX

Determine the correct feedforward gain.

Use data from the two trials showed to the right.

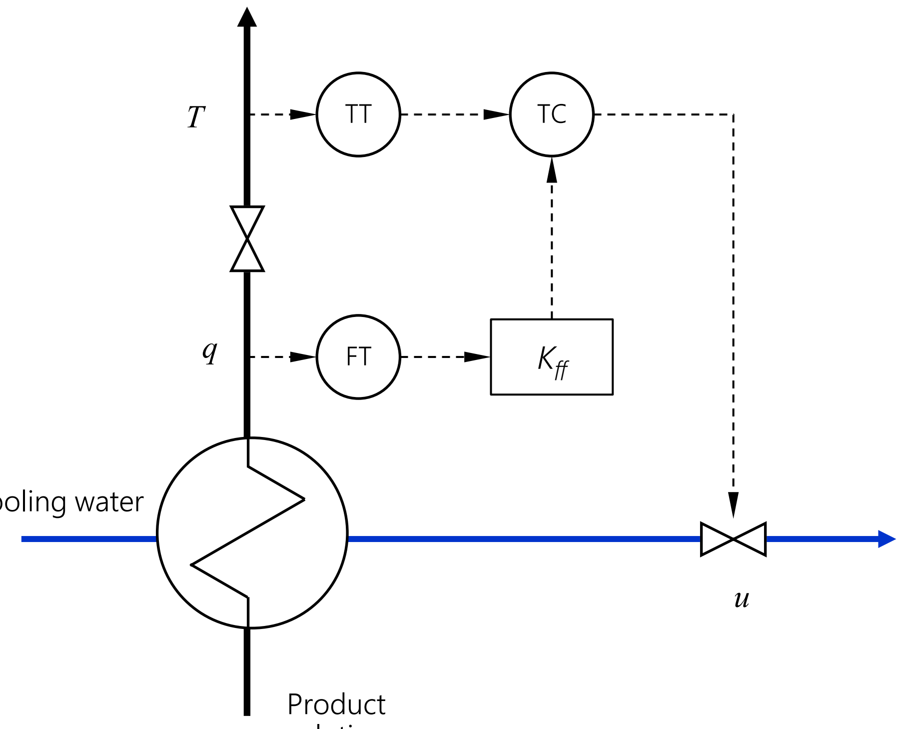


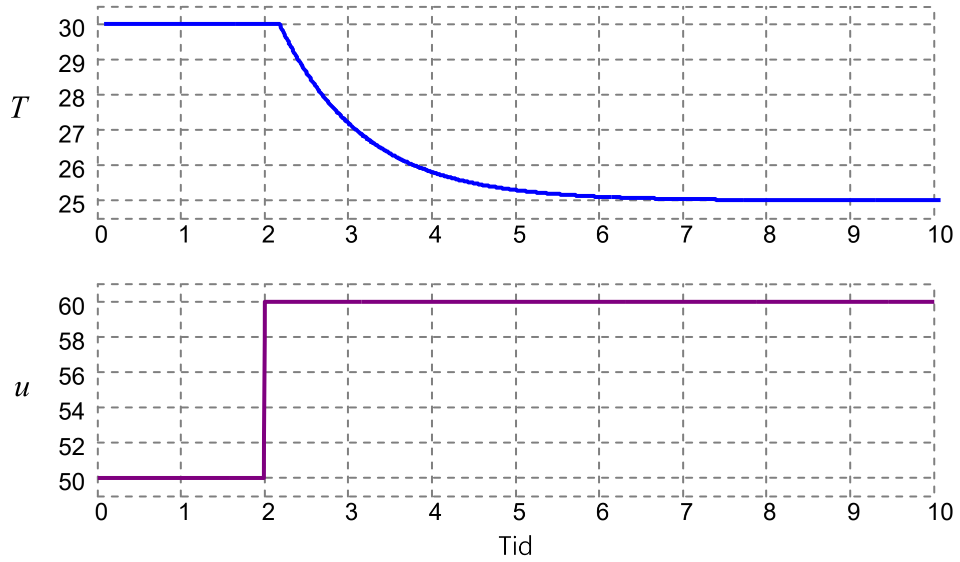


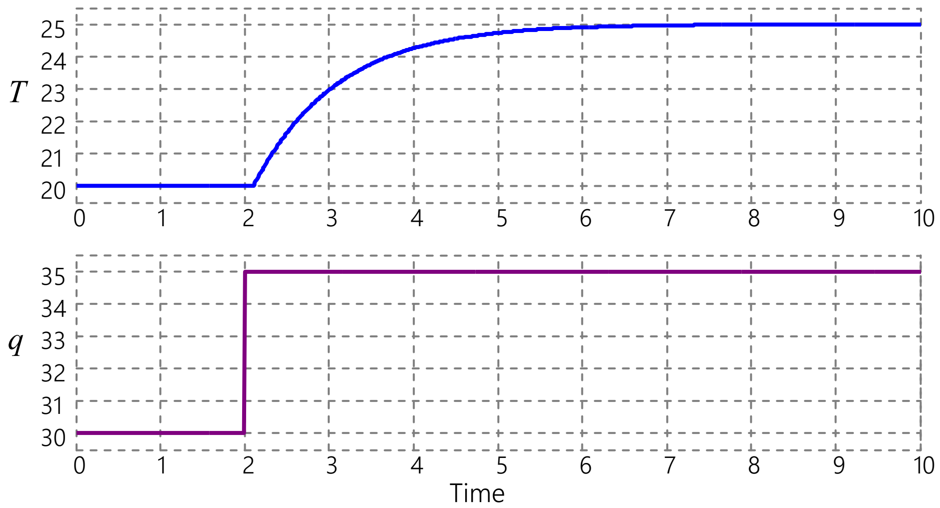
Solution

The effect of q on temperature is "twice as strong" as the effect of u:
5 units in q has the same effect as 10 units on u, but in the opposite direction.

Thus: To counteract a disturbance from q, u has to change twice as much as q.
So: $K_{ff} = 2$



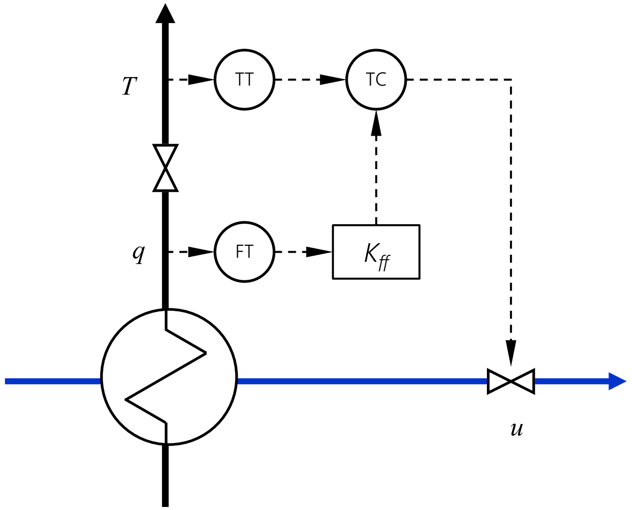


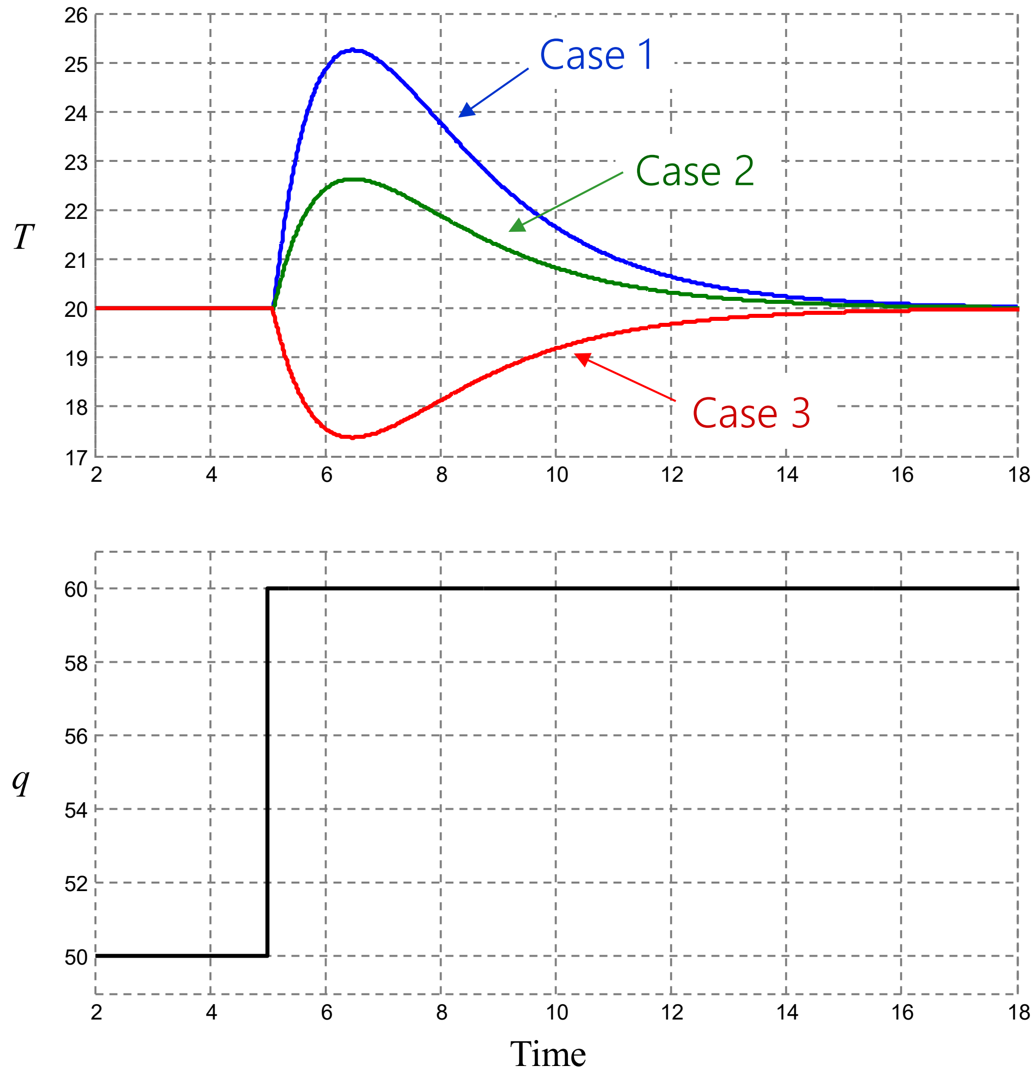


Exercise: Wrong K_{ff} for HEX in the example above

Which behavior corresponds to

$K_{ff} = 0$	Case 1
$K_{ff} = 1$	Case 2
$K_{ff} = 3$	Case 3





Conclusion: Feedforward is not robust

- Feedback control: If the process model used for tuning is not correct, the feedback algorithm compensates for that.

- Feedforward control: A model error can cause more damage:
  - If the FF gain is completely wrong, the FF can do more harm than good.

Use a ”chicken factor”

Feedforward exercise

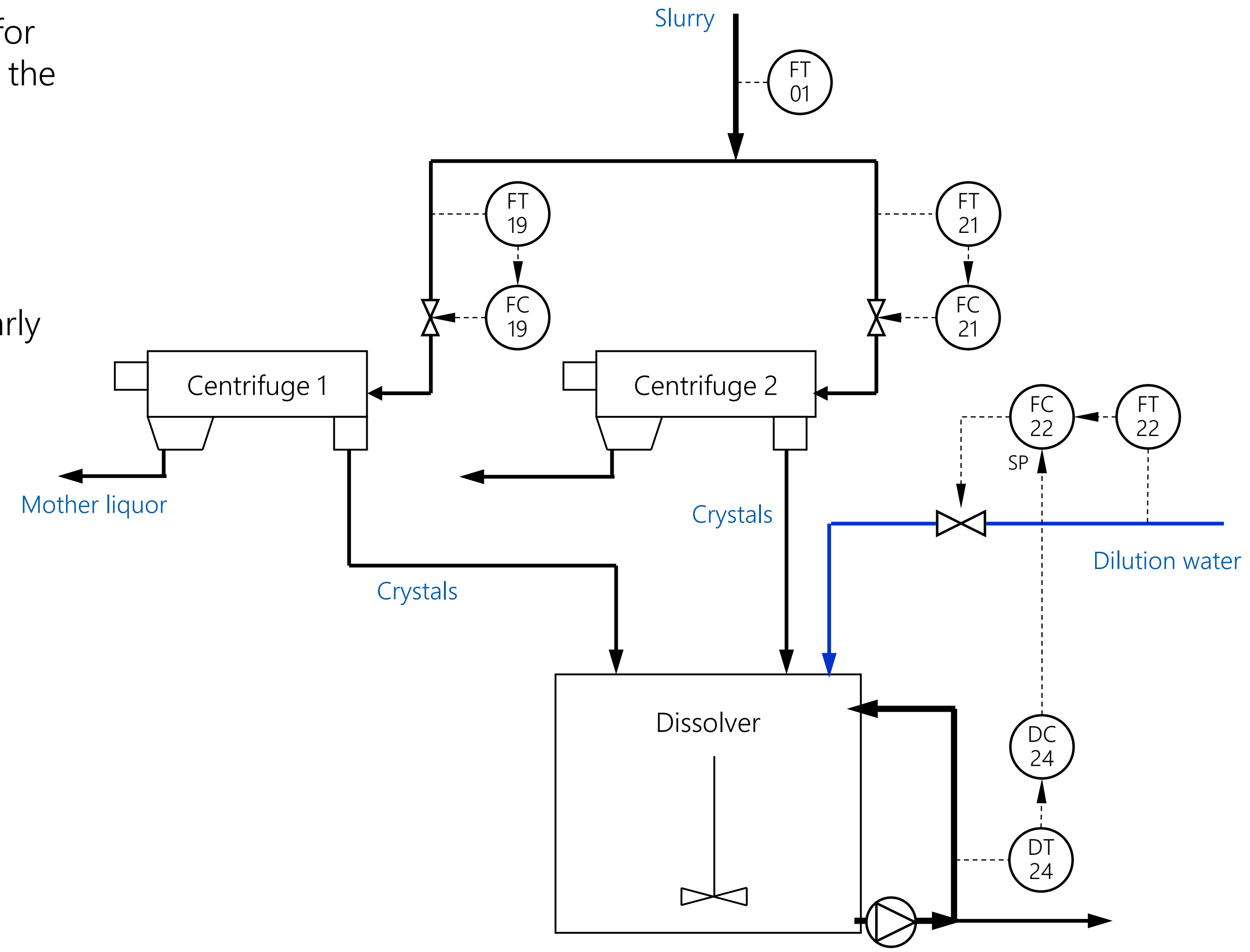
Density control in a dissolver

Problem: Every shift one of the centrifuges is shut off for cleaning, during one hour. Then the flow of crystals to the dissolver decreases by 50%
(but the solids flow is not measured).

Consequence: a large upset in dissolver density.

The density controller DC24 cannot be tuned particularly aggressively because the process dynamics is slow.

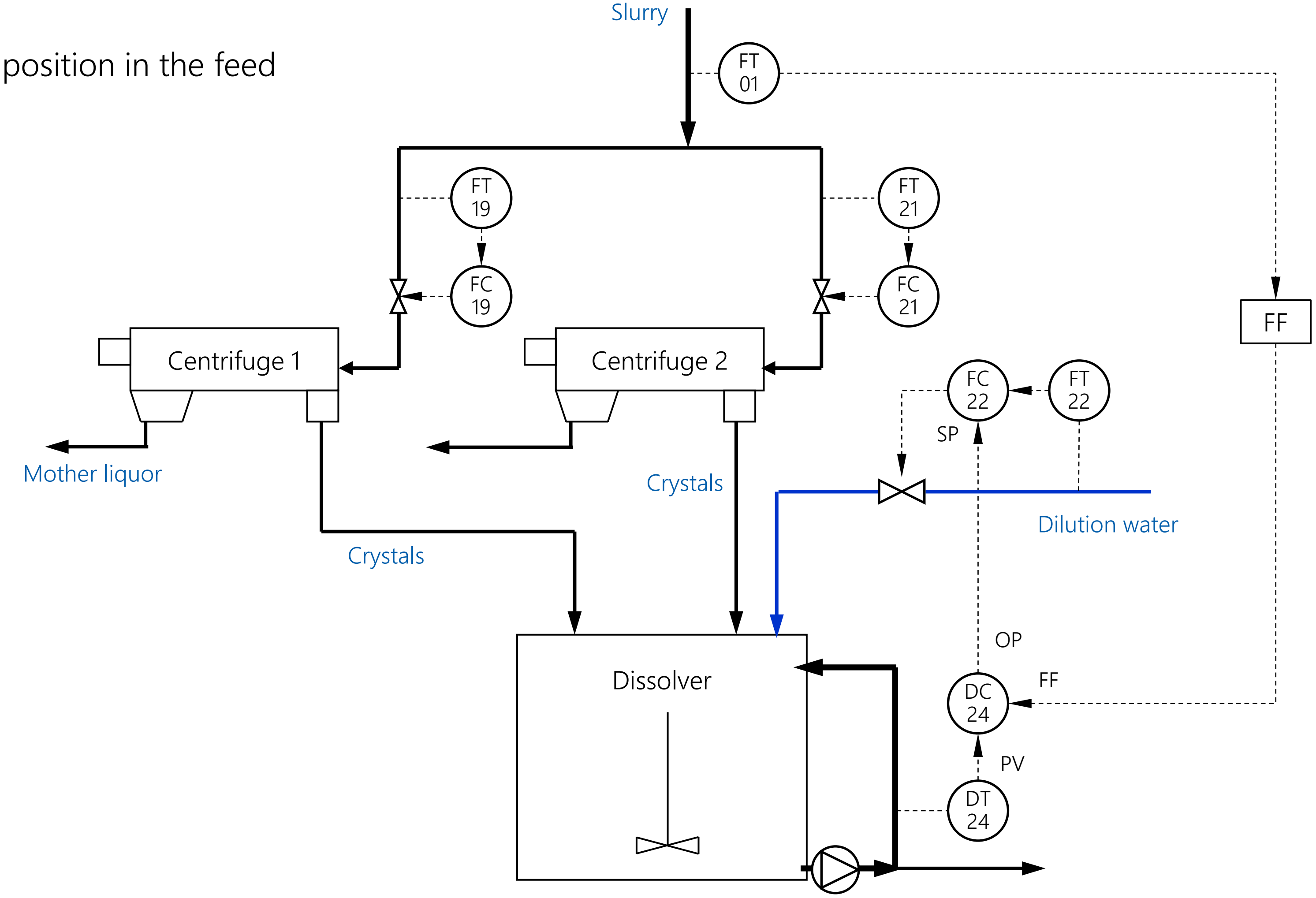
How can we improve density control?



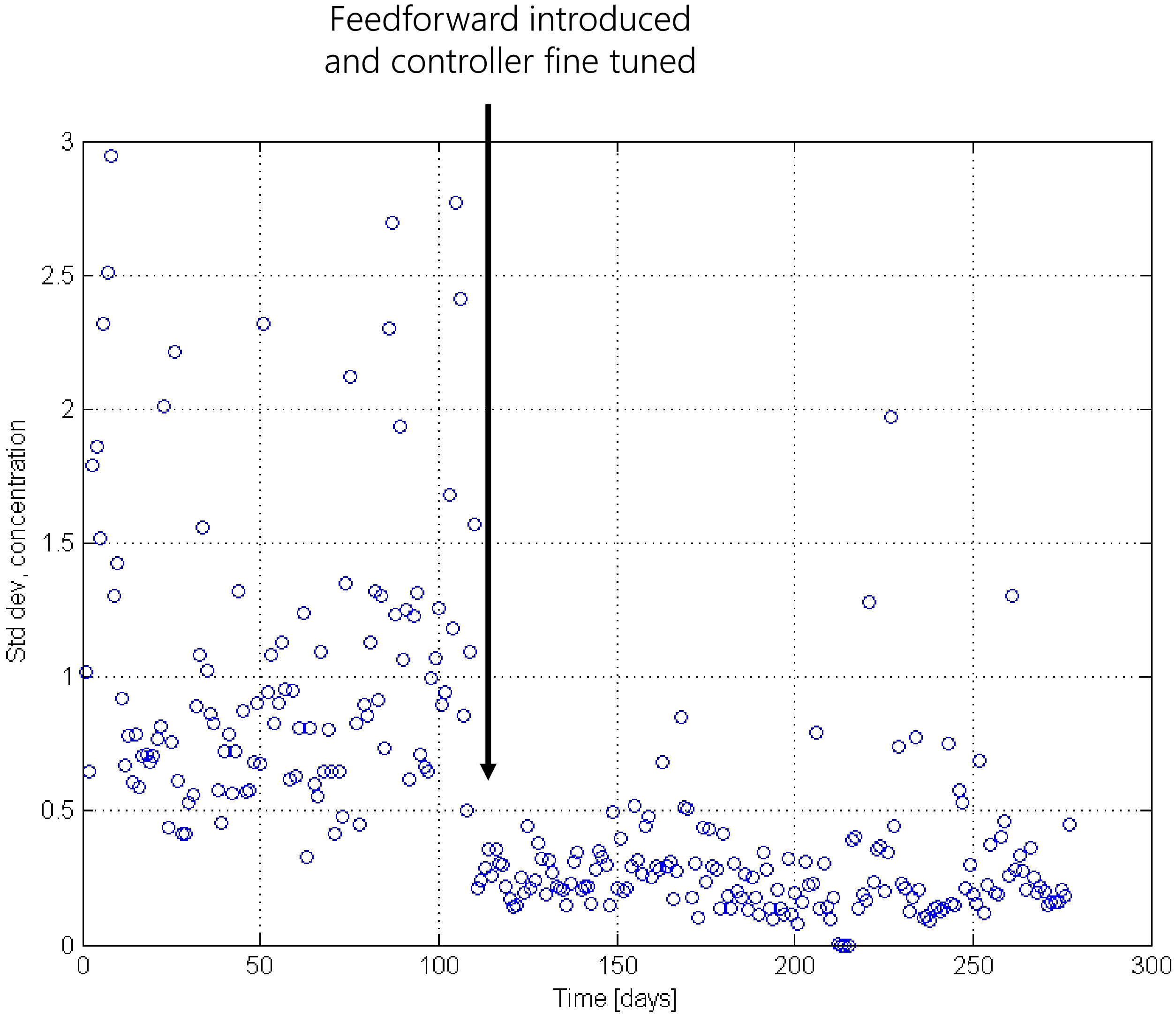
Solution

Introduce a feedforward from centrifuges total feed to the density controller.
A change in feed corresponds to a change in solids flow to the dissolver.

The feedforward gain will not be perfect, since the composition in the feed slurry varies. But in practice it works very well.

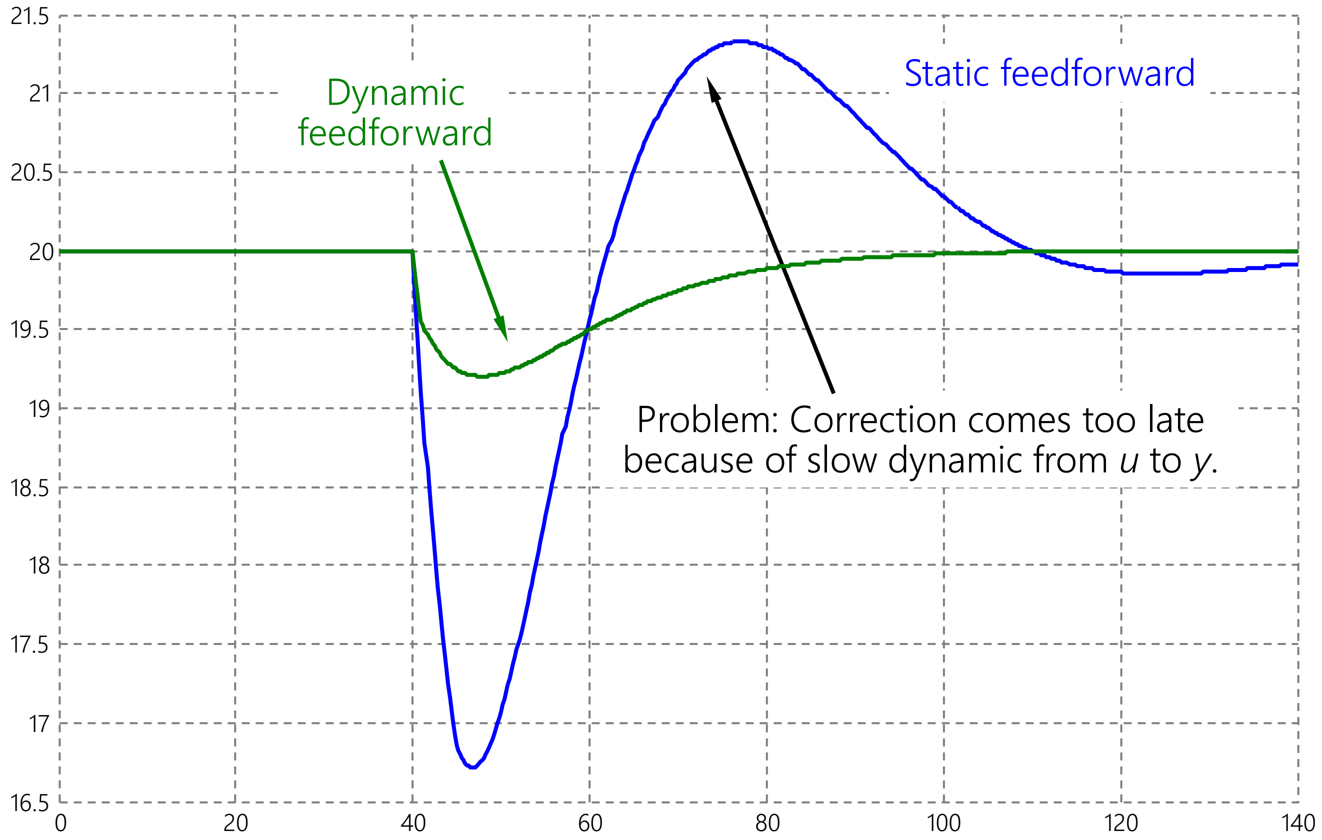


Feedforward and tuning reduced variations



Careful if the two dynamics differ

- In some cases we have control loop where the disturbance effect on PV has a different dynamics than the effect from controller output.
- Then it may be a bad idea to introduce a static feedforward. Why?



Avoid unintentional feedback!

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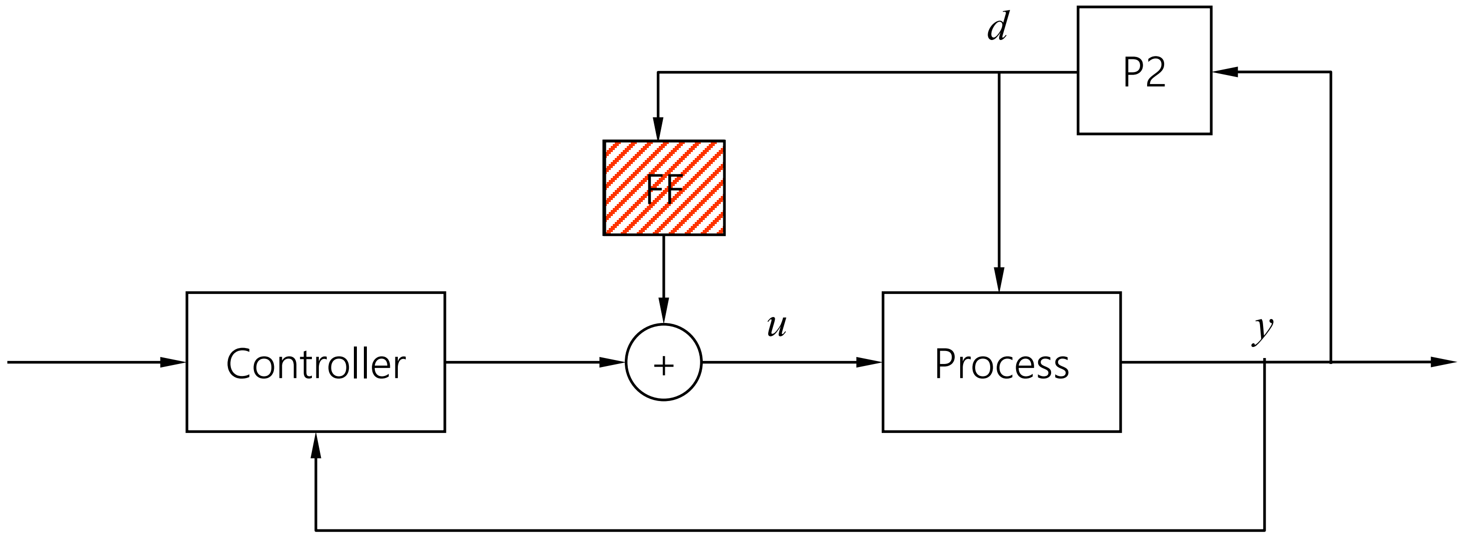
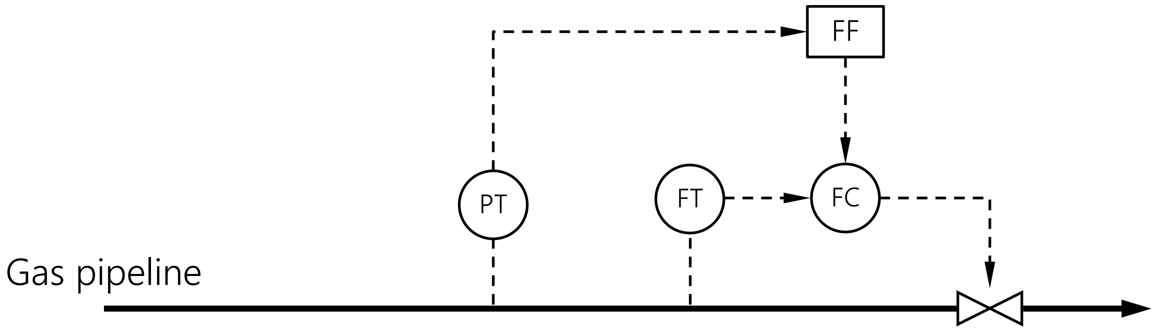
- If the disturbance variable depends on the manipulated variable you have to be careful.
  - If you try to feed forward you introduce an extra feedback!
- *d* is not an externel variable
- It may work, but the previous analyses do not apply.
- Normally a feedforward cannot make the loop unstable, but in this case it may.



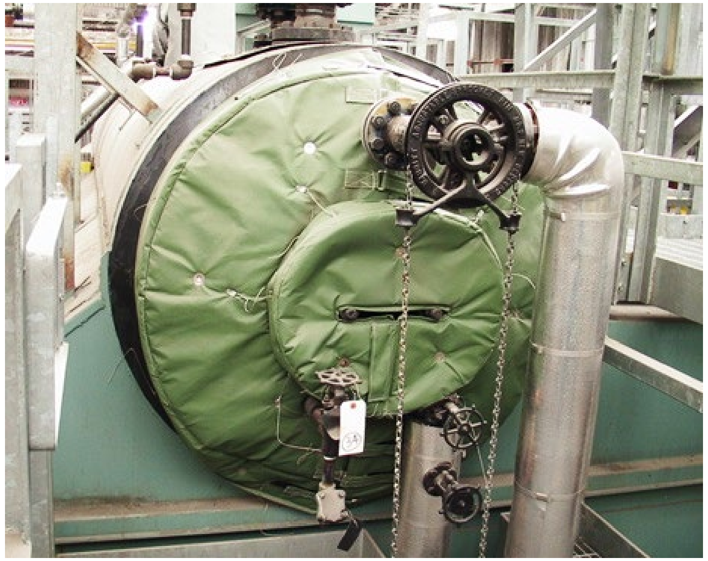
Example of unintentional feedback

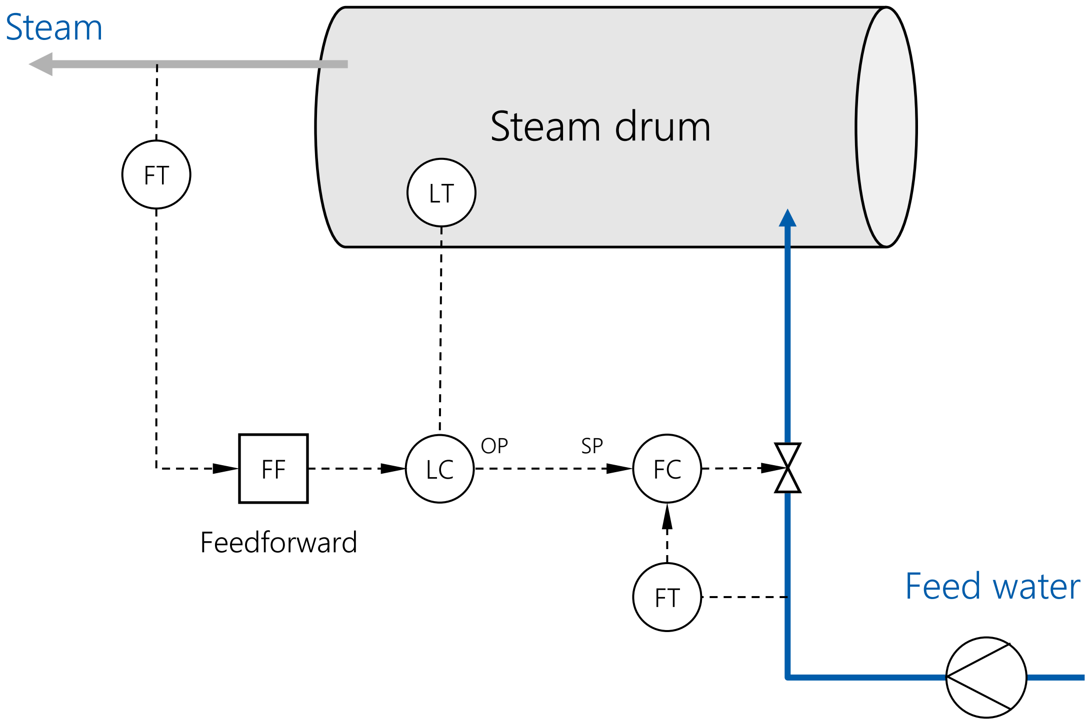


If this is a large consumer, the valve affects both flow and pressure.
Then the FF-block is not a normal feedforward, but an extra feedback.

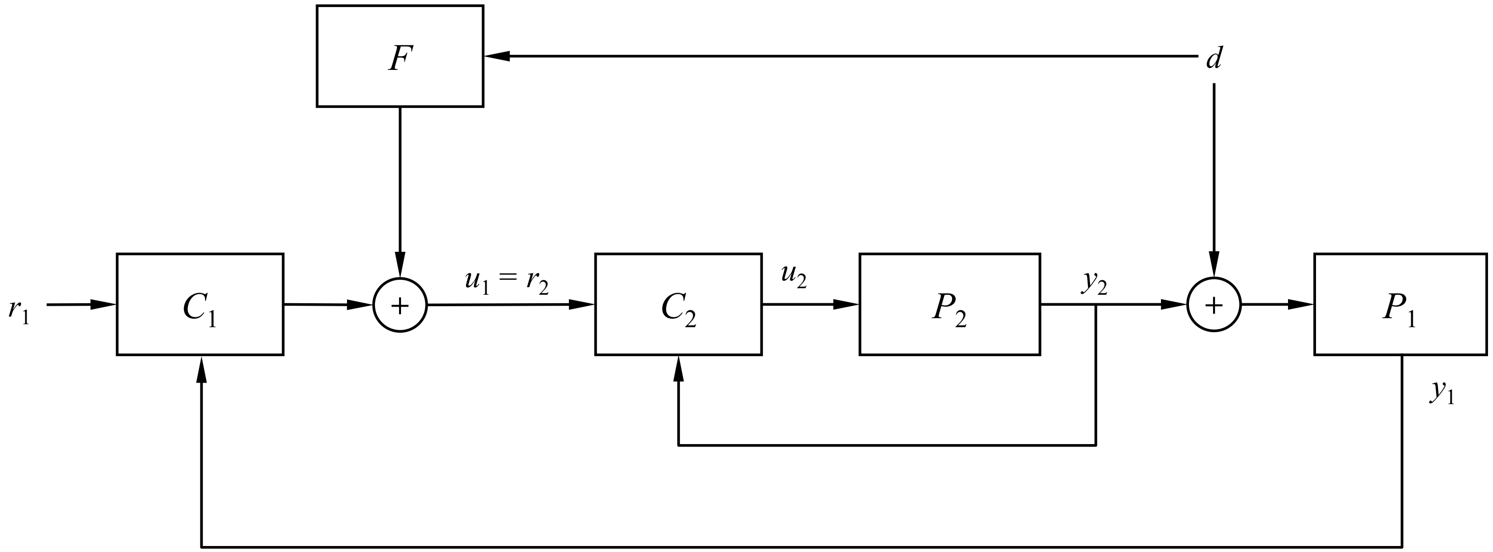
Drum level example: FF + cascade

- A common way of controlling the drum level in a steam boiler is to combine feedforward and cascade control.
- The LC gets a feedforward from steam consumption. Feedwater flow in cascade agains level.





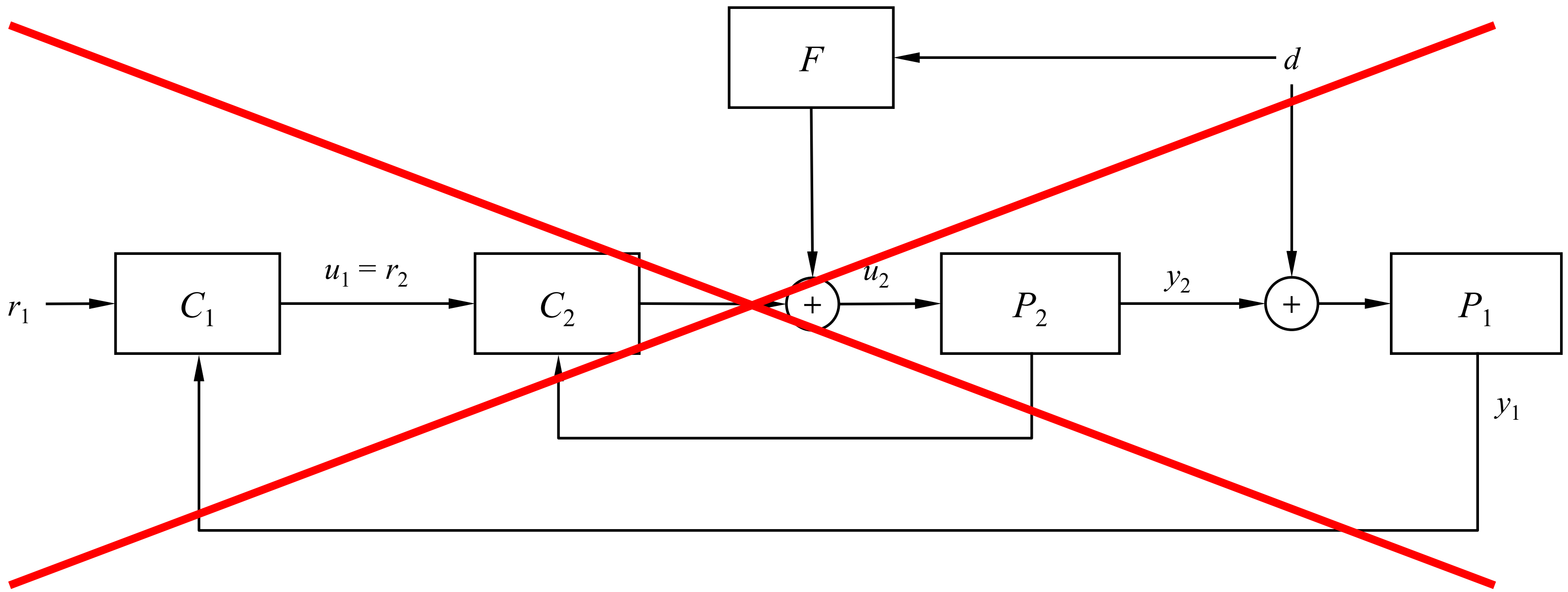
FF and cascade: Block diagram



Comments on the FF/Cas combination

- In fact we already saw several examples of this combo.
  - The first example with feedforward for tank level, the example with dissolver density control, etc
- If you are not careful, the implementation can go wrong.
  Below is a common mistake, feeding forward to the slave controller instead of the master.
  - This is completely pointless: d doesn't even affect y2.
    The FF term will be immediately counteracted by the slave controller integral part.

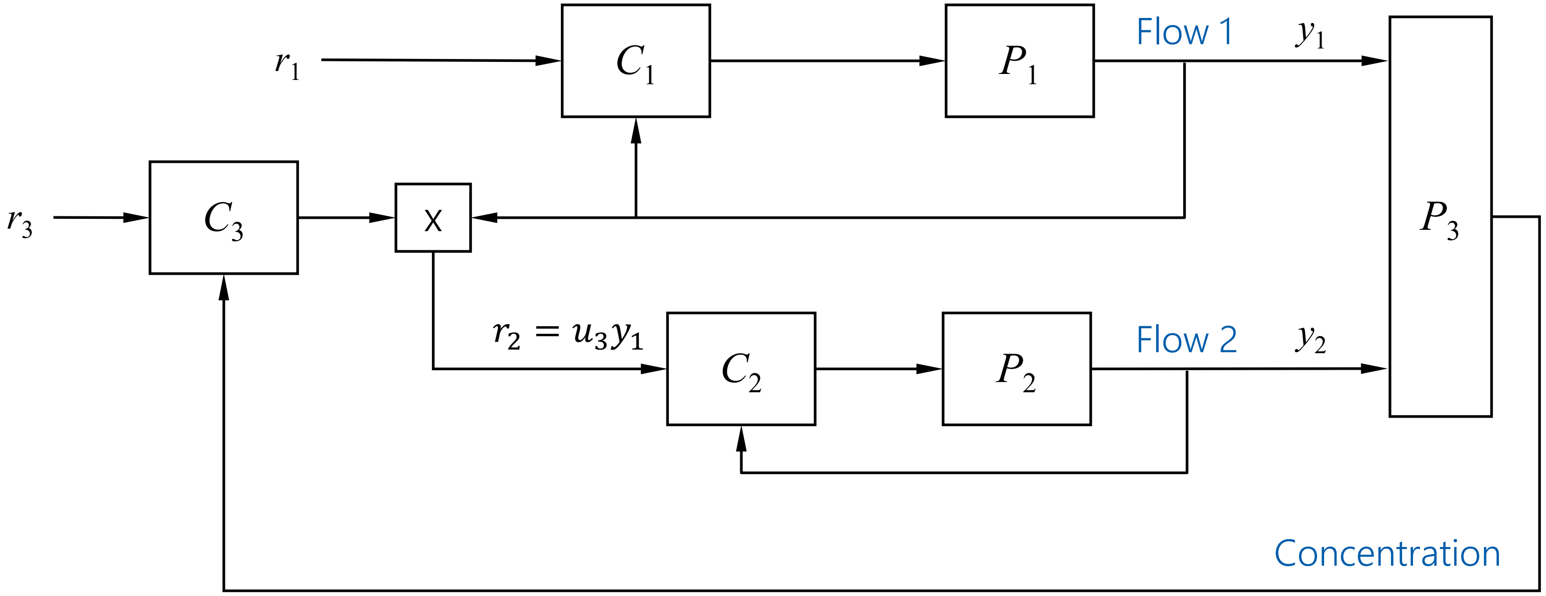
Warning: DON'T do like this!



Ratio control

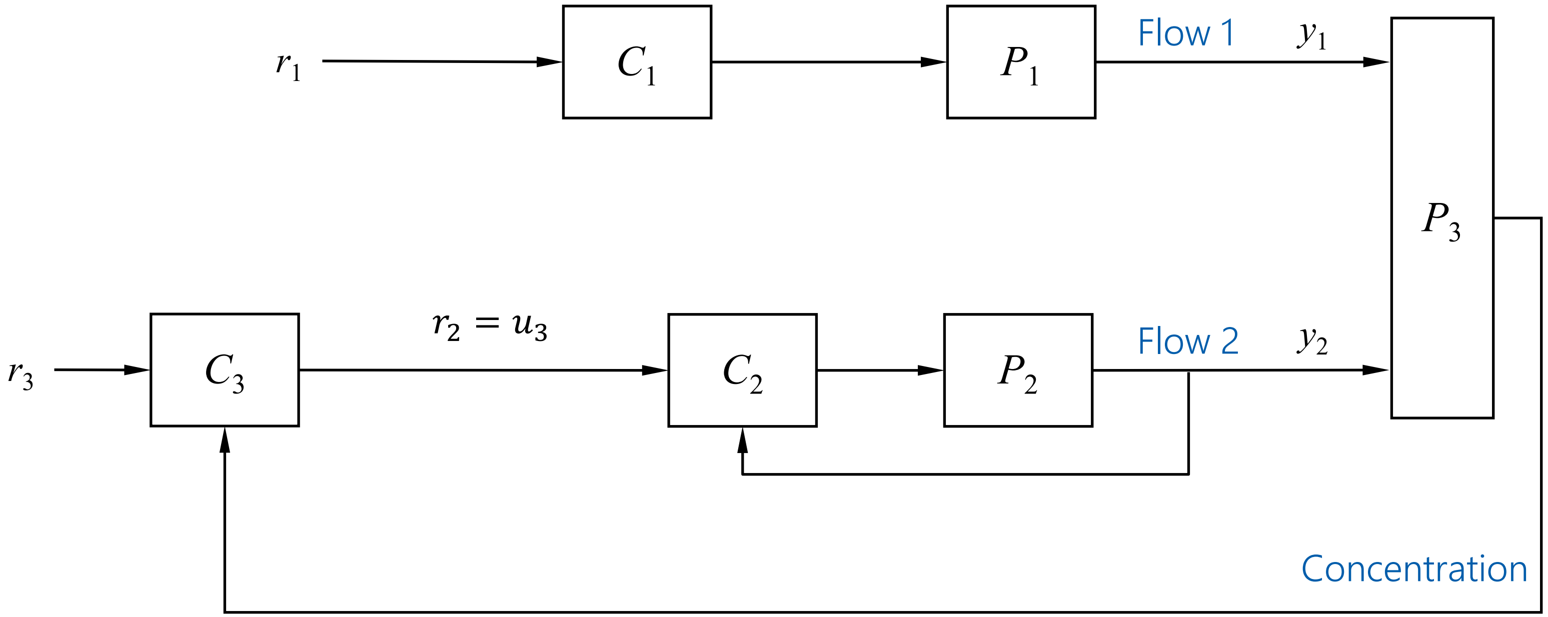
Ratio in cascade

Typical examples: reactor, mixing, dilution
This is an important and useful structure that is not well known by practitioners.

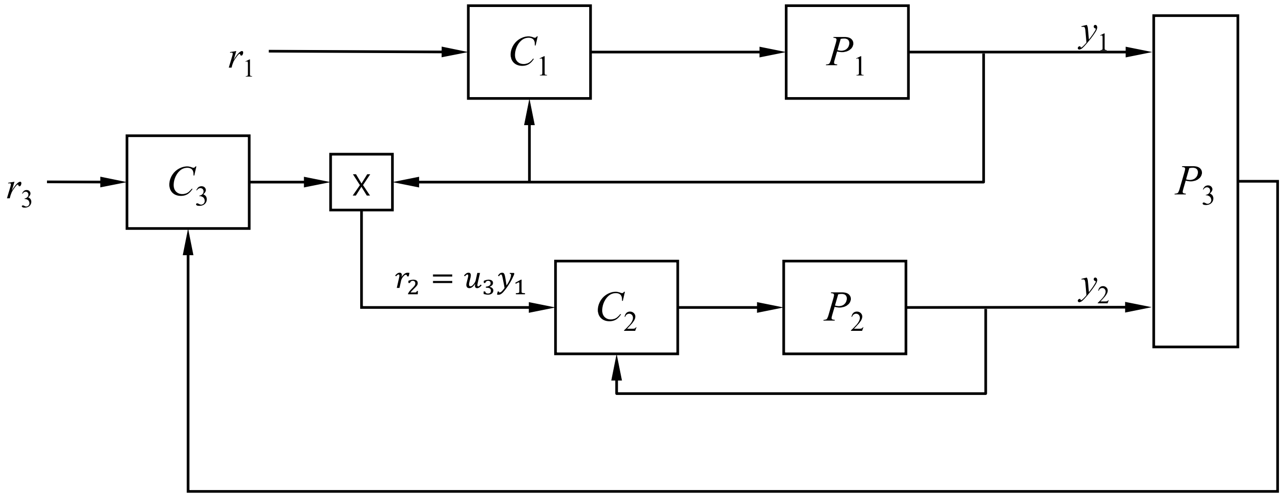


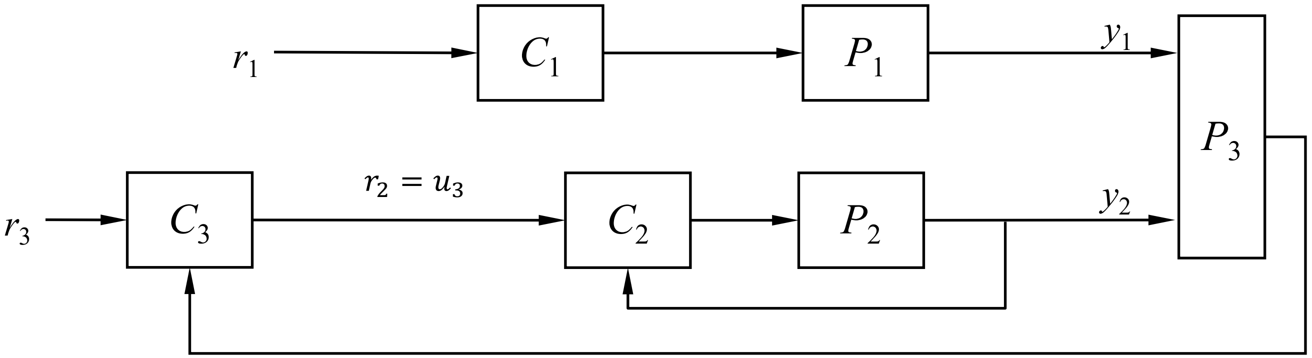
Cascade control, for comparison

This structure is an alternative solution to the same problem.
Exercise: In which way is this structure inferior to ”ratio in cascade”?



Compare structures



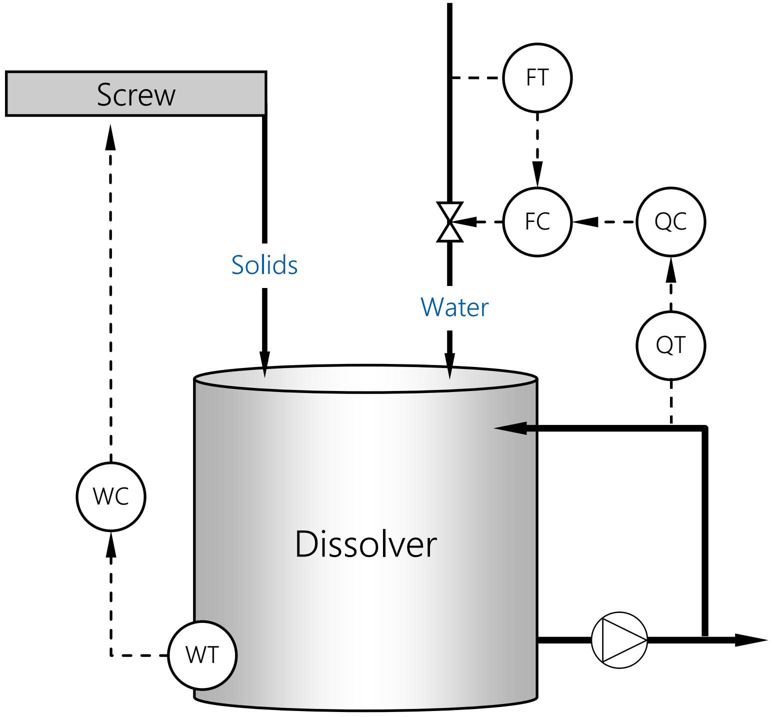


Exercise: In which way is cascade control inferior to "ratio in cascade"?

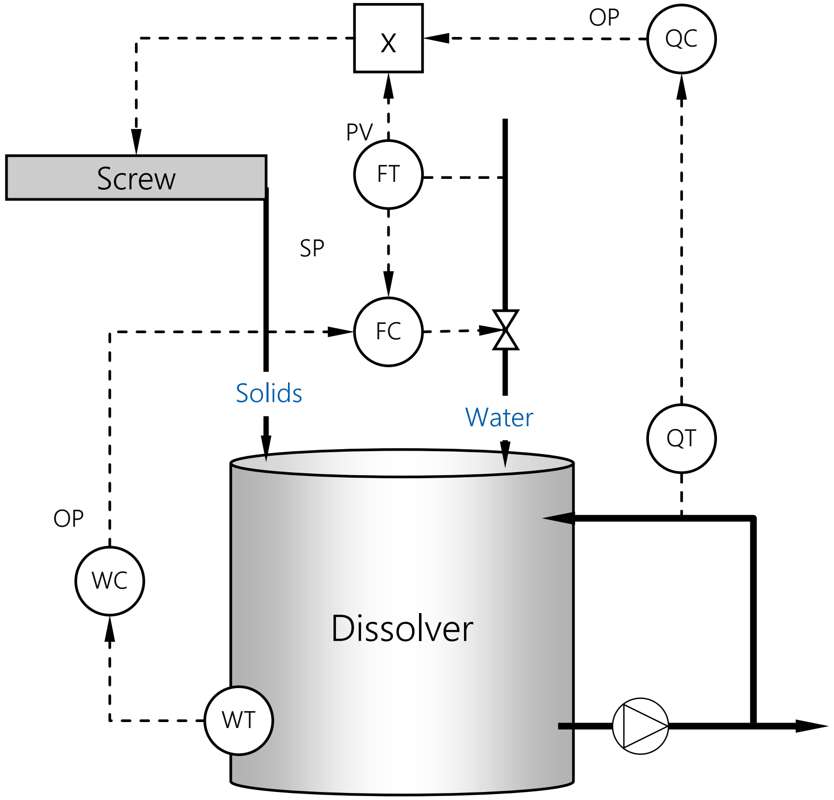
Answer: With ratio control any change in the master flow immediately changes the secondary flow SP, without going through P_3.

Ratio in cascade reduces interaction

Direct cascade control:
Concentration and level controllers may disturb eachother



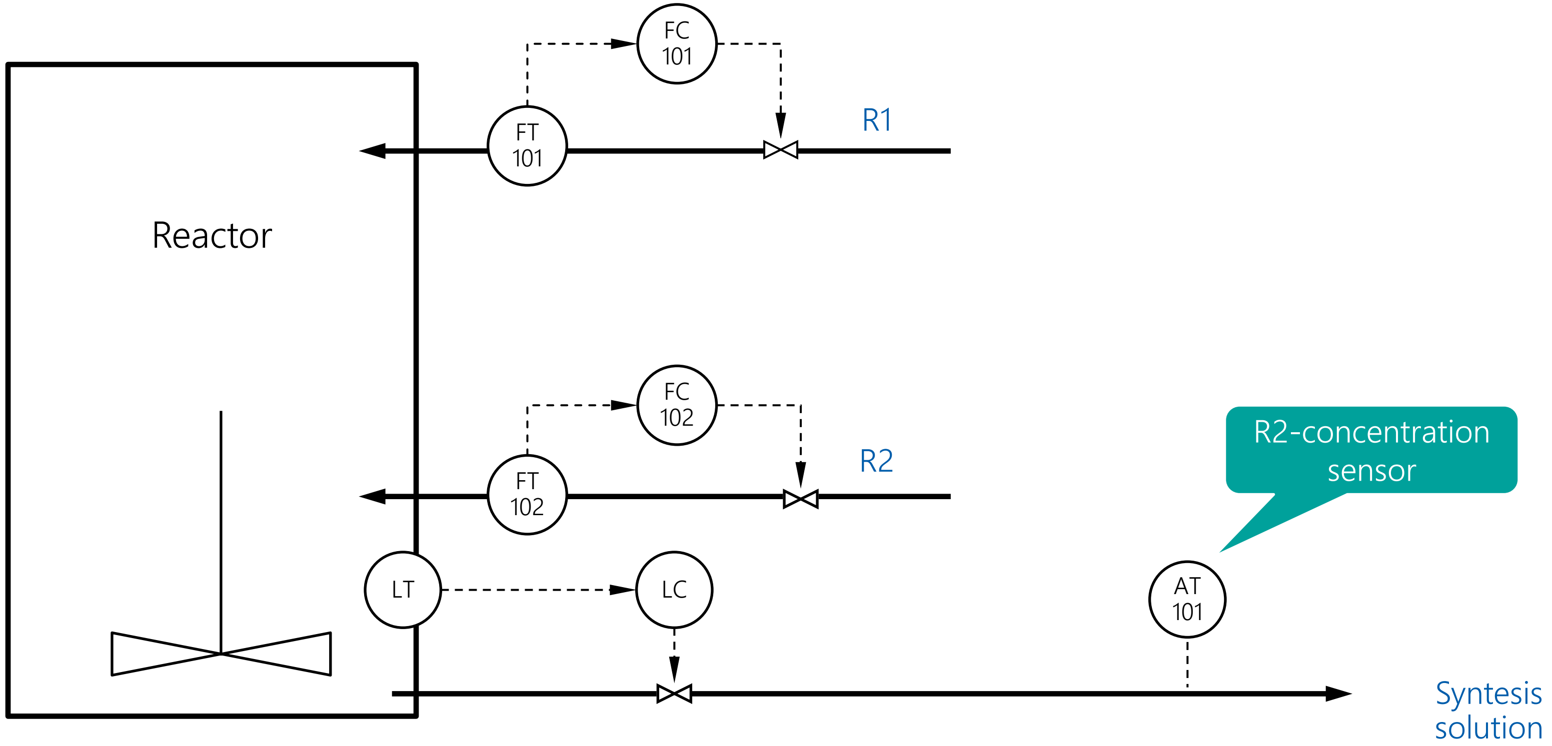
Ratio in cascade:
Level controller does not disturb concentration controller



This is an example of **decoupling**

Exercise: Suggest control structures

- Process: A continuous reactor with two reactant feeds: R1 and R2
- R1 is master flow (throughput manipulator). R2 is in excess.
- Suggest different ways of controlling the R2 concentration in the outgoing flow, and compare them.
  - Which solution is superior in which situation?
  - Don’t care about level control.



Alternative 0: Conc controller manipulates the valve

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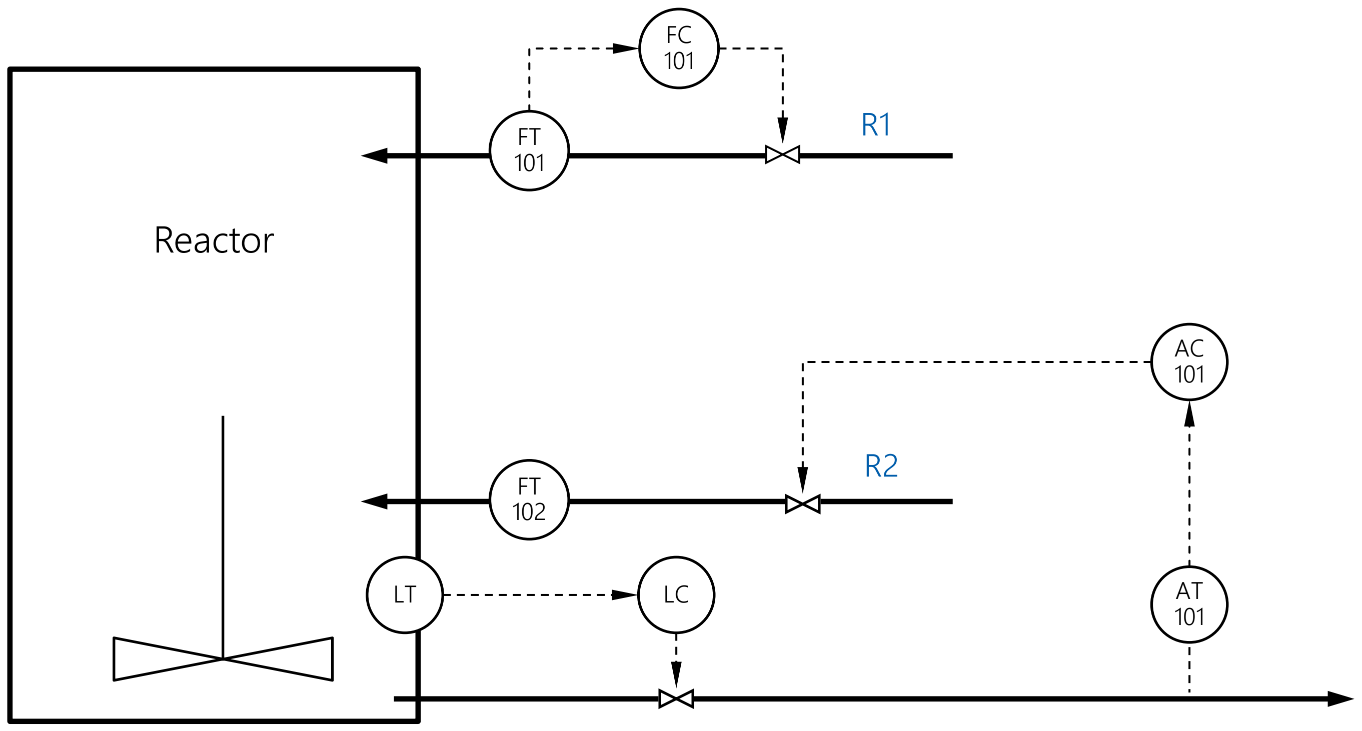

Alternative 1: R2 in cascade against conc

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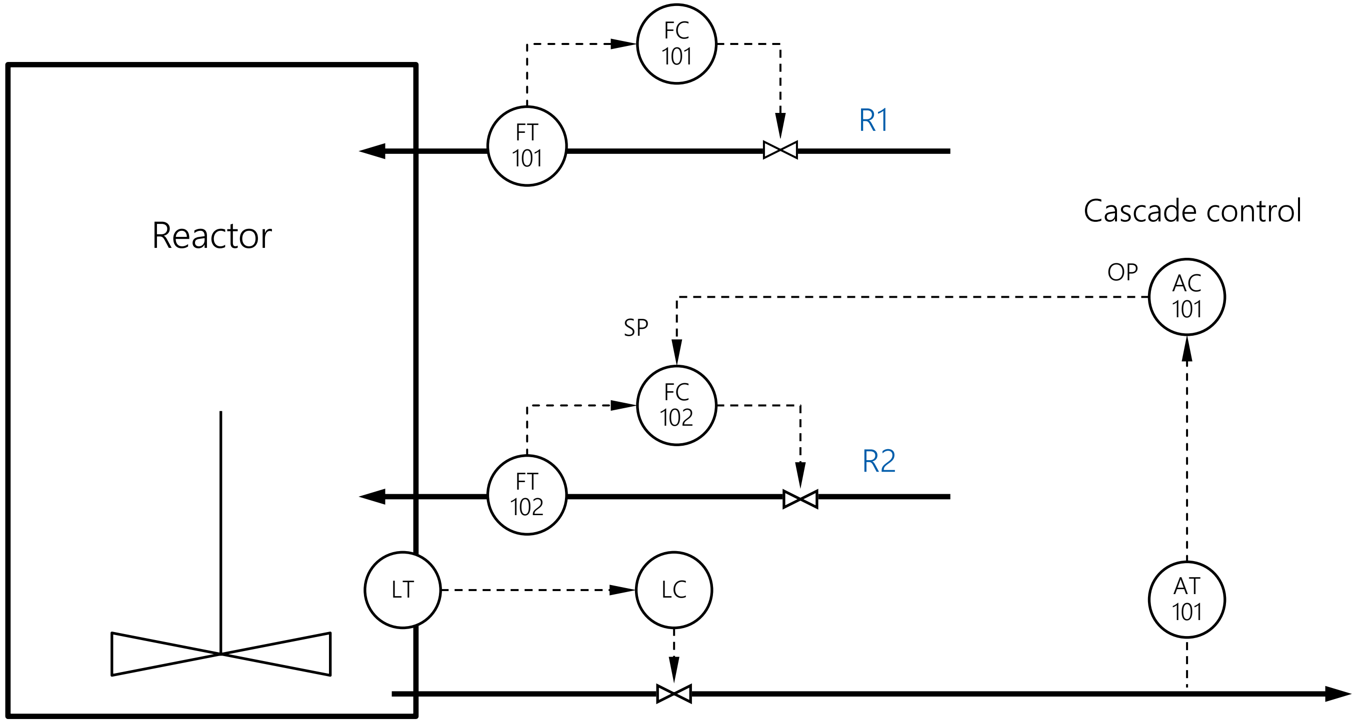

Alt 2: R2 slave in cascade + feedforward from R1

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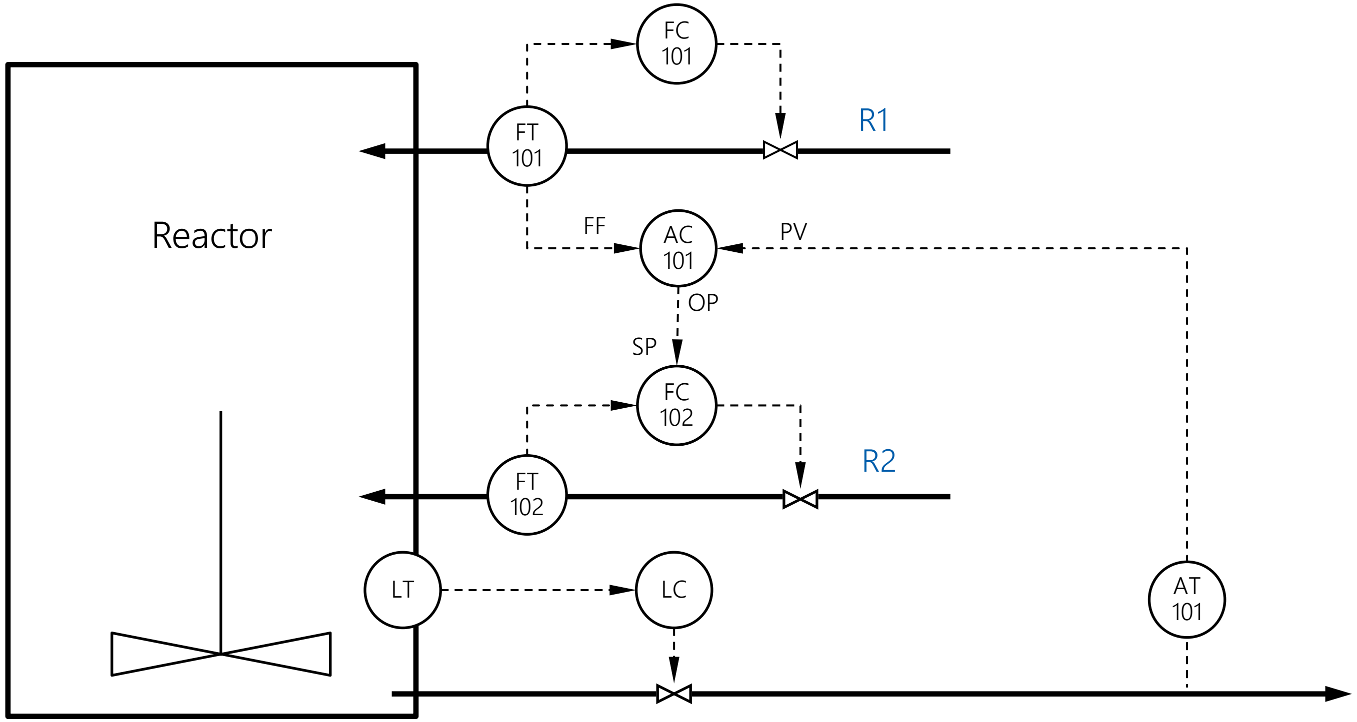
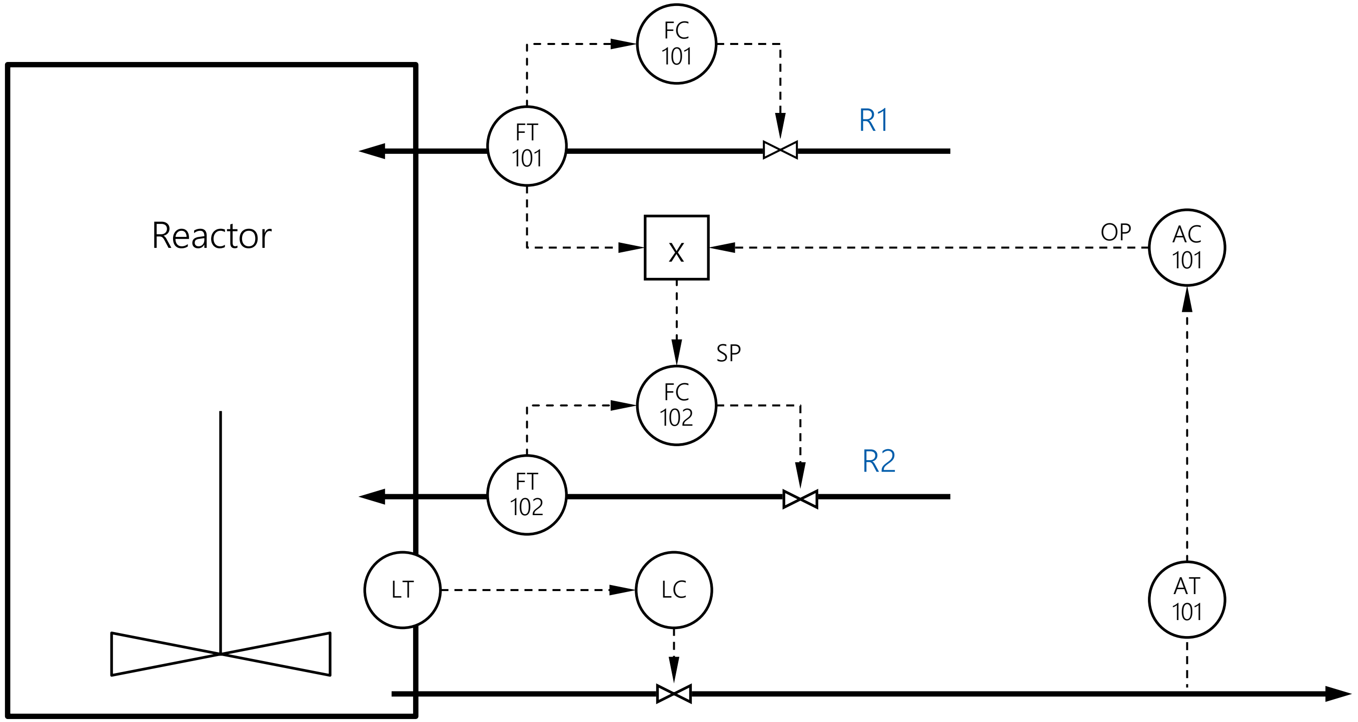
This structure is better than the previous one if the flow of R1 varies, e.g. during start-ups.



Alt. 3: R2 in ratio against R1, cascade to conc

This structure is even better than the previous one. “Adaptive feedforward”

Feedforward requires a model.
In this structure the AC does the modeling.



Question: Is this the same as ratio in cascade?

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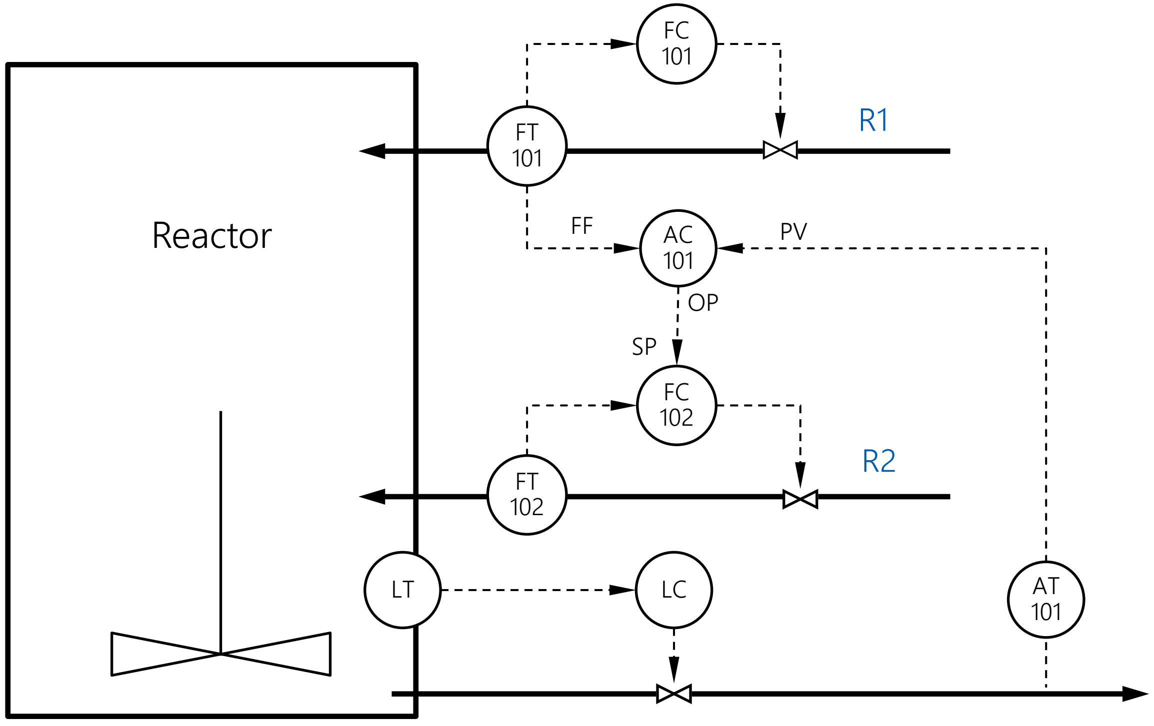


If not, what are the advantages and disadvantages of each structure?

Which disturbances are best handled by which structure?

Answer: No, it is not the same.
If the concentration in stream R1 changes, then flow rate variations are better managed by ratio in cascade.

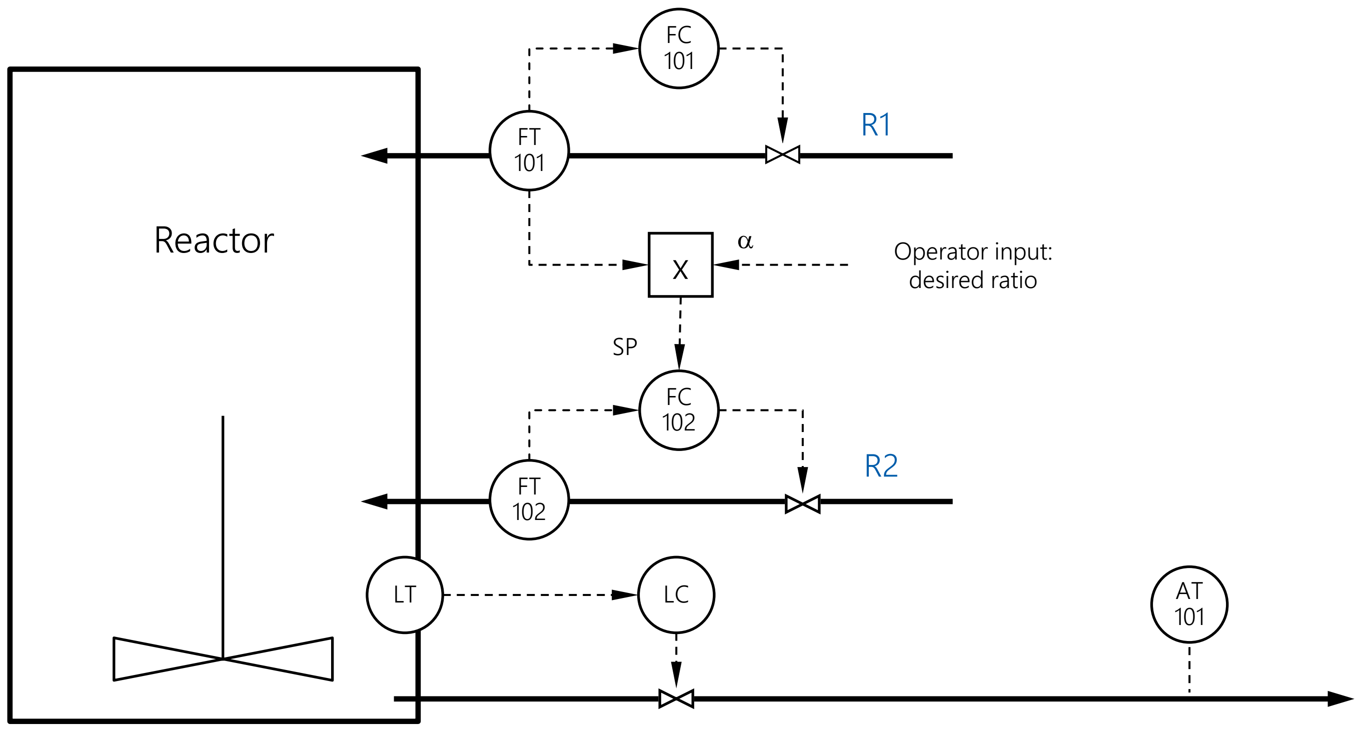
Ratio in cascade can be considered as ”adaptive feedforward”.

Scenario: The conc in R1 is ”permanently” changed.
Ratio-in-cascade corrects for this only once.
The next production rate change does not create a transient.
But for ”cascade + FF” we get the transient every time the production rate changes.

In addition: The feedforward gain would need to be adjusted for different setpoints of the AC when using ”cascade + FF”.

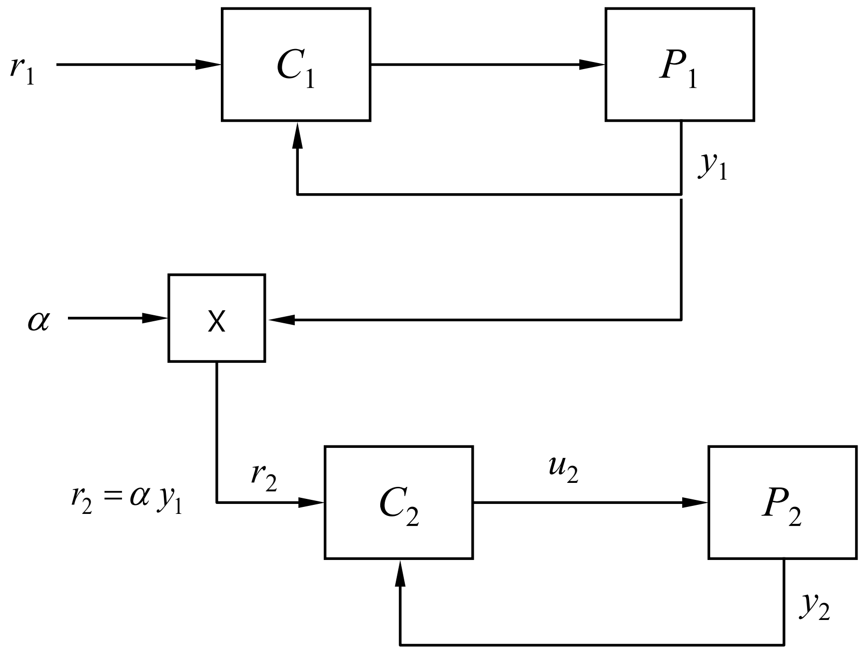
Not an option: Here the conc is not controlled

Ratio control without feedback is not a solution. Here, the concentration is not controlled.



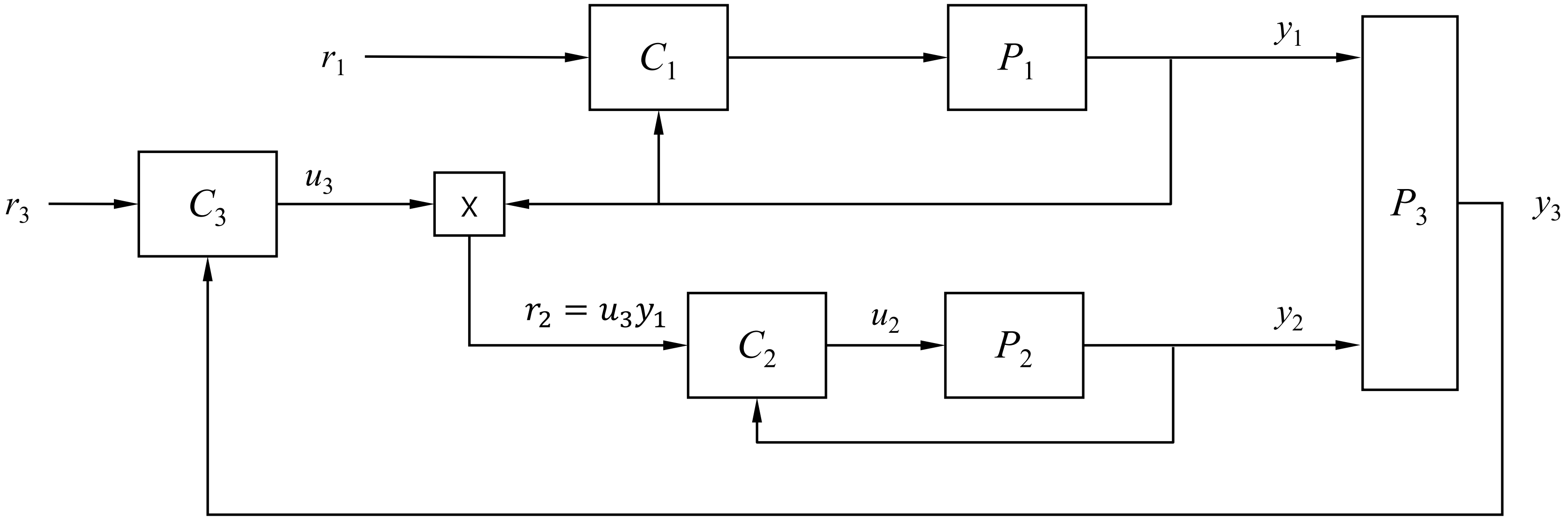
Traditional ratio control, without feedback

- A potential problem in ratio control is that if the slave controller output (manipulated variable) saturates, e.g. having output u=100%, then the control is lost.
  - The desired ratio α is set by the operator.
- If u_2 reaches its max value, then the ratio will deviate from the desired value in steady state.

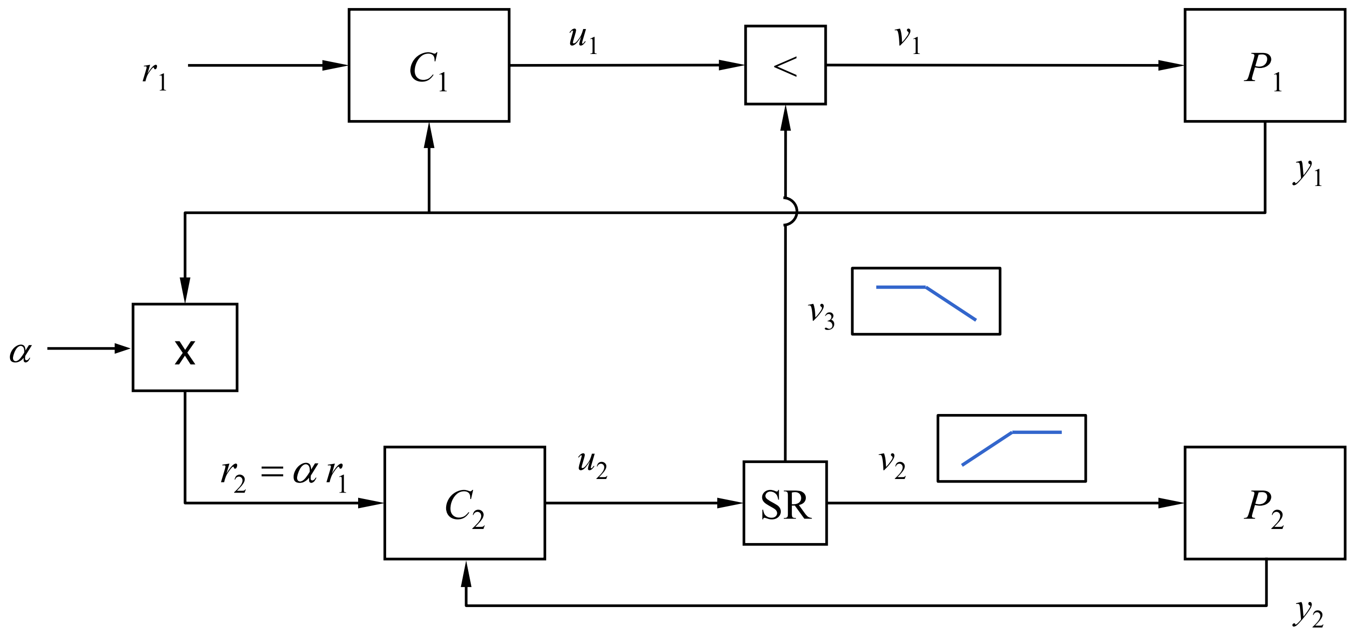


Ratio control with feedback; same issue

- This structure can be used when there is a measurement y_3 dependent on the ratio of flows 1 and 2.
- However, it suffers from the same problem when u_2 saturates.



Solution: New structure using SRC and selector

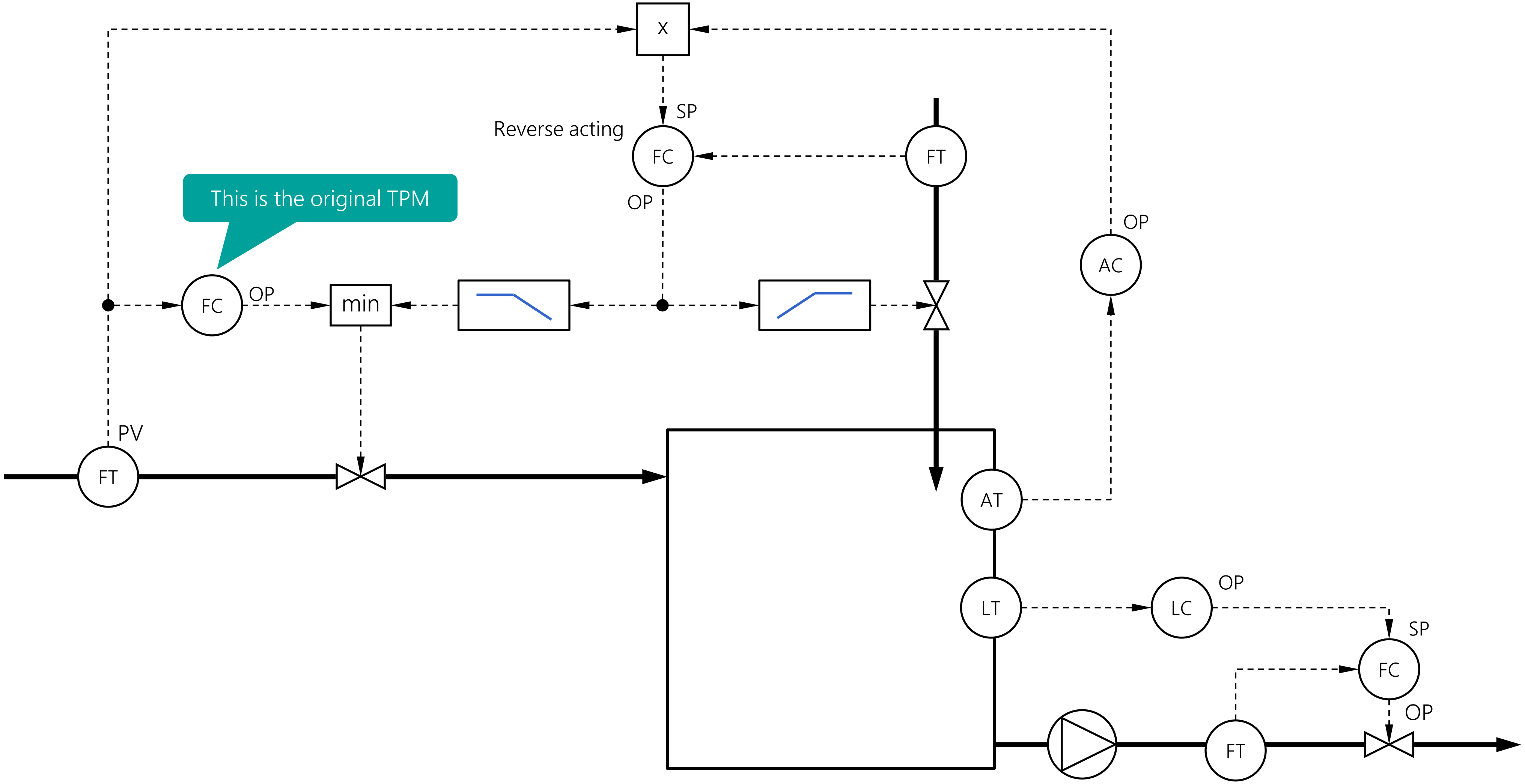


$$v_1 = \min(u_1, v_3)$$

SR-table (example)

u_2	v_2	v_3
0	0	100
50	100	100
100	100	0

Application example: Blending



A note on "multiplicative feedforward"

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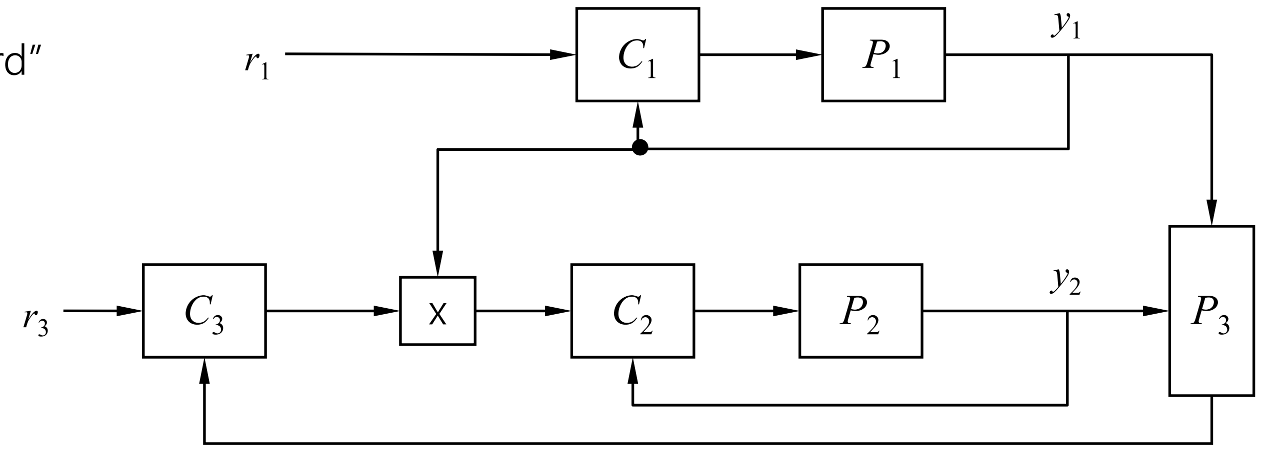
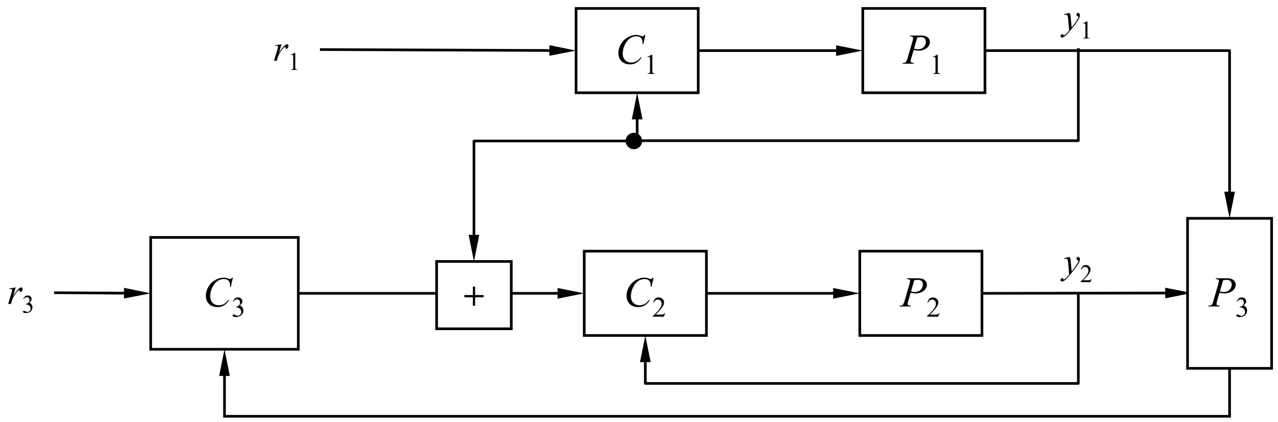
"Multiplicative feedforward"



Honeywell:
"Additive feedforward"

(This is not ratio control)



Physical interpretation of the ratio

- Specification above: keep the ratio $\frac{y_2}{y_1}$ constant

- However in most applications, this is not the relevant physical quantity.
- Consider a mixing process, e.g. dilution (concentration control).
  - Then the relevant quantity is

xk = feed stream conc

fk = flow

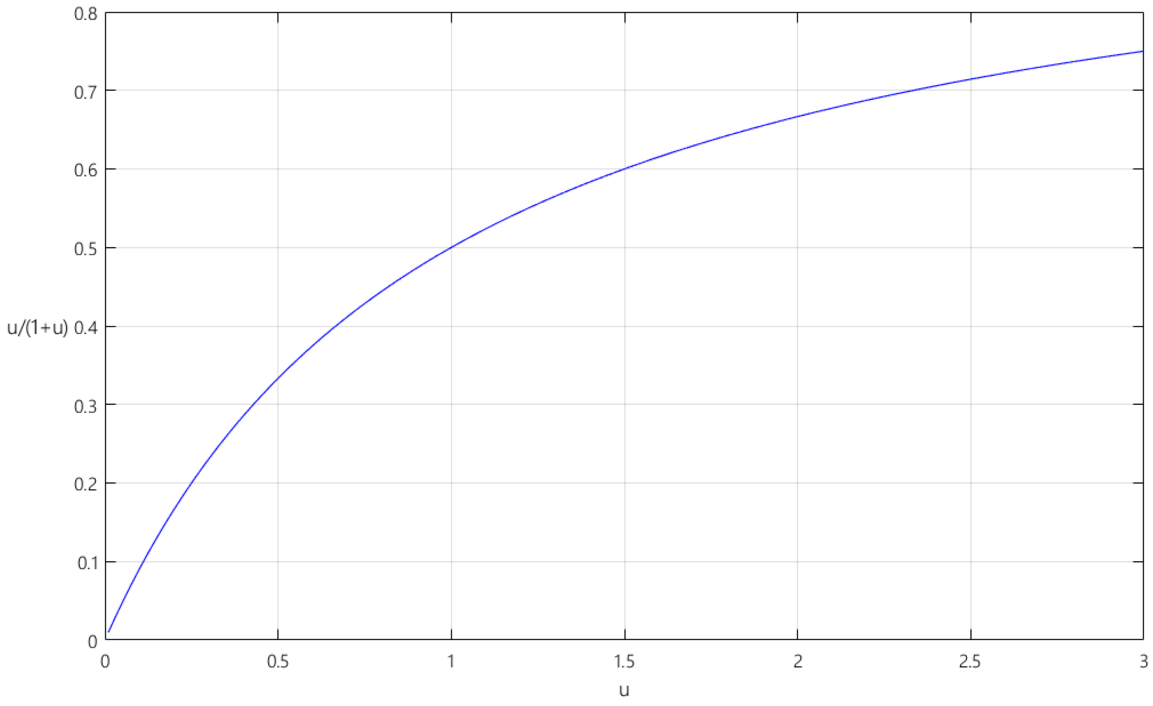
$$\frac{x_1 f_1 + x_2 f_2}{f_1 + f_2}$$

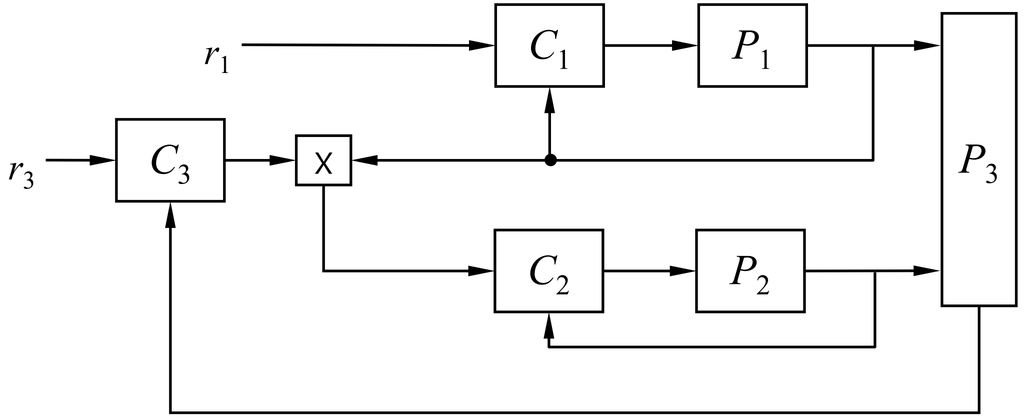
- In the case of dilution (f2 being water) we get $\frac{x_1 f_1}{f_1 + f_2}$

- One guiding principle is to choose an architecture that makes the static relation between MV and CV as linear as possible.
- If the CV is concentration, then "classical" ratio control is not optimal in this sense.

Relation between the two ratios

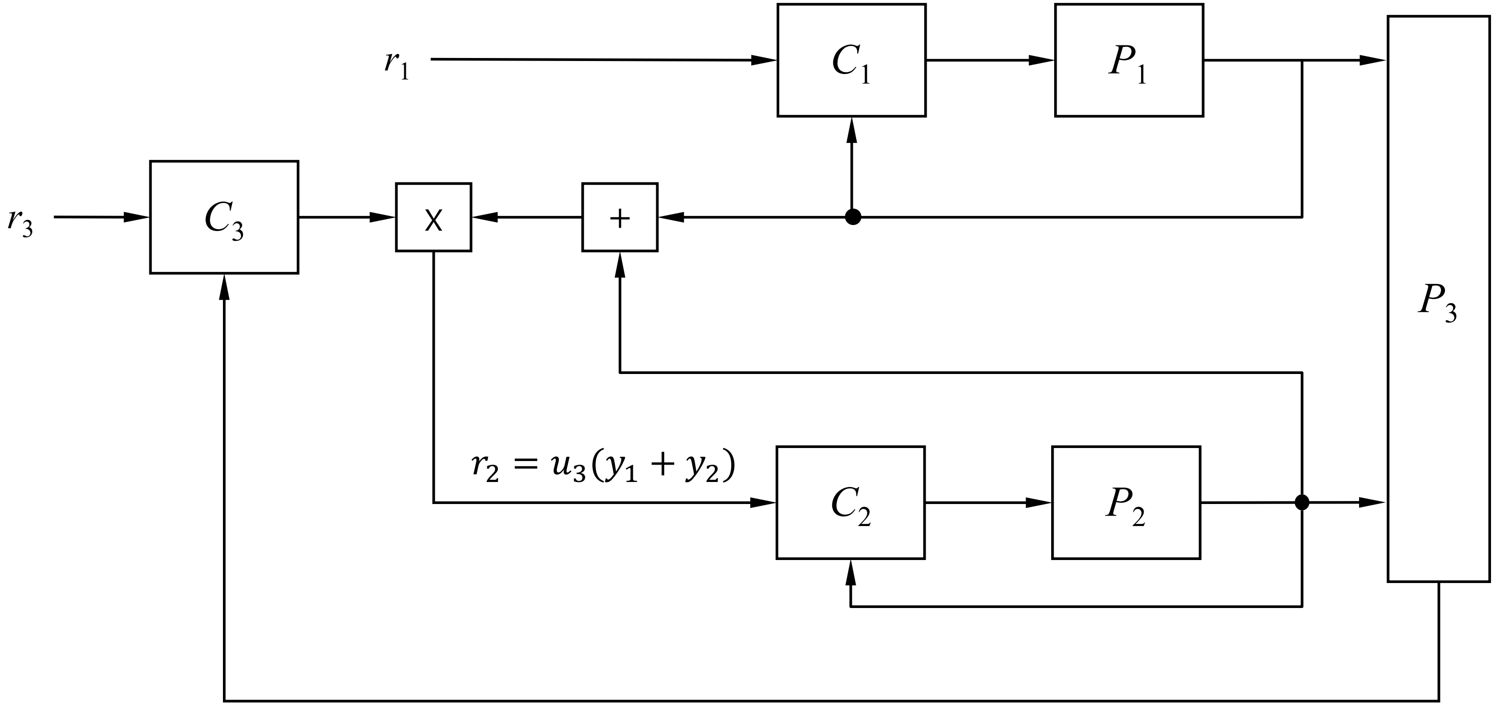
$$u = \frac{y_1}{y_2} \qquad r = \frac{y_1}{y_1 + y_2} \qquad r = \frac{1}{1 + \frac{1}{u}} = \frac{u}{1+u}$$





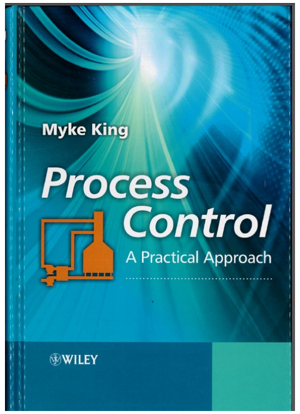
This architecture is suboptimal,
because it manipulates u to control r.

This is the "physical" control architecture



I'm not sure I like it. What dynamics does C_2 try to control?

Mentioned in M. King: *Process Control. A Practical Approach.* Wiley, 2011.
Found under "Feedforward control" 🤣



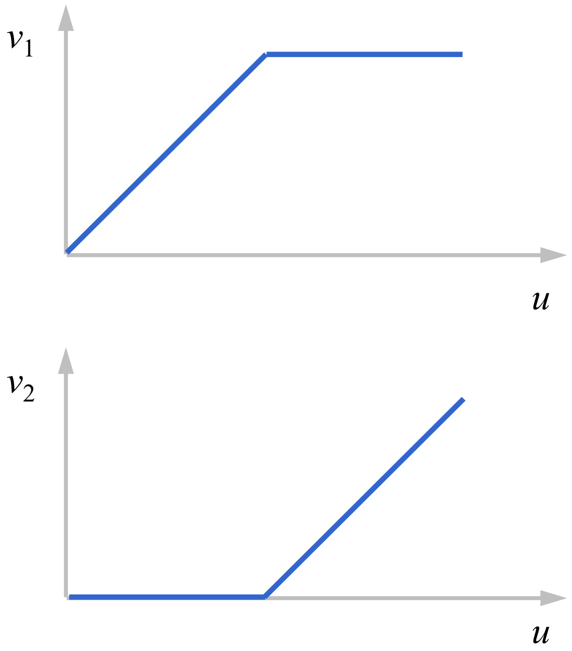
Split-range

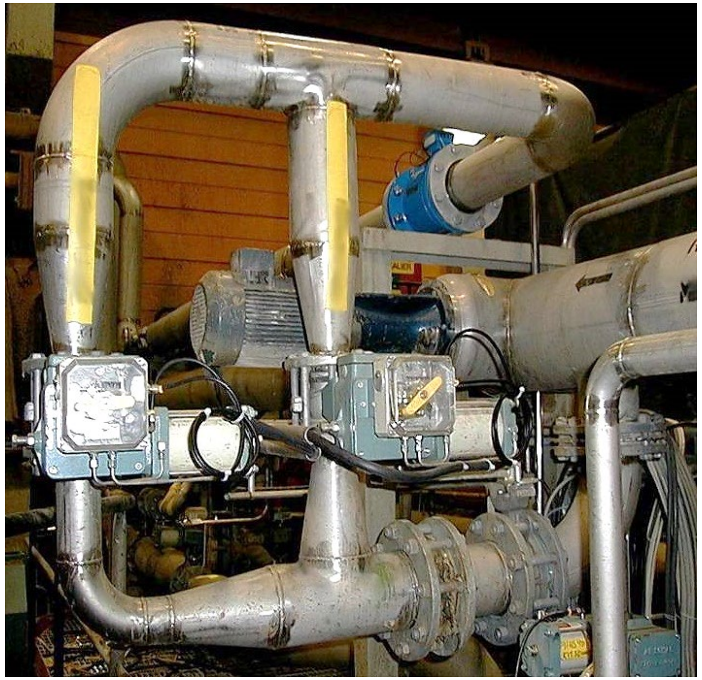
(SRC: Split-range control)

Split-range: two valve – one controller

- In some applications we have two MVs and want the controller to use one "first" and the other one next.
- The most common solution is to send the controller OP to two tables; one per valve.
  For example:

OP	v_1	v_2
0	0	0
50	100	0
100	100	100





Level control with two manipulated flows

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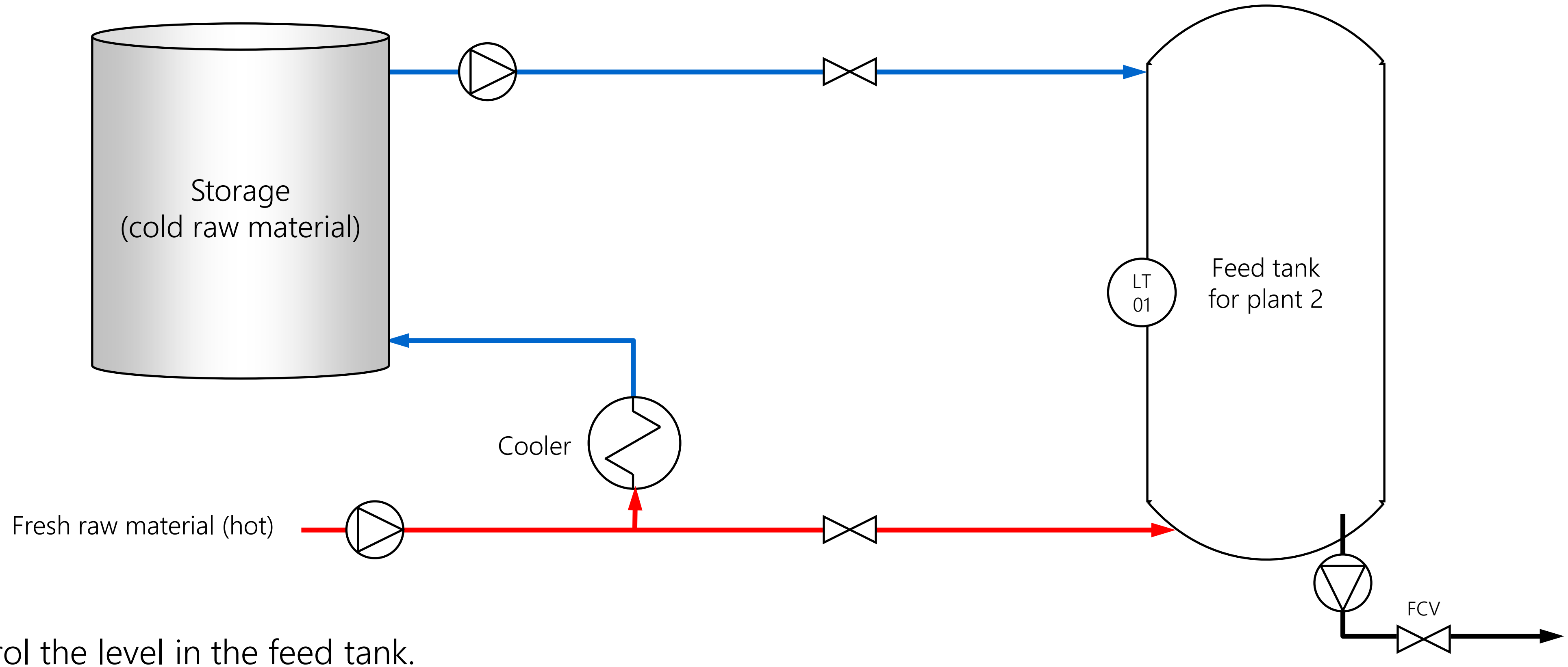


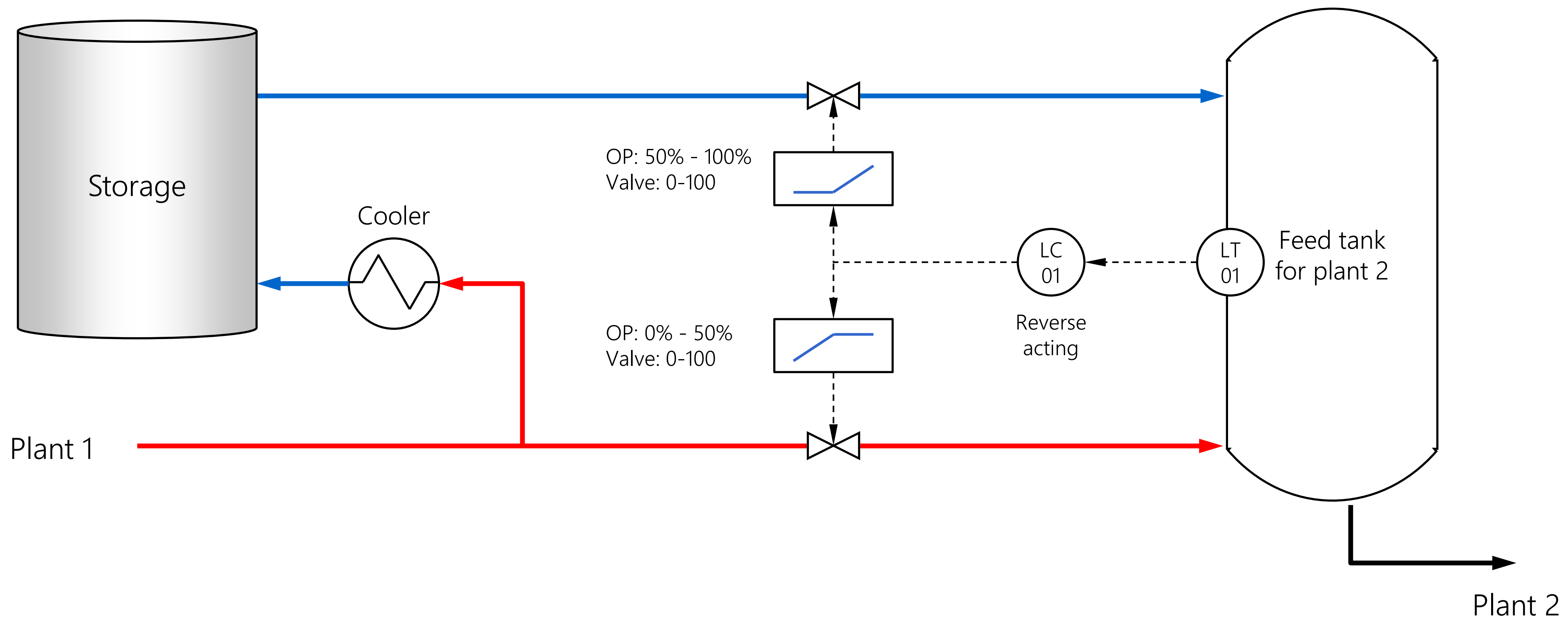
Main task: Control the level in the feed tank.

Use the extra degree of freedom for energy optimization: Primarily take fresh, hot, raw material.

If that doesn't suffice, take from storage.

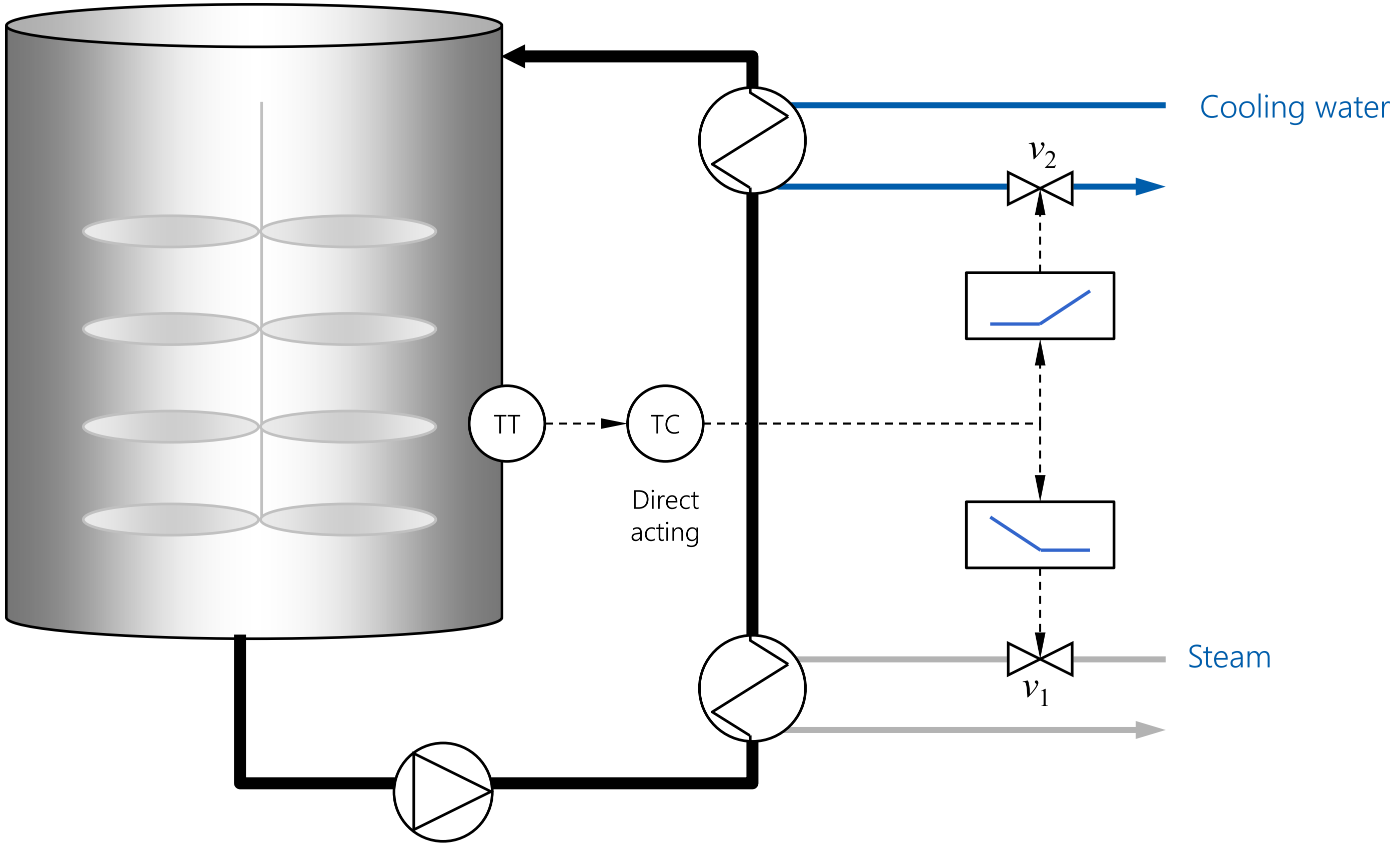
If we don't have optimizing control we may cool unnecessarily, and then re-heat.

SRC application: level control with optimization



Another SRC application: Cooling - Heating

- Another example of split-range: cooling or heating for temperature control.
  - Here the MVs are "qualitatively different" and likely to have different dynamics etc. Could be hard to tune.
  - E.g. exothermal reaction in a continuous reactor. Heating for start-ups, cooling during normal operations.



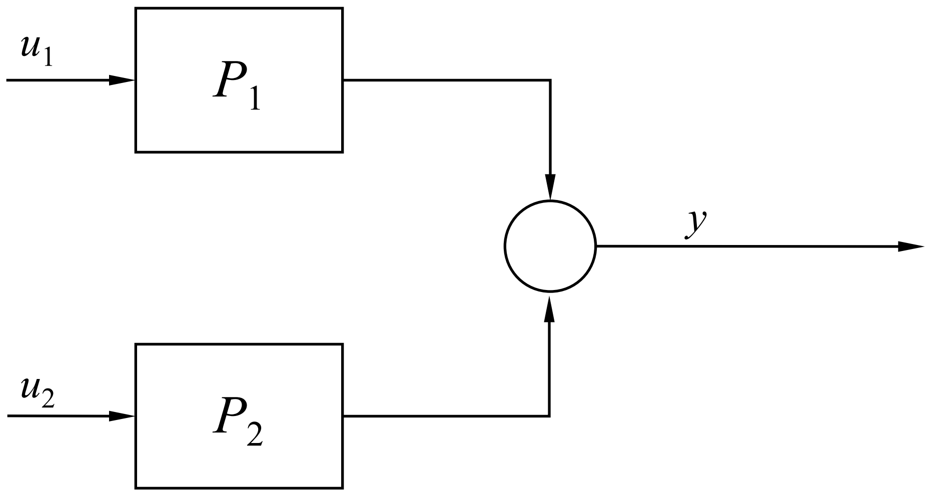
OP	Steam	C.w.
0	100	0
50	0	0
100	0	100

Valve position control (VPC)

("Mid-ranging")

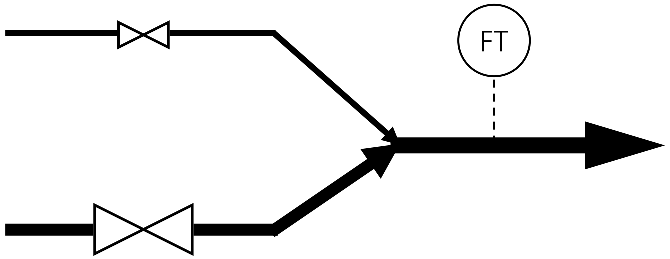
Extra degree of freedom: A coarse and a fine MV

- Scenario: We have two different MVs at our disposal to control one process variable.
- The dynamics from u_1 may be different from the one from u_2.
  Typically one has larger gain and slower response than the other.
- How do we utilize this extra DoF?

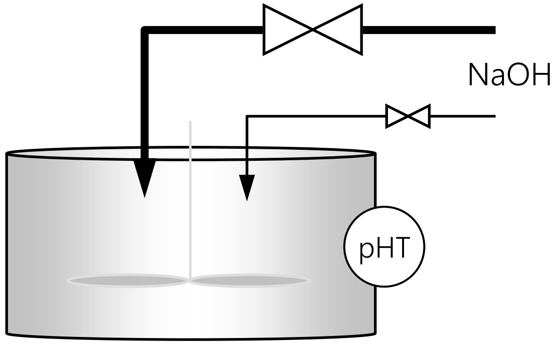


Typical processes with extra DoF

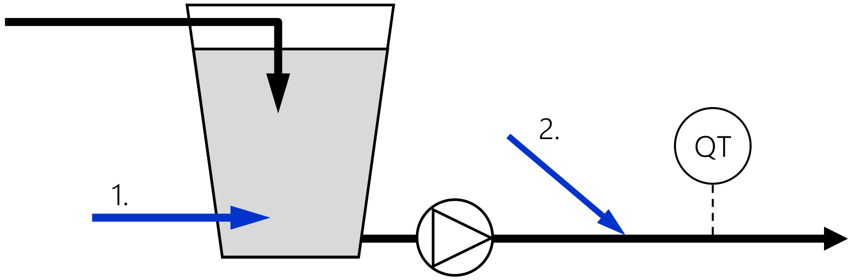
- Ex 1: Two parallel valves of different sizes manipulate the same flow:



- Ex 2: Control pH in a tank, by manipulating caustic addition. Fine and coarse valve



- Ex 3: Dilution in two steps: in tank and pipe



VPC: New controller with u_1 as controlled variable

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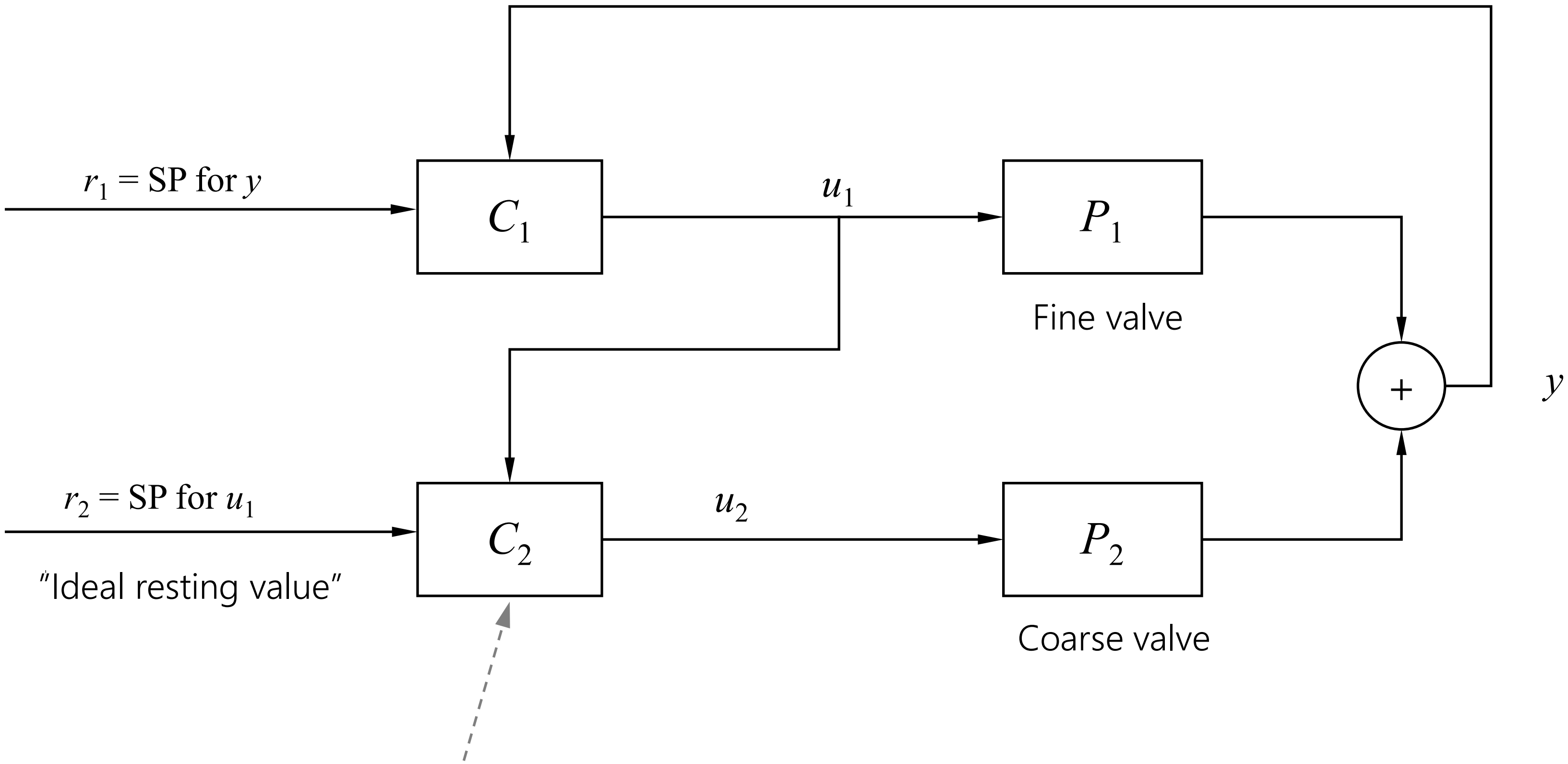
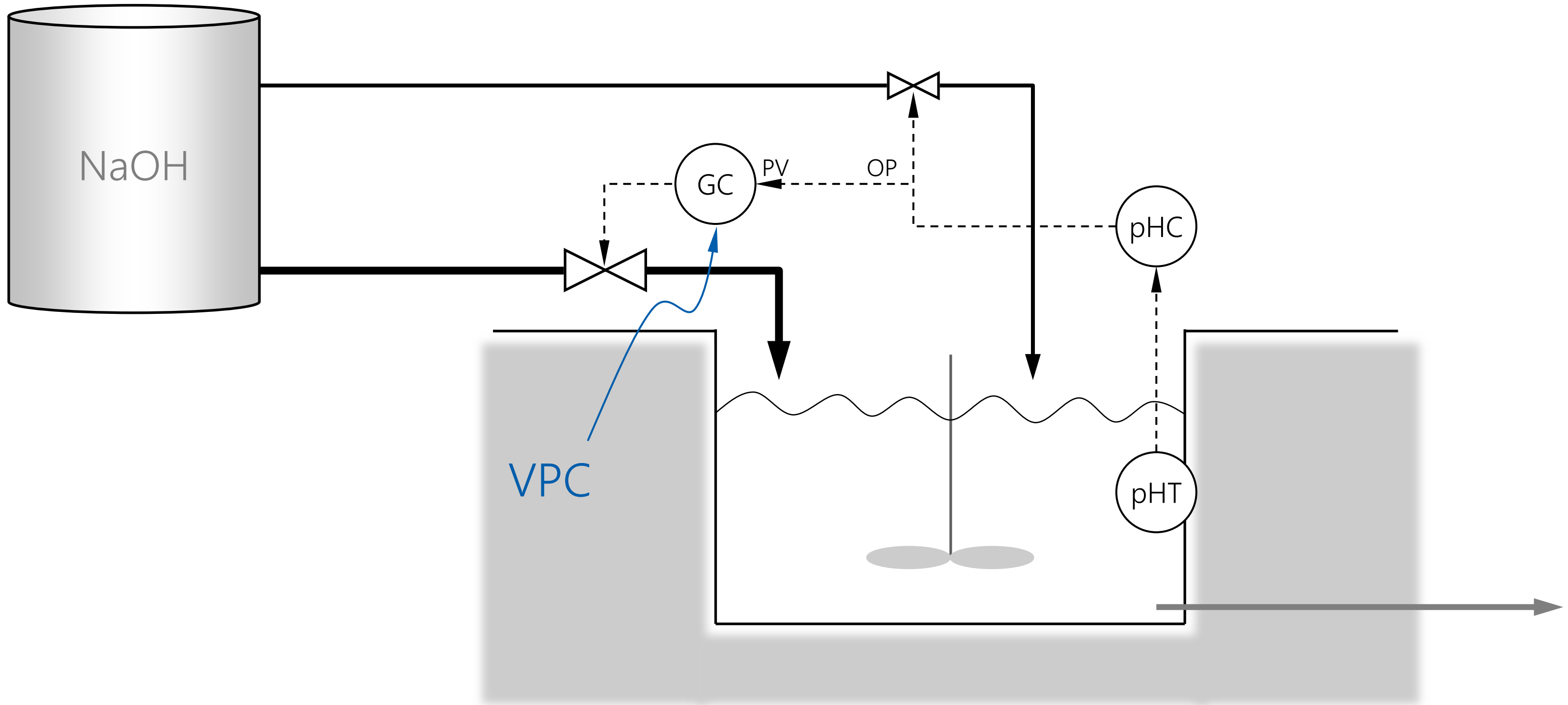


C_2 is a valve position controller (VPC)
This controller should be considerably slower than C_1

Example: Sewer pH control; Mid-ranging

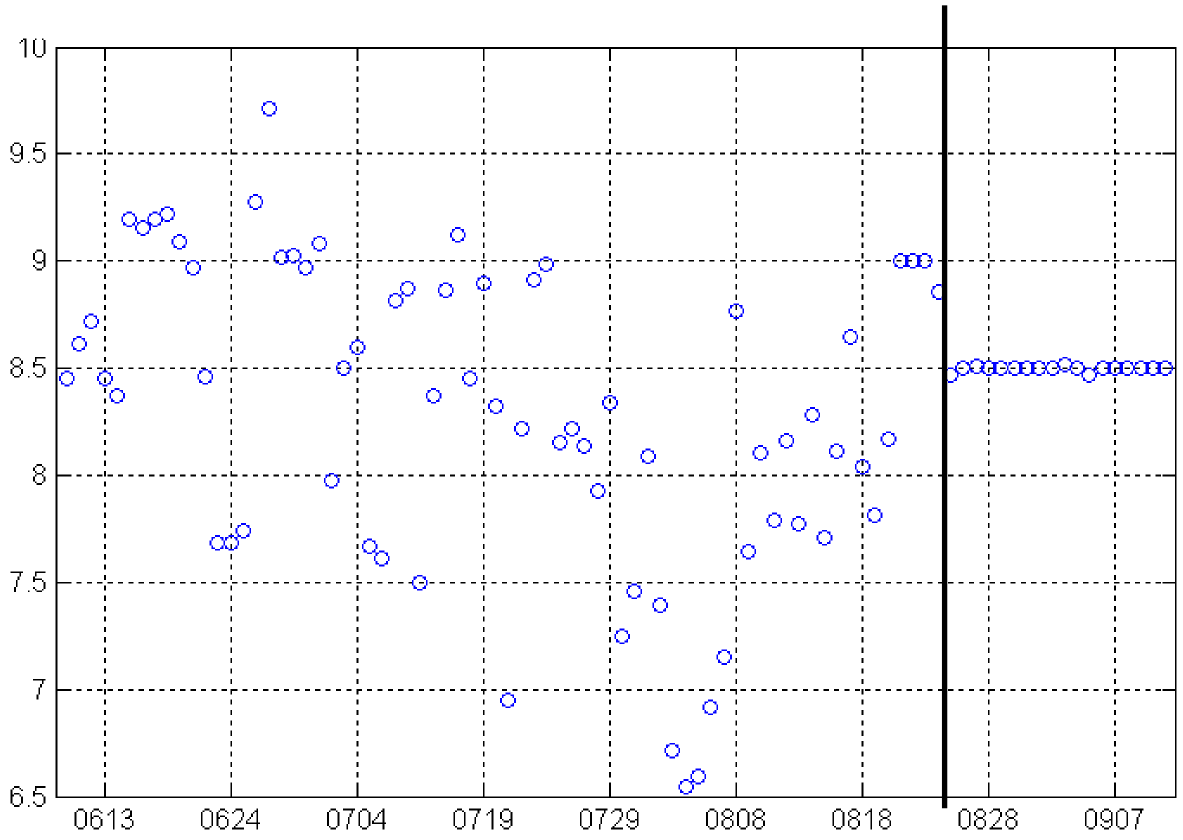
Exercise: How should we control the pH in the pit, using both valves?

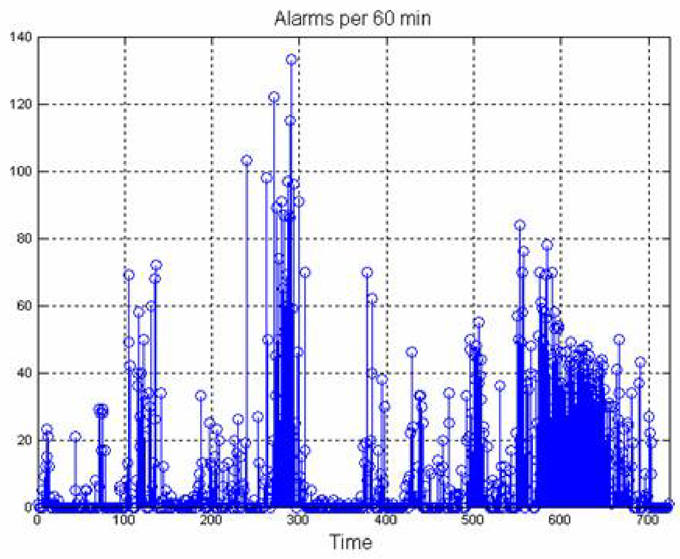


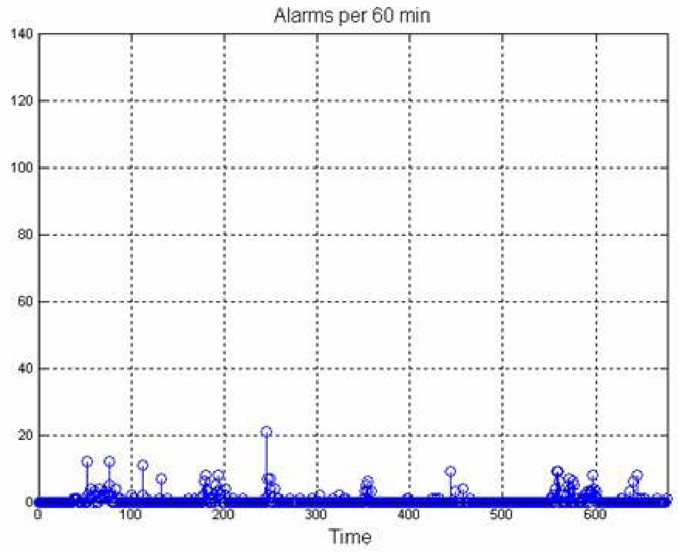
pH control: Results

Before: 10 488 alarms in one month, After: 418

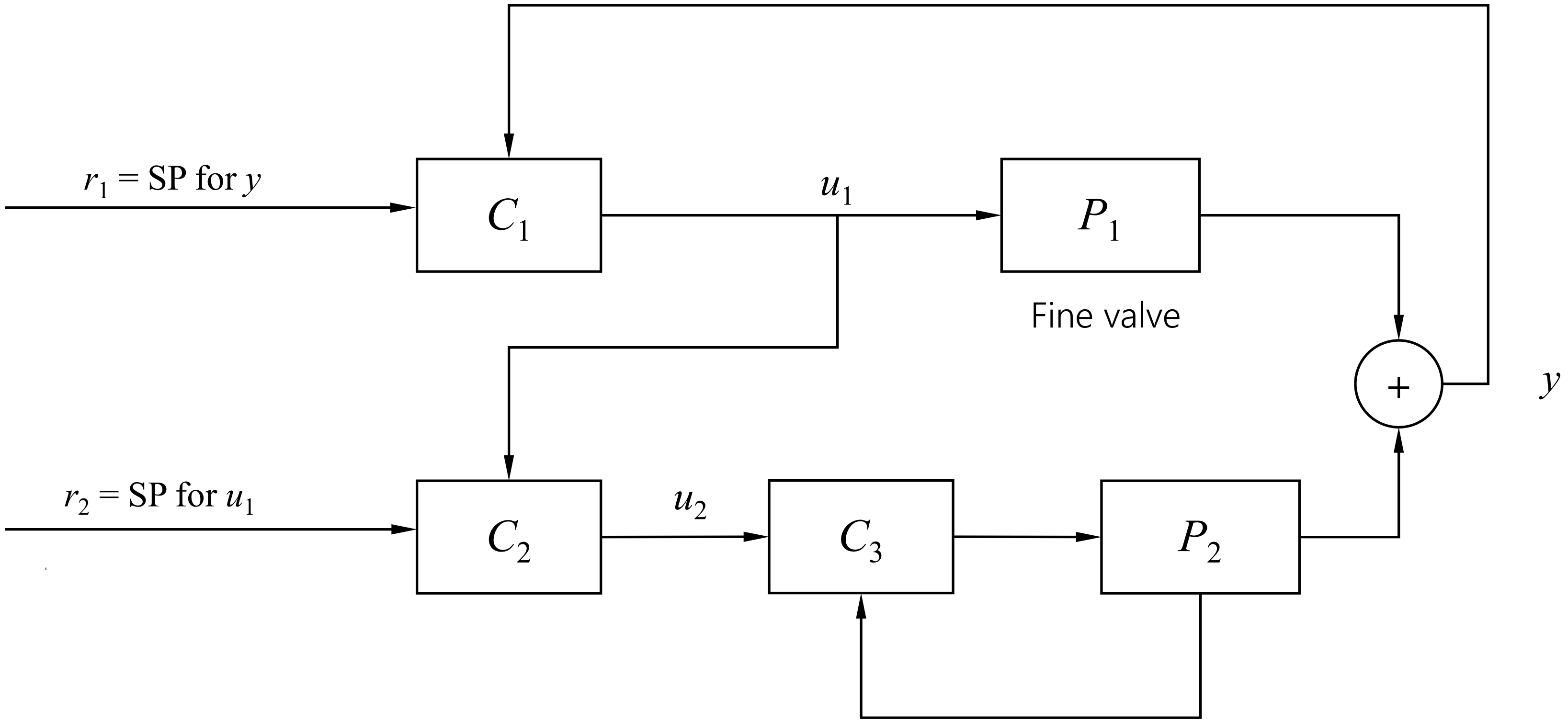
pH: Daily averages before and after VPC







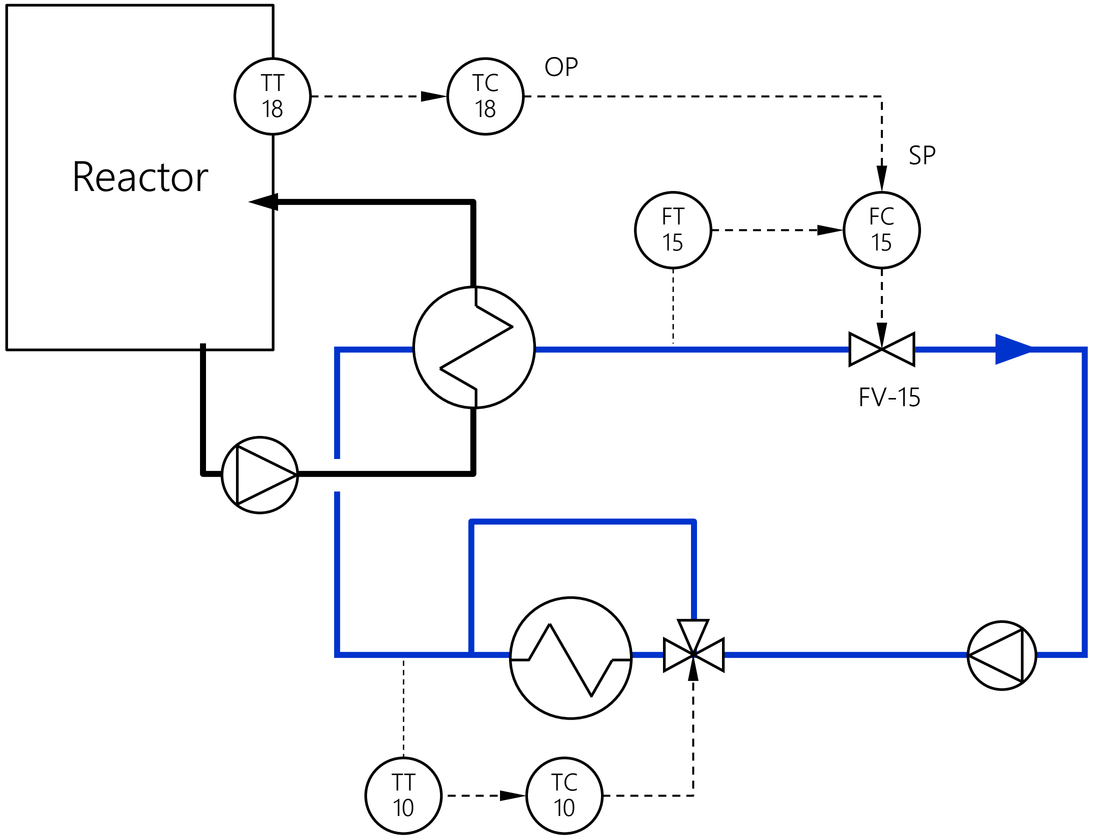
Variant: The VPC gives the SP for a slave controller



Mid-ranging application: Cooling water system

Problem: Sometimes the flow valve FV-15 saturates.
Then the operator has to change the SP for cooling water temperature.

Exercise: Suggest a mid-ranging solution that does this automatically!



Solution: New controller (VPC) giving the SP for TC

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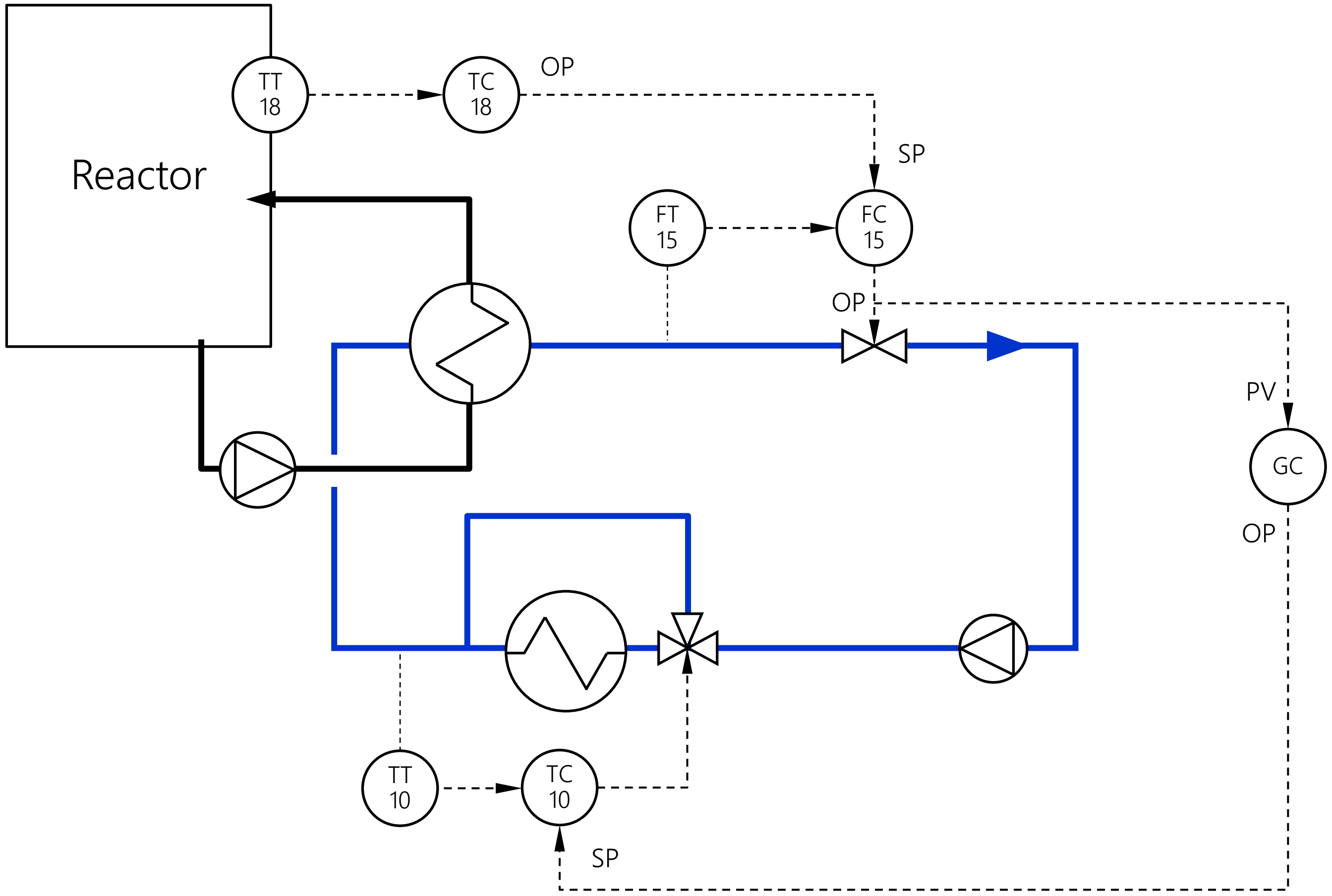
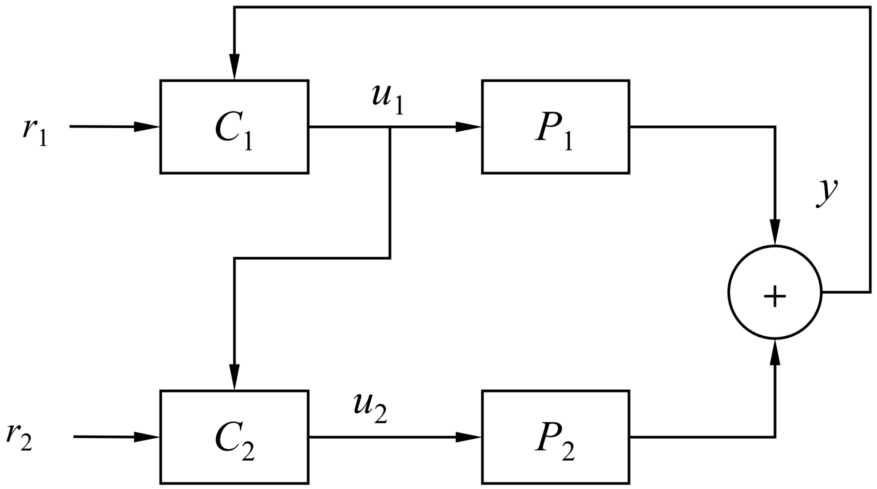
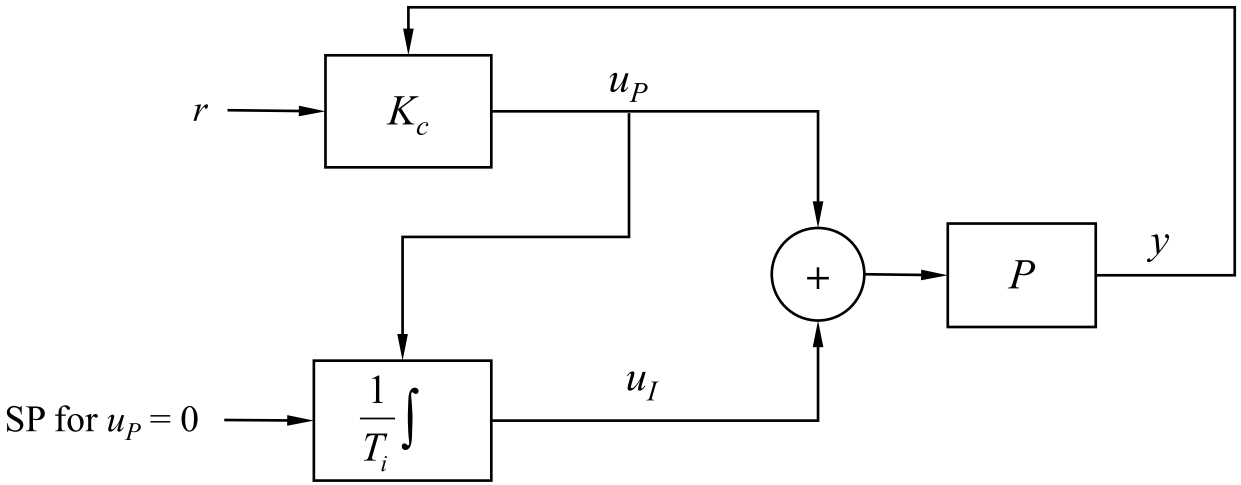


VPC = valve position controller

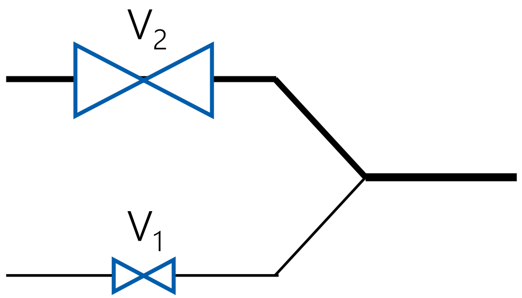
Fun fact: VPC ~ PI-controller





- The PI-controller can be seen as a special case of mid-ranging, internally.
  - The I-part and the P-part can be seen as two controllers. The task of the I-part is to reset the P-part to 0.

Split-range or mid-ranging?



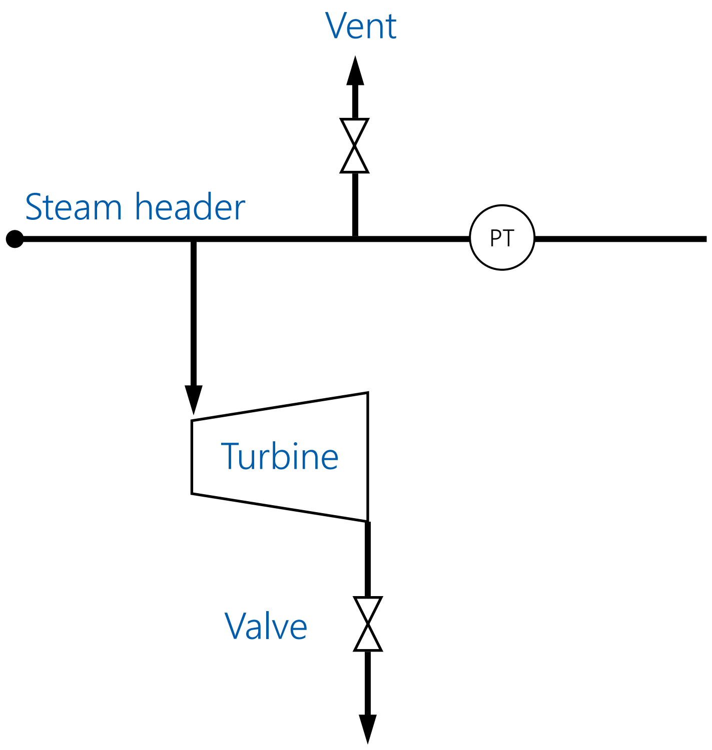
- If we have a process with big difference in valve dynamics and gain: Should we use mid-ranging or split-range?

- Answer: It depends on the specifications (the optimization criterion).
  - Split-range lets us use stream 1 to the max before taking from stream 2. Could be optimal if number 1 is cheaper.
  - Mid-ranging gives maximum control precision.

Split - parallel control

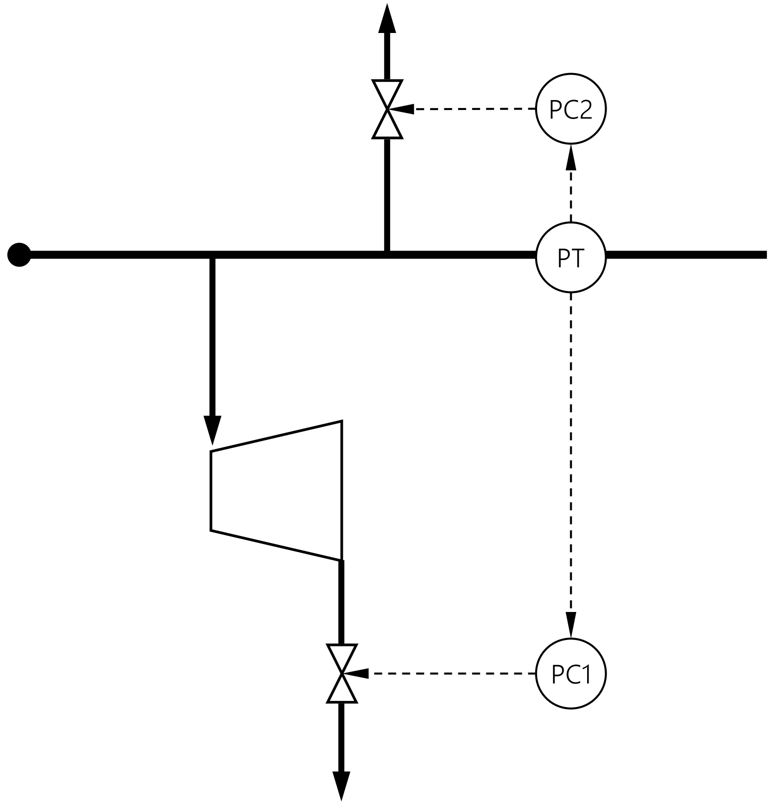
Control with several MVs

- Task: Control the pressure in the high pressure header by manipulating flow through the turbine.
- If the HP header pressure gets too high, open the vent valve.
  - But only in that case.



Solution: Two controllers with different SP

"Split parallel control"



The two controllers have the same PV, but different MVs.
They should have different setpoints, and can be tuned differently.

In this application PC2 could even be a P-controller:
It doesn't really need to enforce setpoint adherence.

If we use P control the valve position for the vent valve will be simple function of the header pressure.

Test question: Which of PC1 and PC2 should have the highest SP?

Answer: PC2. Its OP should be 0% almost always.

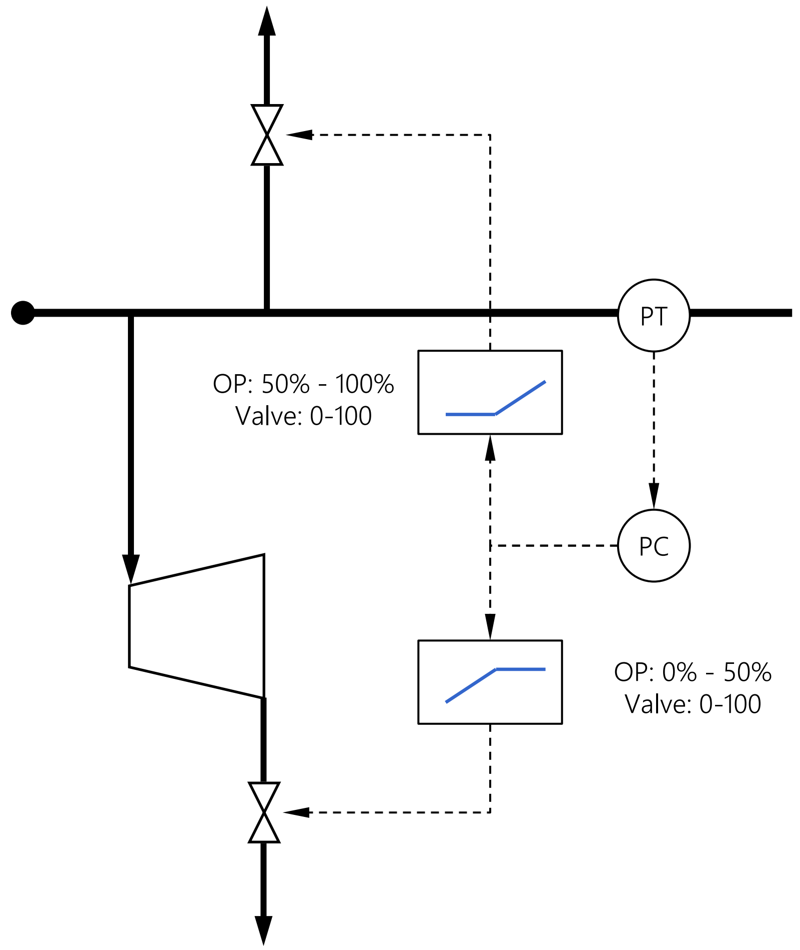
A different solution: Split-range

A split-range could also solve the problem.

Does split parallel control have advantages compared to SRC?

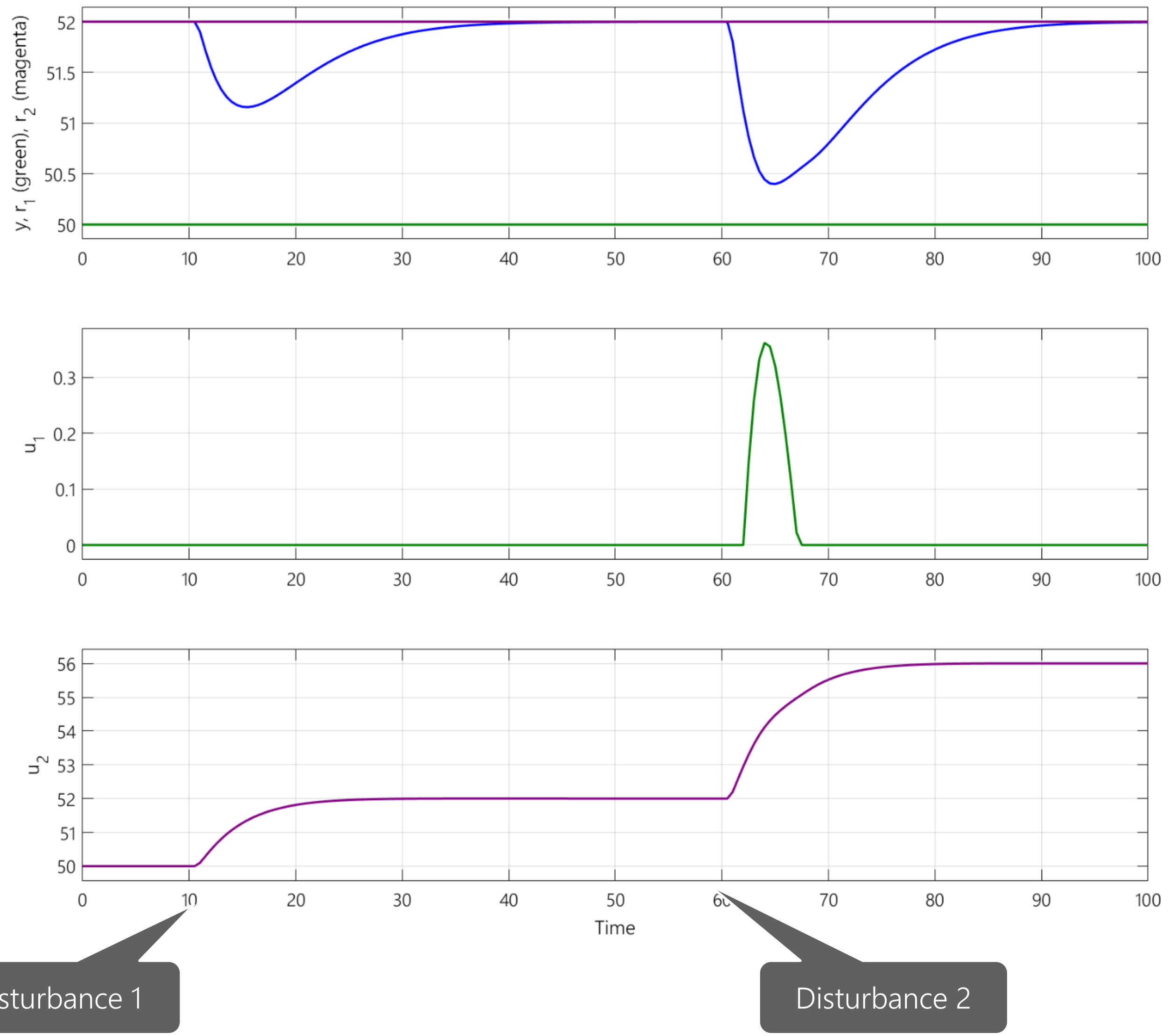
Answer:

1. With parallel controllers you can have different tuning parameters if the processes have different dynamics.
2. With parallel controllers the operator can run each valve manually, independently of each other.
3. Split parallel control allows disturbances of short duration without immediately switching MV. "Grace period"
4. In some cases it is optimal to have two different steady state levels.



The third advantage (don’t switch MV immediately)

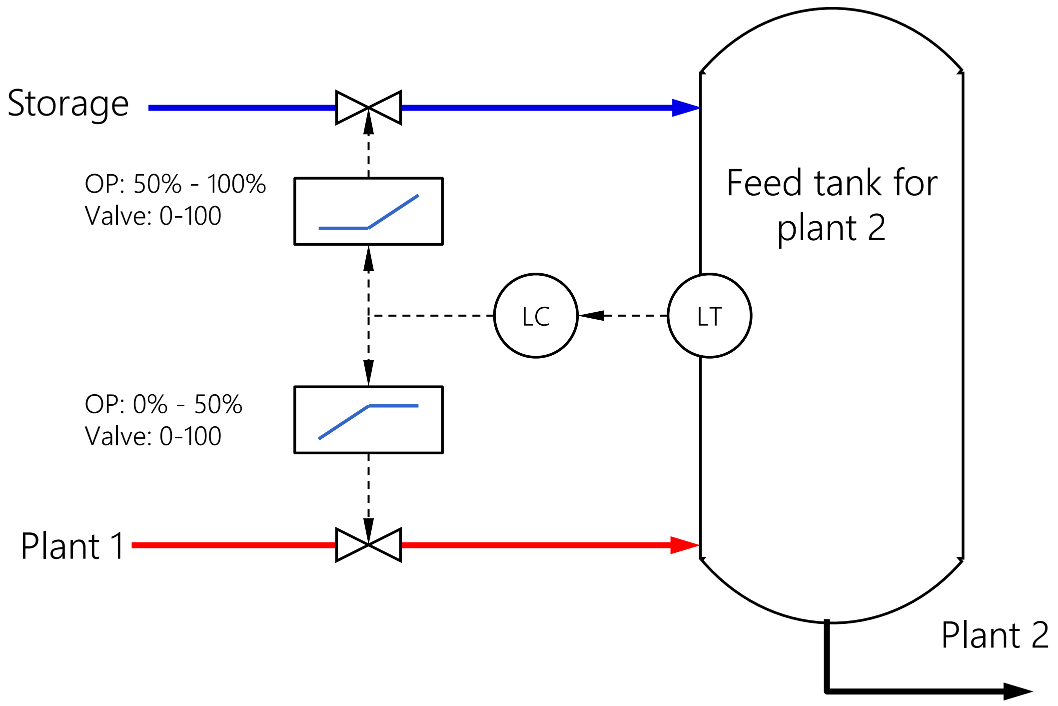
- Disturbance 1 is rather small.
  - It is enough that MV2 works.
- Disturbance 2 is larger.
  - Here both valves have to work.



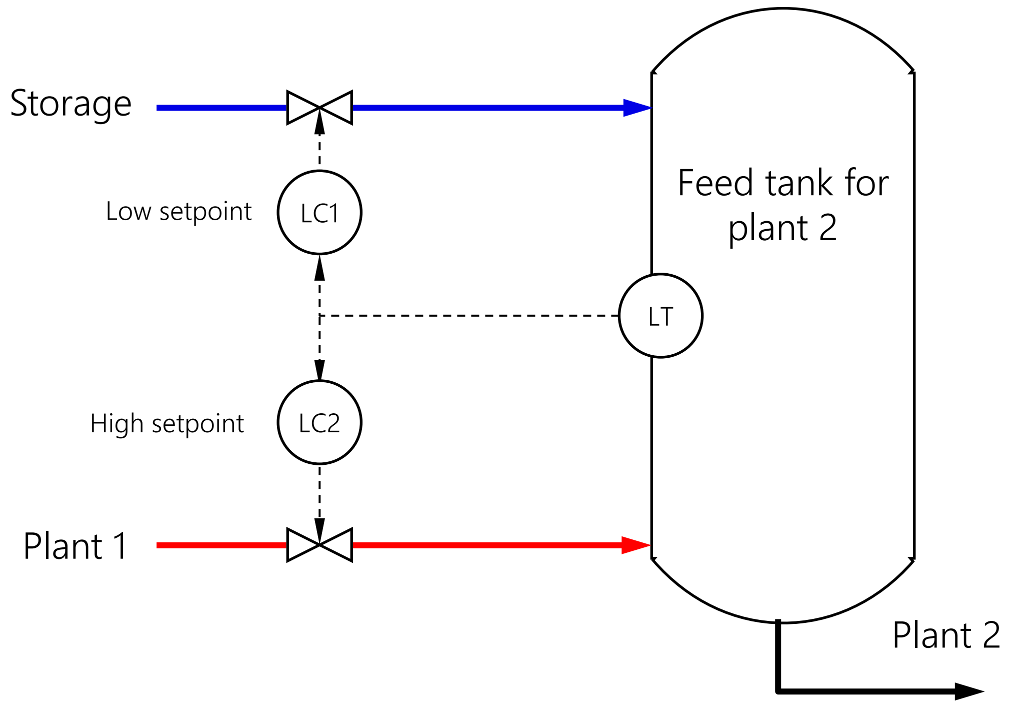
Level control with two feed streams

- In the example of a feed tank with two feed stream, we could have used split parallel control as well.

Split-range control

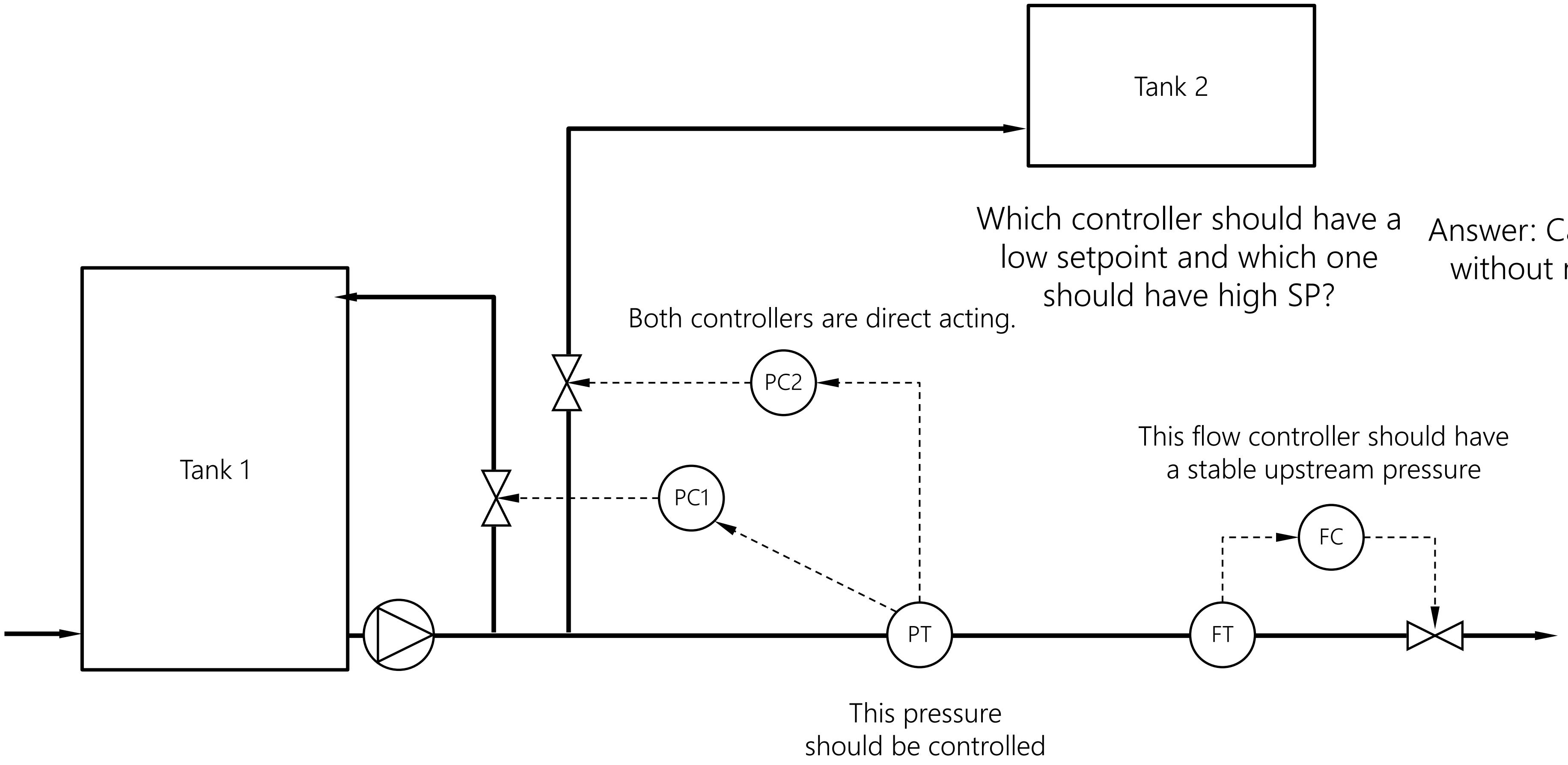


Split parallel control



Exercise: Pressure control, liquid system

Suggest a pressure control scheme based on split parallel control



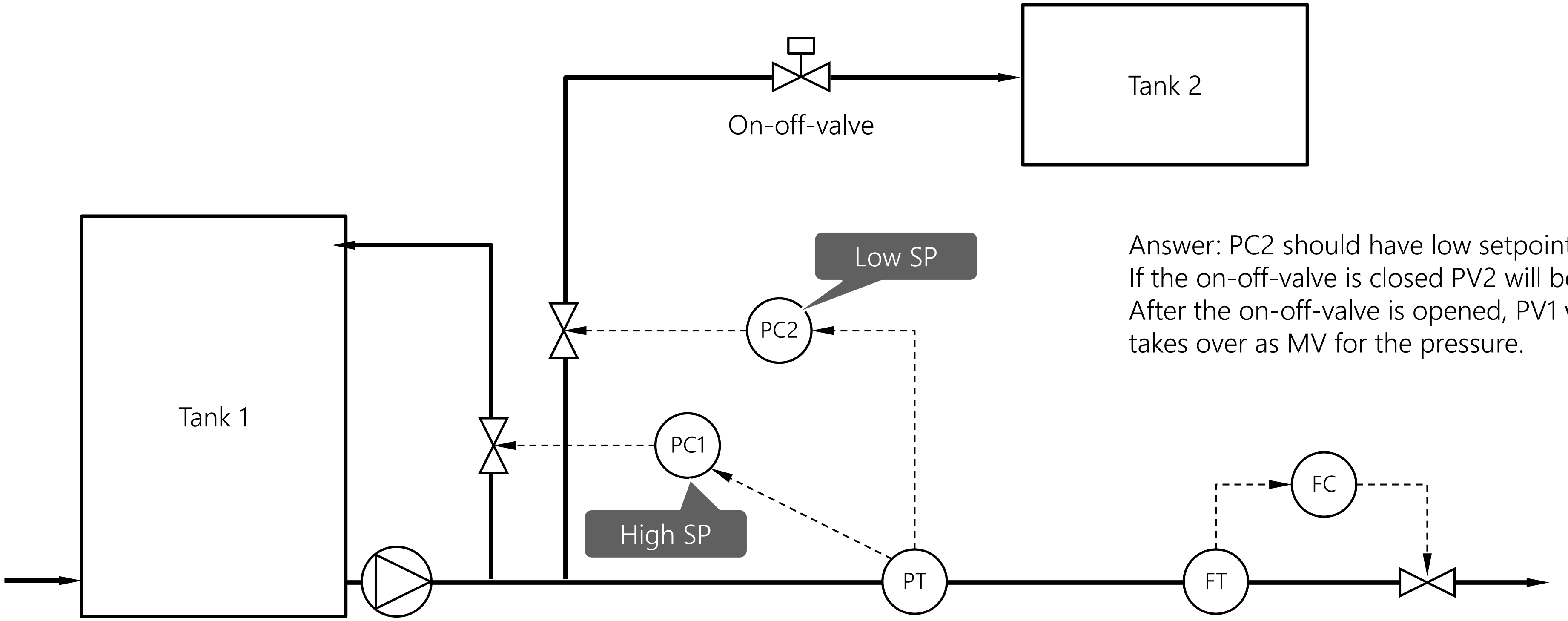
Which controller should have a low setpoint and which one should have high SP?

Answer: Cannot be determined without more specifications.

Cont'd: Pressure control, liquid system

New information: There is an on-off valve that operators open and close to fill tank 2.

Is it now possible to answer which controller should have high or low SP respectively?



Answer: PC2 should have low setpoint.
If the on-off-valve is closed PV2 will be 100% open.
After the on-off-valve is opened, PV1 will close to 0% and PV2 takes over as MV for the pressure.

Example: Cooling of steam condensate

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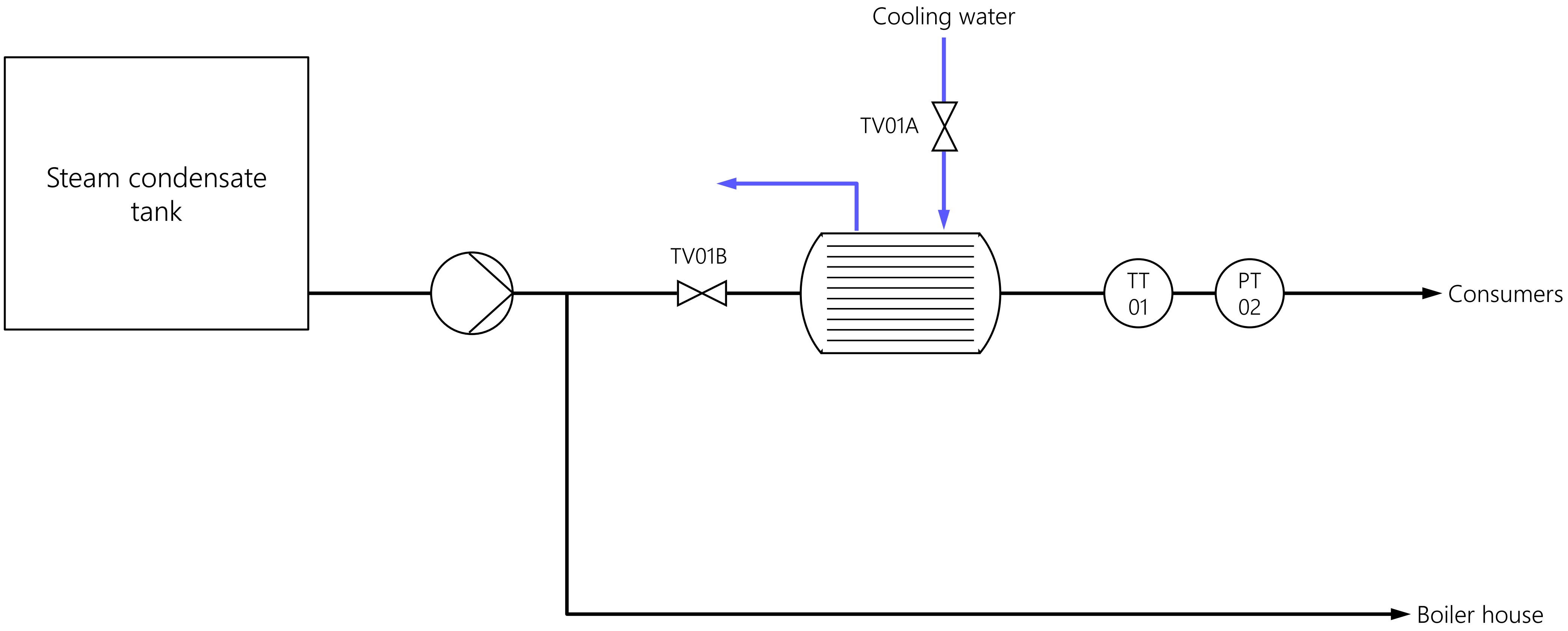


Control specifications

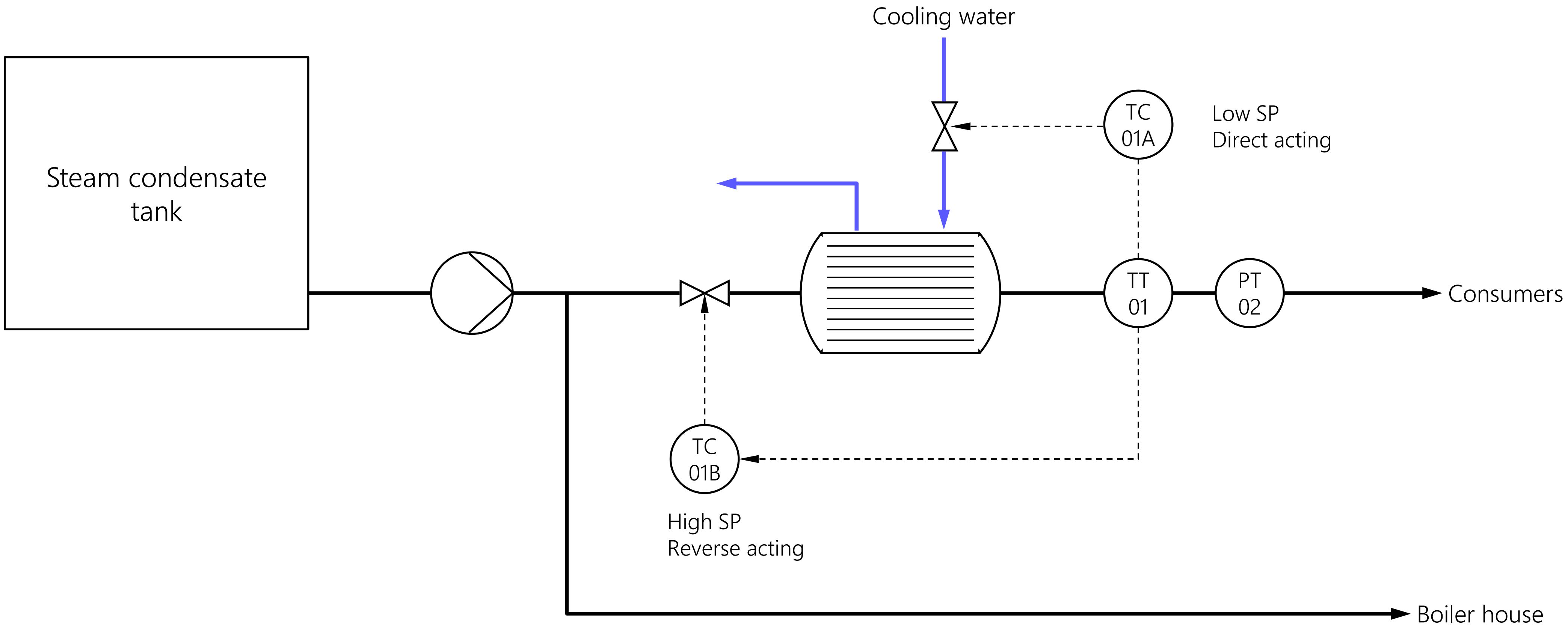
- The steam condensate coming in should be cooled to below 60 C.
- The condensate is used for cleaning with hoses, and other things.
  - It could be a risk for personnel if it is too hot.
- Whatever condensate is not used for flushing is sent to the boiler house
  - This means that there is no material balance aspect of this.
- The heat exchanger for cooling often has too low capacity.
- We can choose to reduce the flow, if there is not enough cooling.
- However, the pressure after the exchanger must not be too low,
  because then there is no pressure in the hoses on the fourth floor.

- All these requirements cannot always be fulfilled, I think.
- But we can make a control architecture that works fine
  as long as the equipment does not create an "infeasible" condition.

- This task arose from alarm management: there are frequent alarms for high temperature

Control architecture

Comment: This architecture does not solve the issue with too low pressure at times.

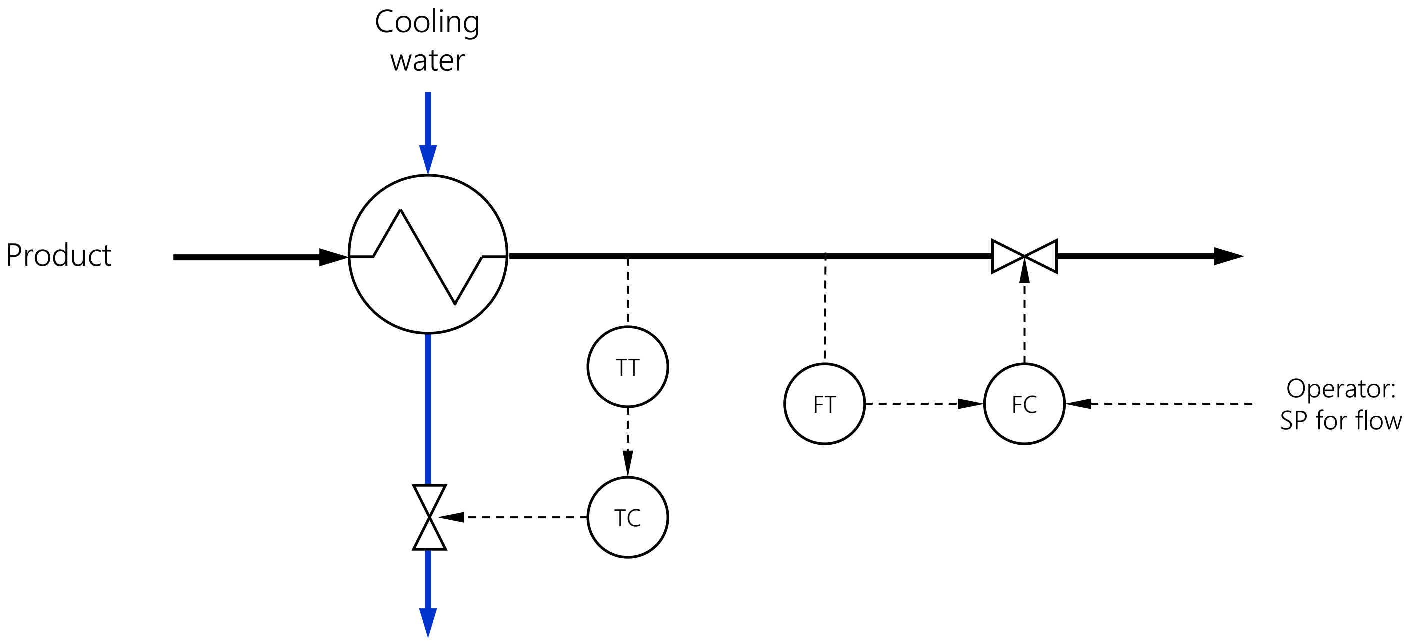


Maximizing control

Ex: Maximize flow in a heat exchanger (cooler)

Temperature must be kept at a given setpoint, for example 45 degrees.

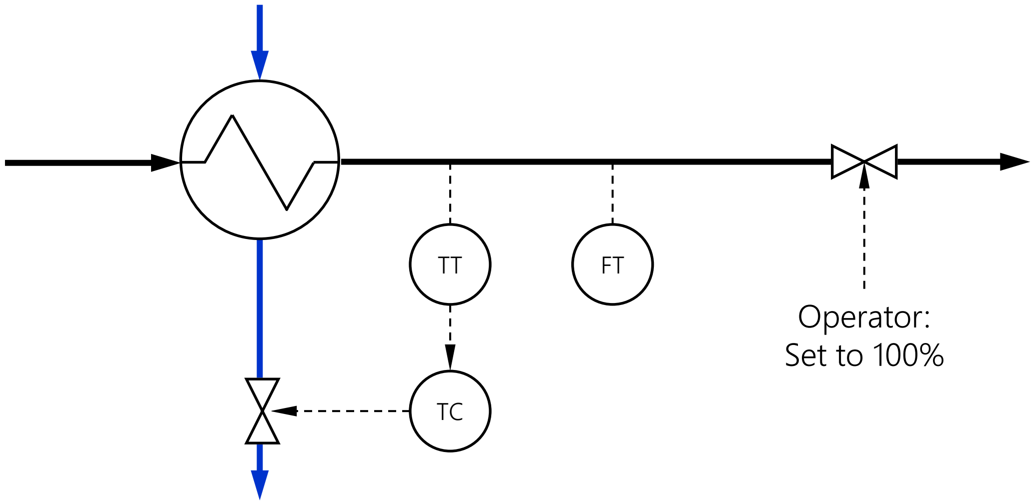
”Traditional” structure:



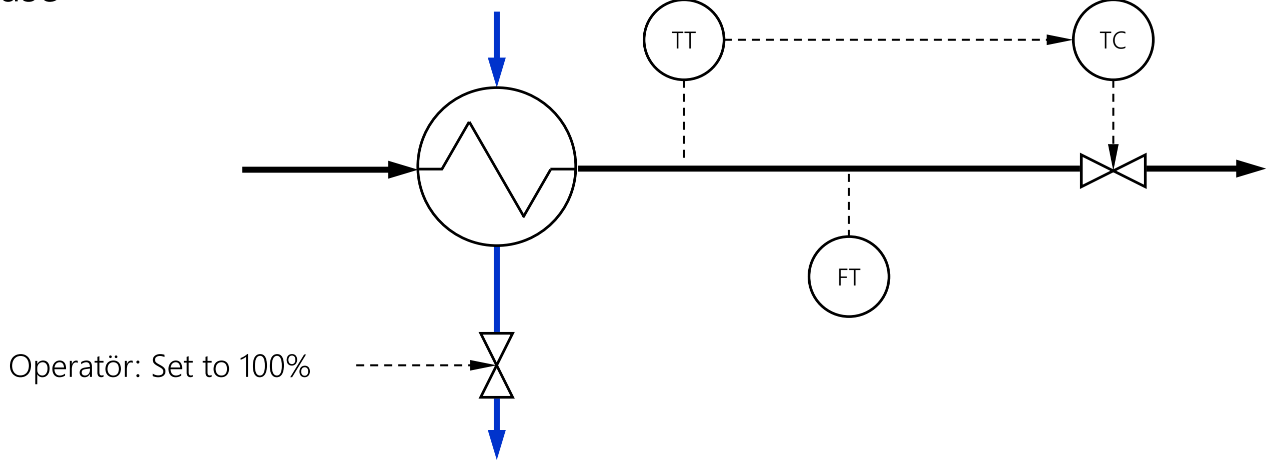
In this structure there is no automatic mechanism that guarantees maximal flow.

Cont; Maximize flow in a heat exchanger

If the flow valve is the limiting factor, this structure should be used.



If the cooling water valve is limiting, use this structure.



Maximizing control

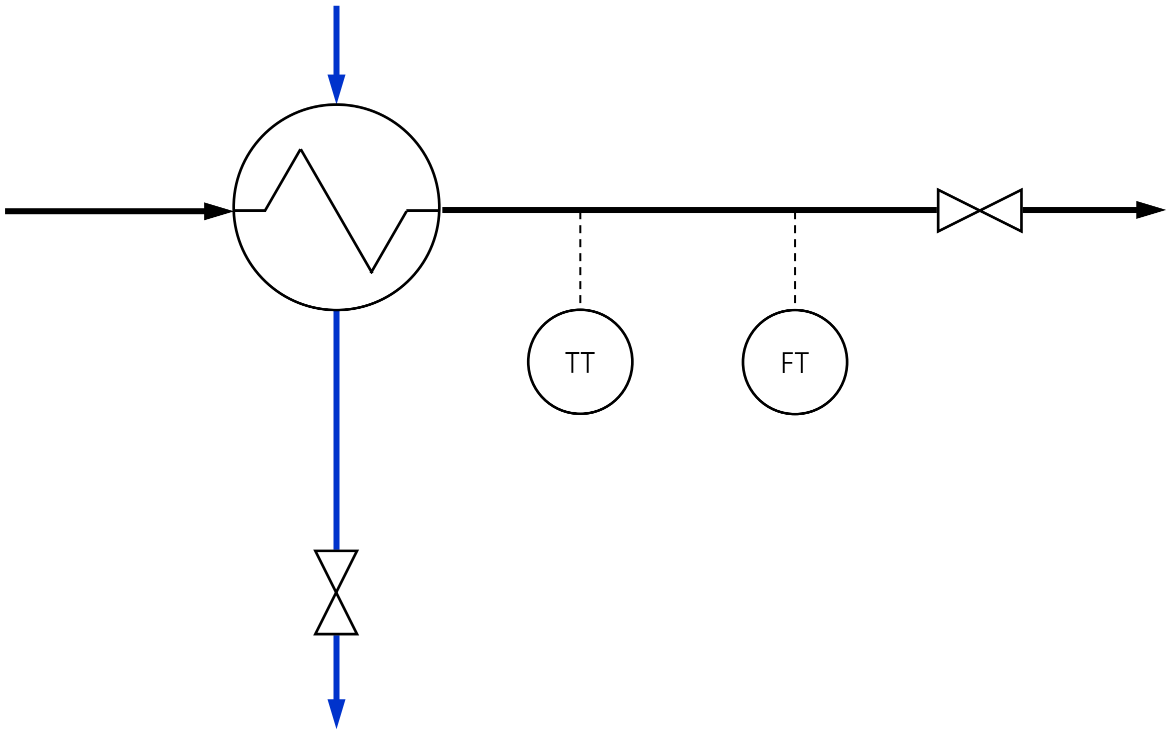
- Main idea: Use the throughput manipulator (TPM) as manipulated variable
  - Lock the bottleneck variable and manipulate the master flow to control what the bottleneck variable used to control.

- Example: Drying limited paper machine.
  Open steam valves 100% and manipulate machine speed
  to control the moisture content in the sheet.

Exercise: Handle both scenarios

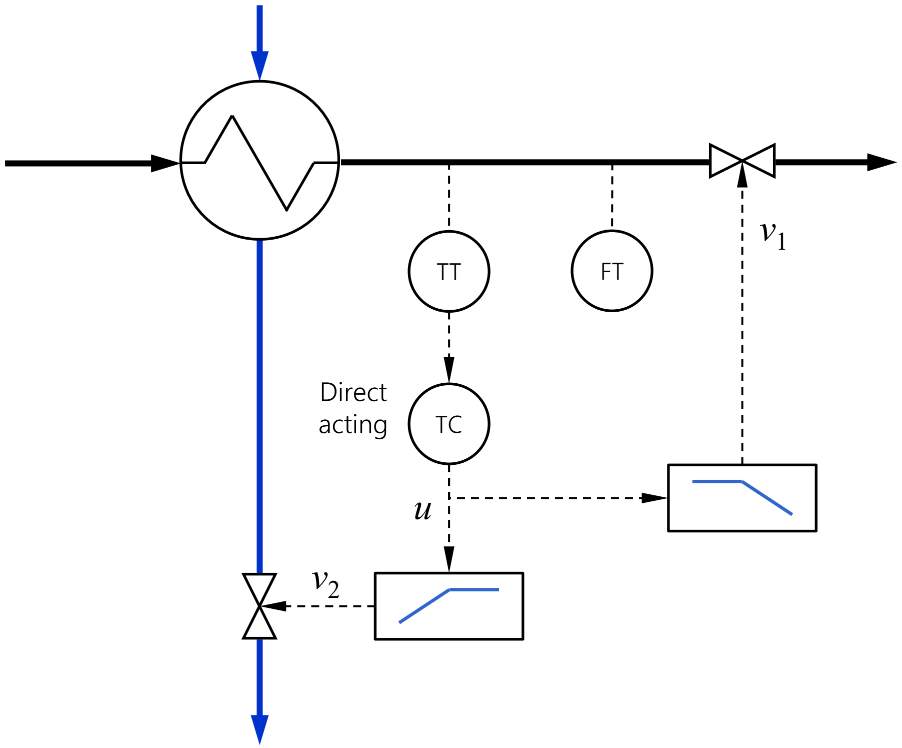
- It’s not uncommon that sometimes the flow valve and sometimes the cooling water valve is the limiting variable, e.g. due to variations in cooling water temperatur.

- **Exercise**: Find a structure that handles both cases.
  Clue: The structure relies on split-range.



Solution: Globally maximizing structure

- Temperature controller does the following:
  - Primarily: keep the production valve fully open and manipulate cooling water valve to control the temperature
  - Secondarily: keep cooling water valve fully open and manipulate the production valve to control the temperature

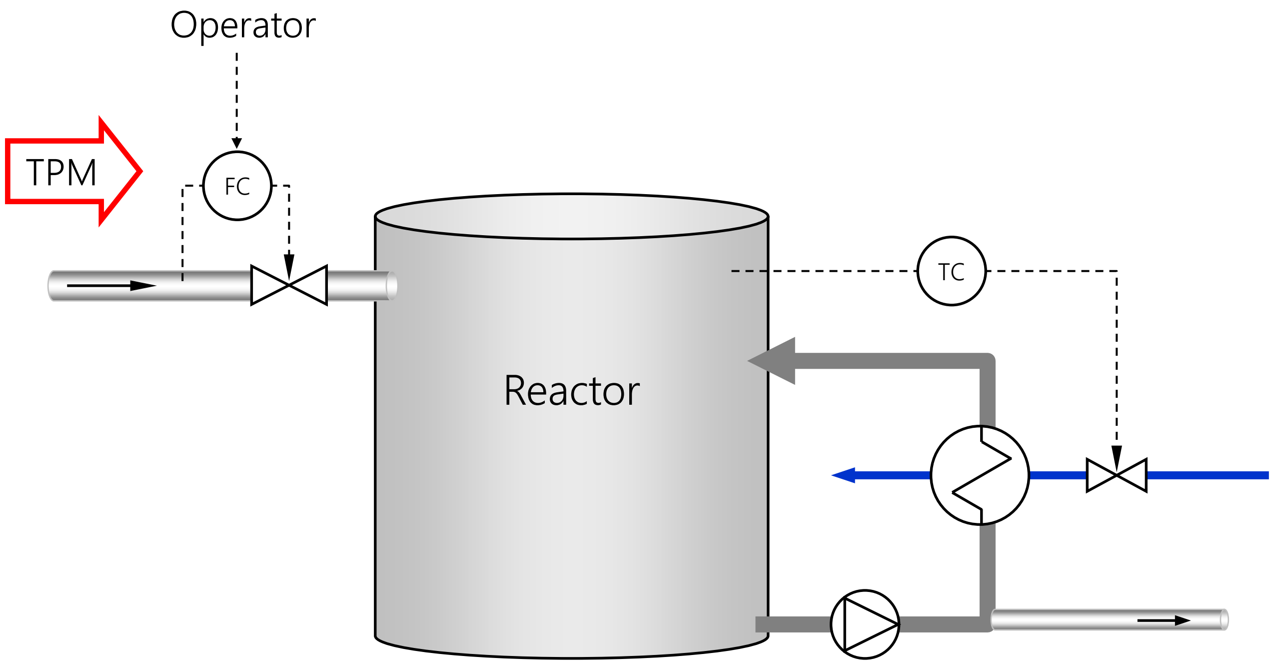


Variant of maximizing control

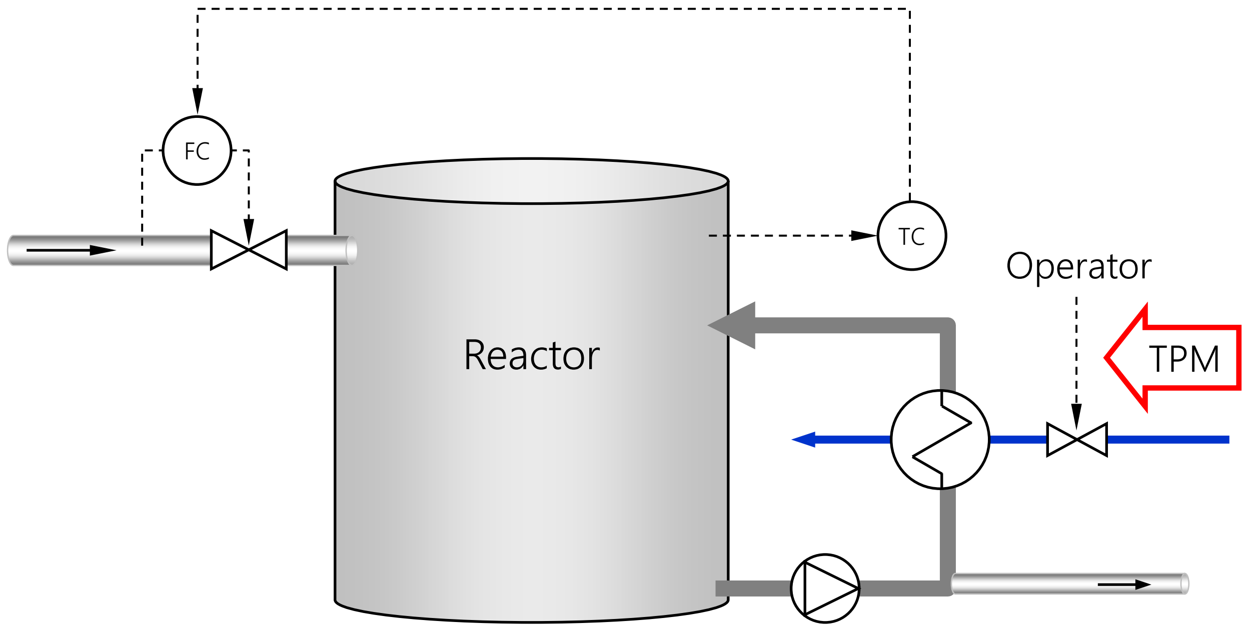
- Don’t switch manipulated variable, but slowly adjust SP for master flow so that controller output for the limiting process is at a suitable distance from its limit.
  - Example: Dryer limited paper machine. Machine speed automatically increased until the steam valves are 90% open, on average.
  - A **mid-ranging controller** can be used for this.

Example: Cooling capacity limited reactor

Normal operation



Cooling limited Rx: Maximizing control 1



Maximizing control 2: Use VPC

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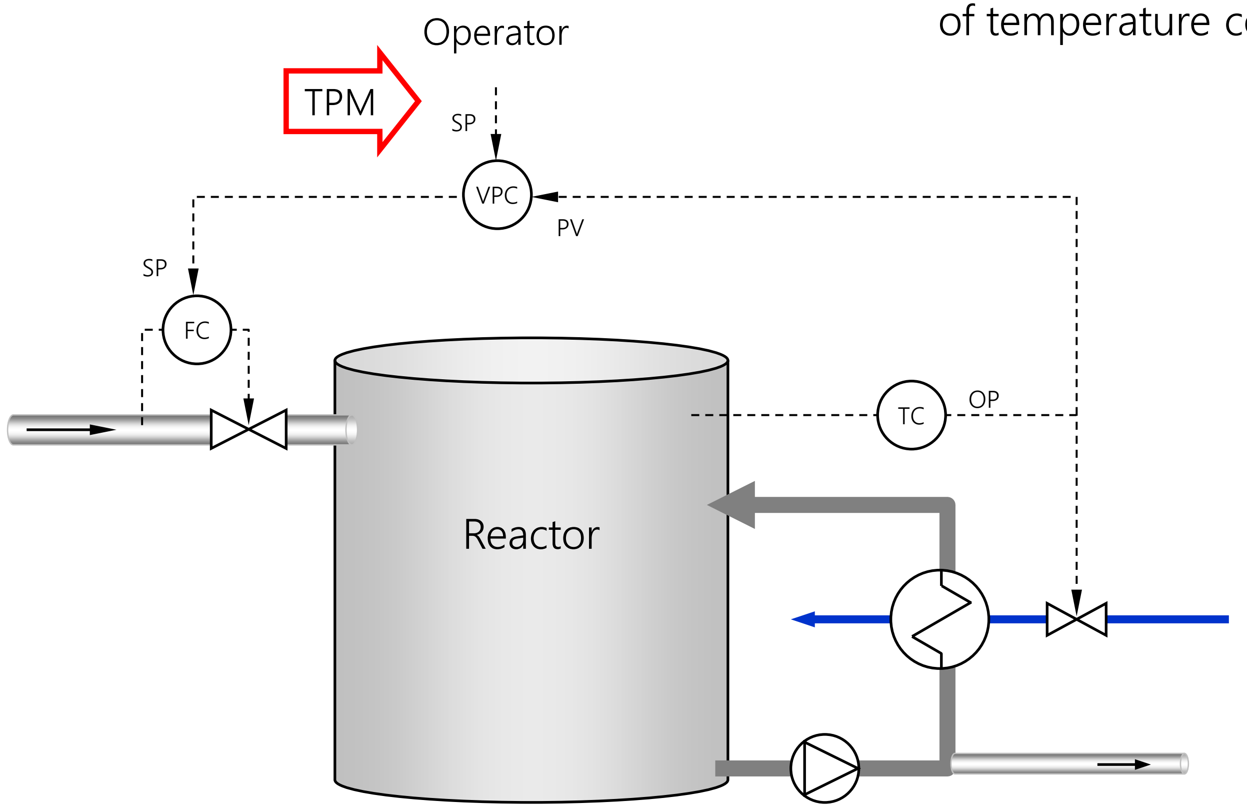
Not as high production rate as with solution 1, but better performance of temperature controller.

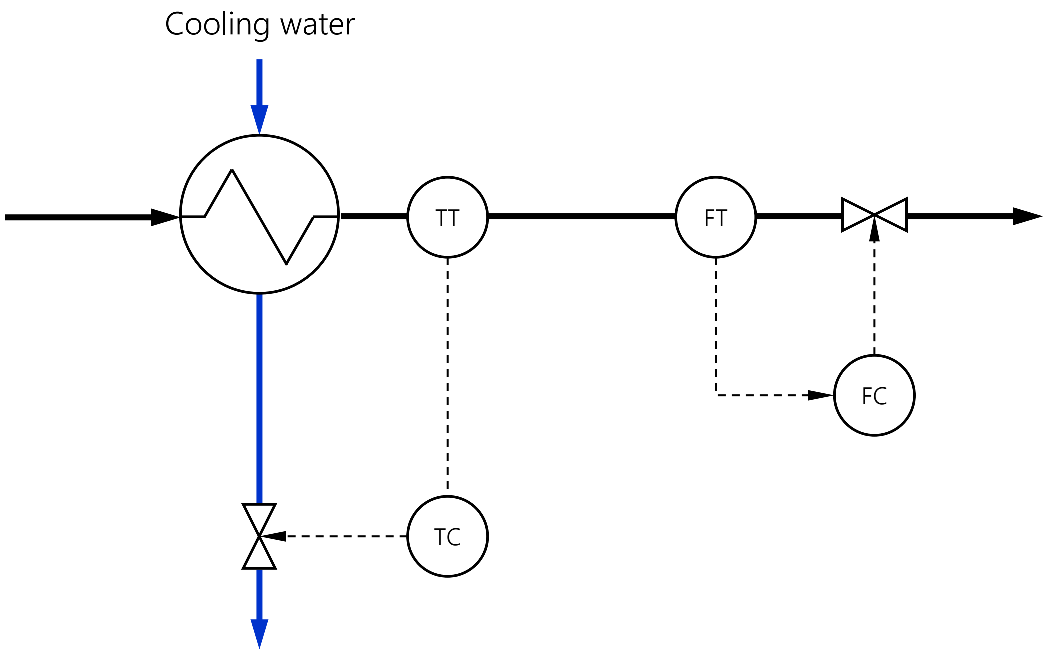


Exercise: HEX flow maximization with exact temp control

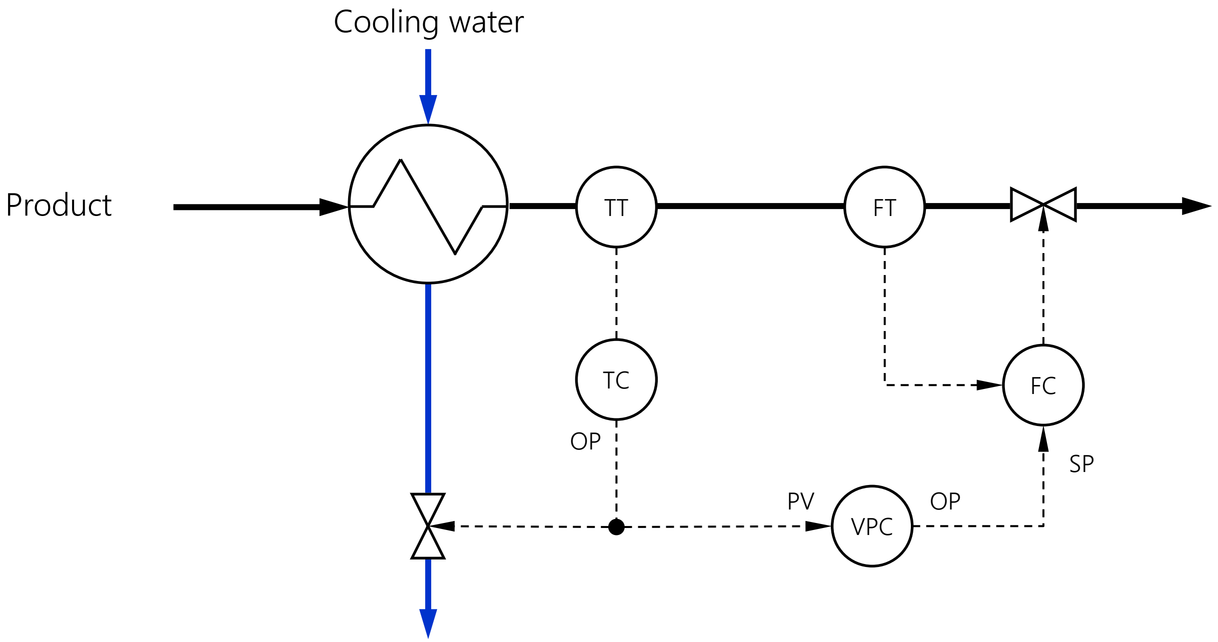
Cooler example.
Design a control structure where the flow can be set close to its max limit,
but only the cooling water valve is manipulated for controlling temperature.

Start from this structure:

Product

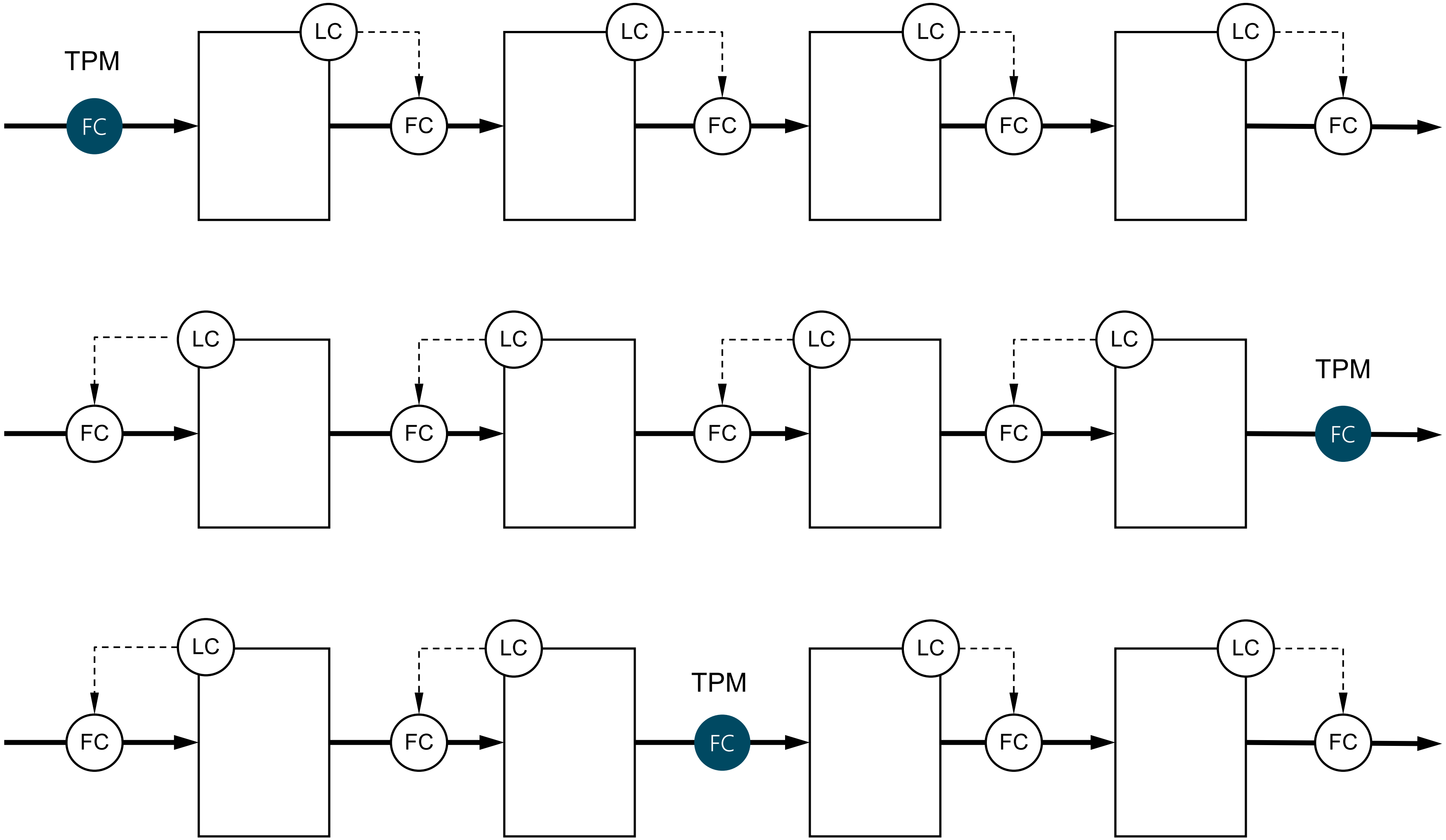


Solution: VPC for maximizing control of HEX



Bidirectional control

Common scenario: Bottleneck/TPM moves over time



”Bidirectional control” automates when TPM moves

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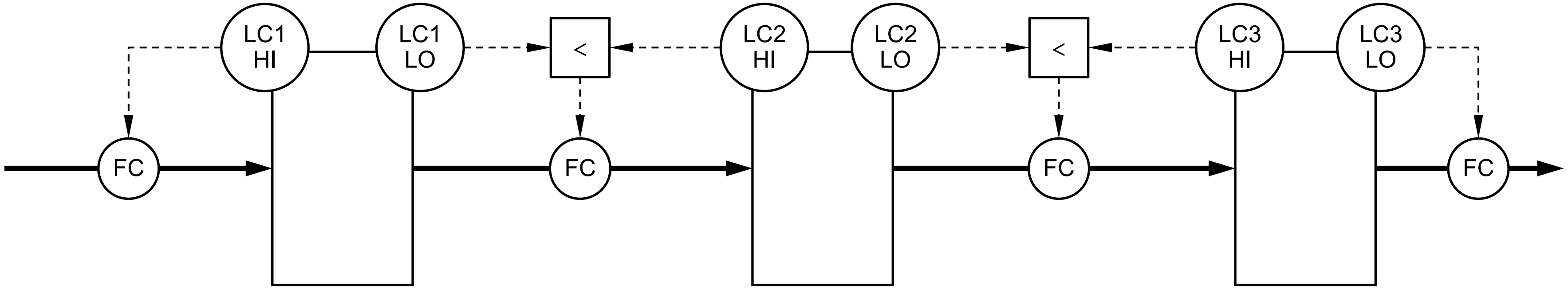
- Often, there is a different bottleneck in different operating points
- Sometimes you want to handle this automatically. Solution is **bidirectional control**.
- Can be implemented in different ways. Simplest: dual level controllers with different setpoints and min-selector.



HI-controllers: reverse acting
LO-controllers: direct acting

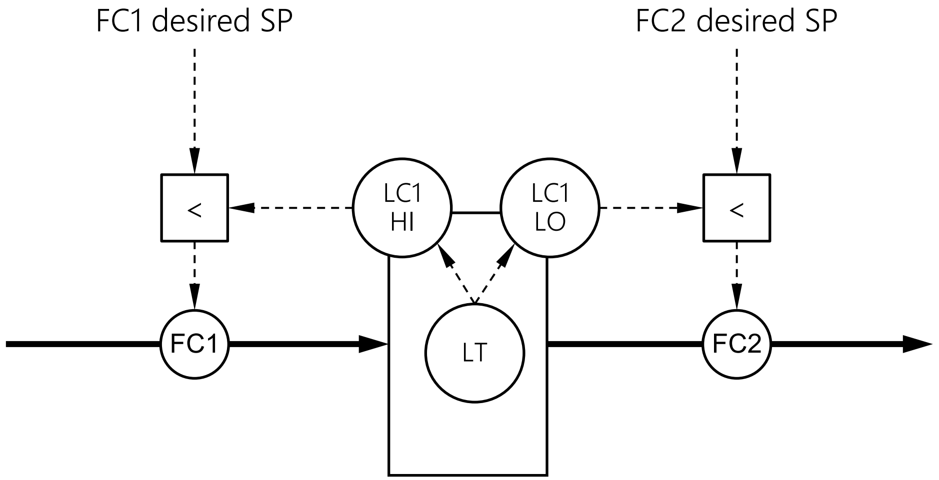
A simple bidirectional inventory control structure with optimal buffer management

Cristina Zotică[a], Krister Forsman[b], Sigurd Skogestad[a,]

[a] *Department of Chemical Engineering, Norwegian University of Science and Technology (NTNU), 7491, Trondheim, Norway*
[b] *Perstorp Specialty Chemicals, 284 80, Perstorp, Sweden*

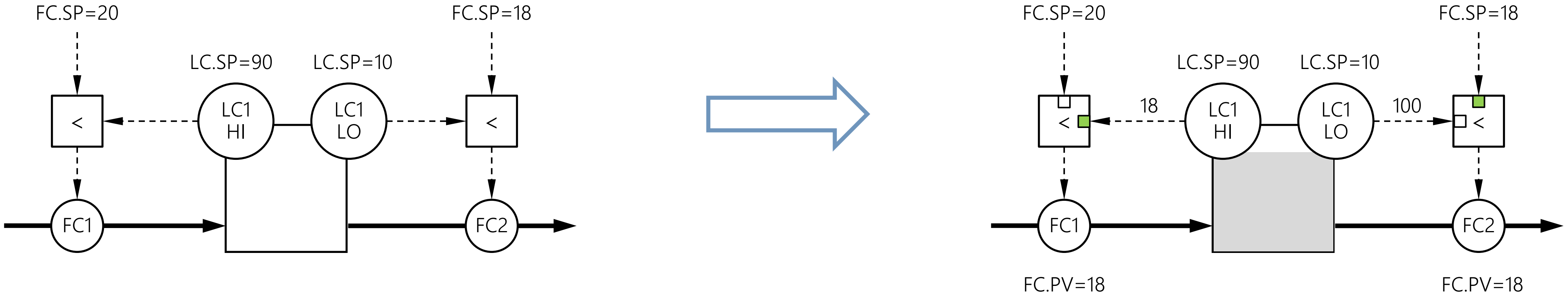
Basic element of bidirectional control

- A system of bidirectional controllers is made up from the following basic building block.

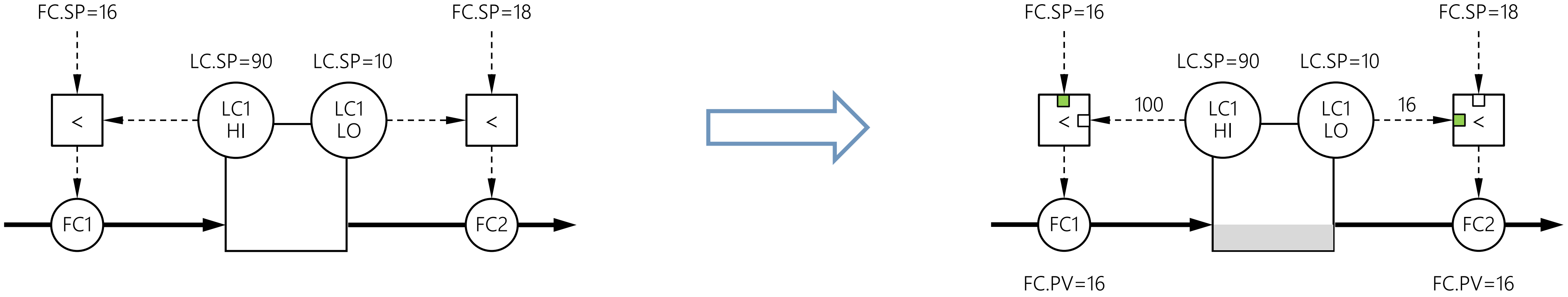


HI-controller: reverse acting
LO-controller: direct acting

Examples



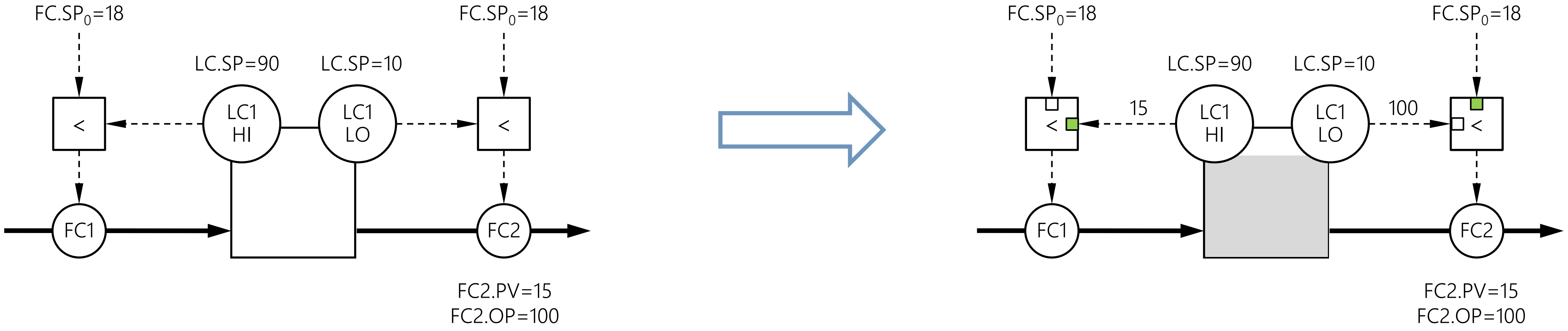
FC.SP=20
FC.SP=18
LC.SP=90
LC.SP=10
LC1
HI
LC1
LO
<
<
FC1
FC2
FC.SP=20
FC.SP=18
LC.SP=90
LC.SP=10
18
100
LC1
HI
LC1
LO
<
<
FC1
FC2
FC.PV=18
FC.PV=18



FC.SP=16
FC.SP=18
LC.SP=90
LC.SP=10
LC1
HI
LC1
LO
<
<
FC1
FC2
FC.SP=16
FC.SP=18
LC.SP=90
LC.SP=10
100
16
LC1
HI
LC1
LO
<
<
FC1
FC2
FC.PV=16
FC.PV=16

More examples

3



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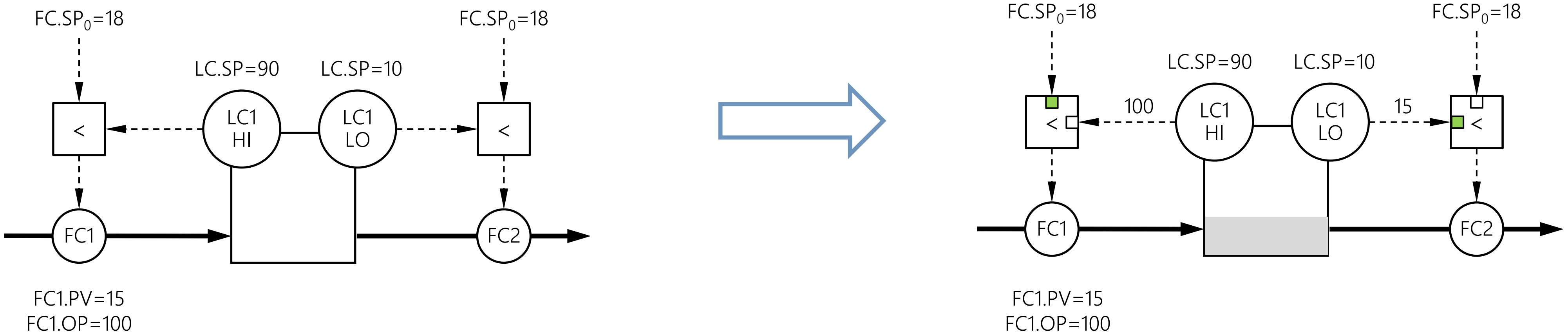
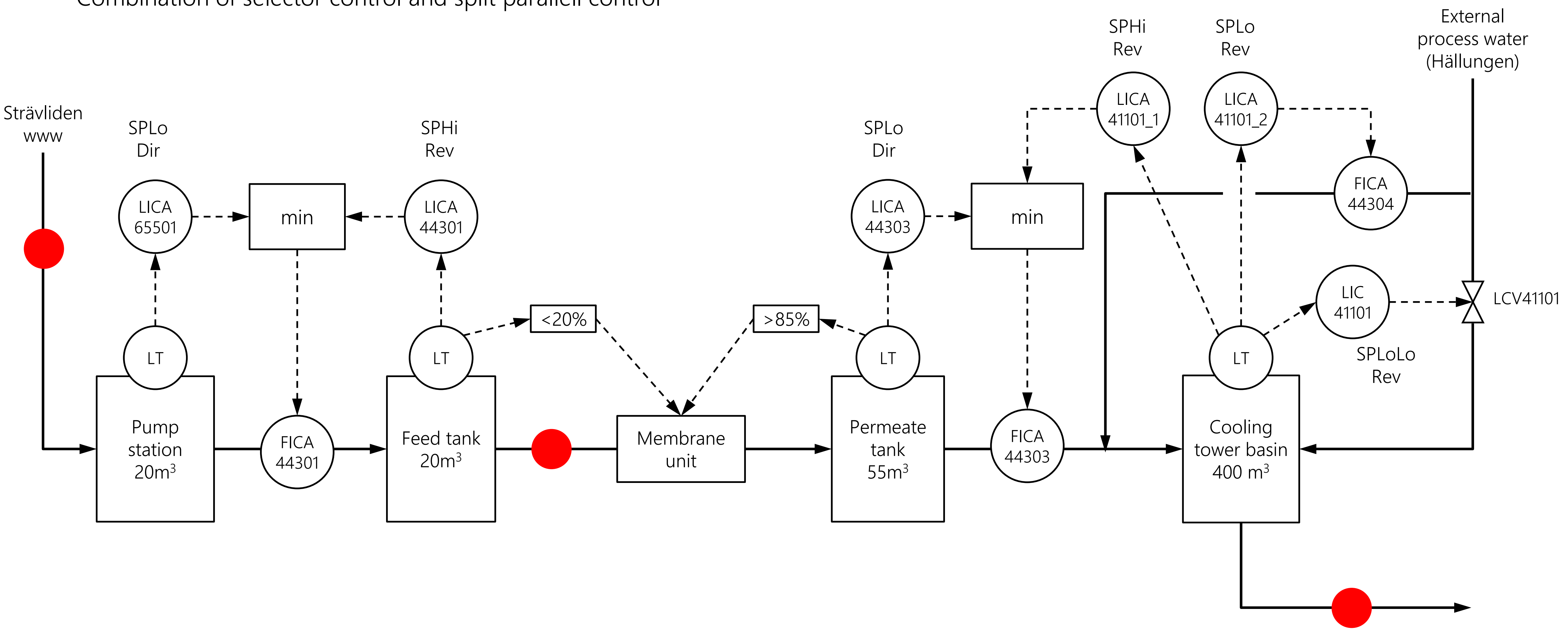


Example: New raw water system in Stenungsund

Combination of selector control and split parallell control

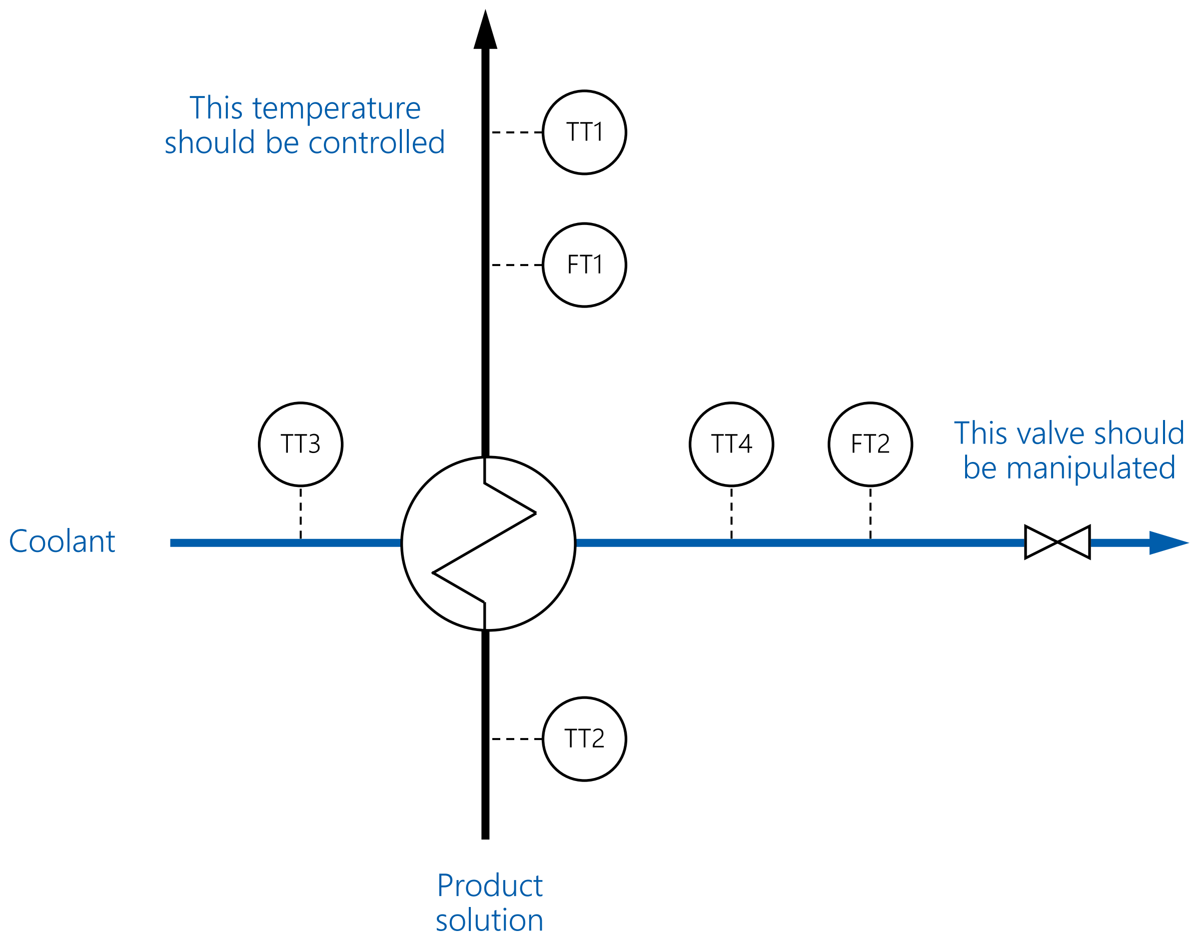


Heat exchanger control

How to control a heat exchanger?

- There are lots of control structures for heat exchangers. The choice depends on
  - Which instrumentation is present
  - Which disturbances that are dominating
- The exercise below gives guidelines for the choice of structure.

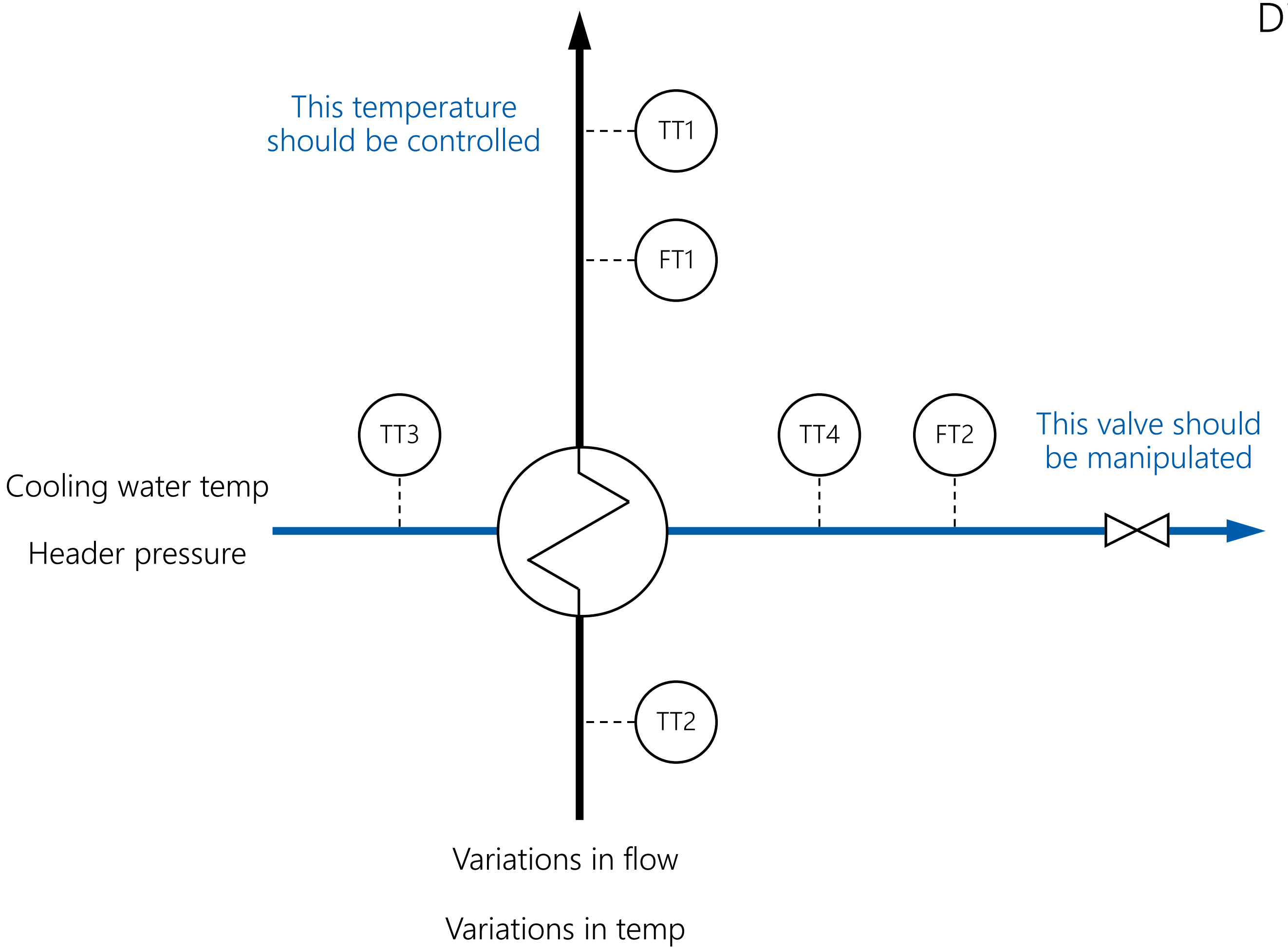
Heat exchanger with generous instrumentation



Exercise: Which disturbances could occur in this process?

HEX: Typical disturbances

Exercise: Suggest different control structures
Discuss which disturbances are best handled by which structure.



HEX: The simplest control structure

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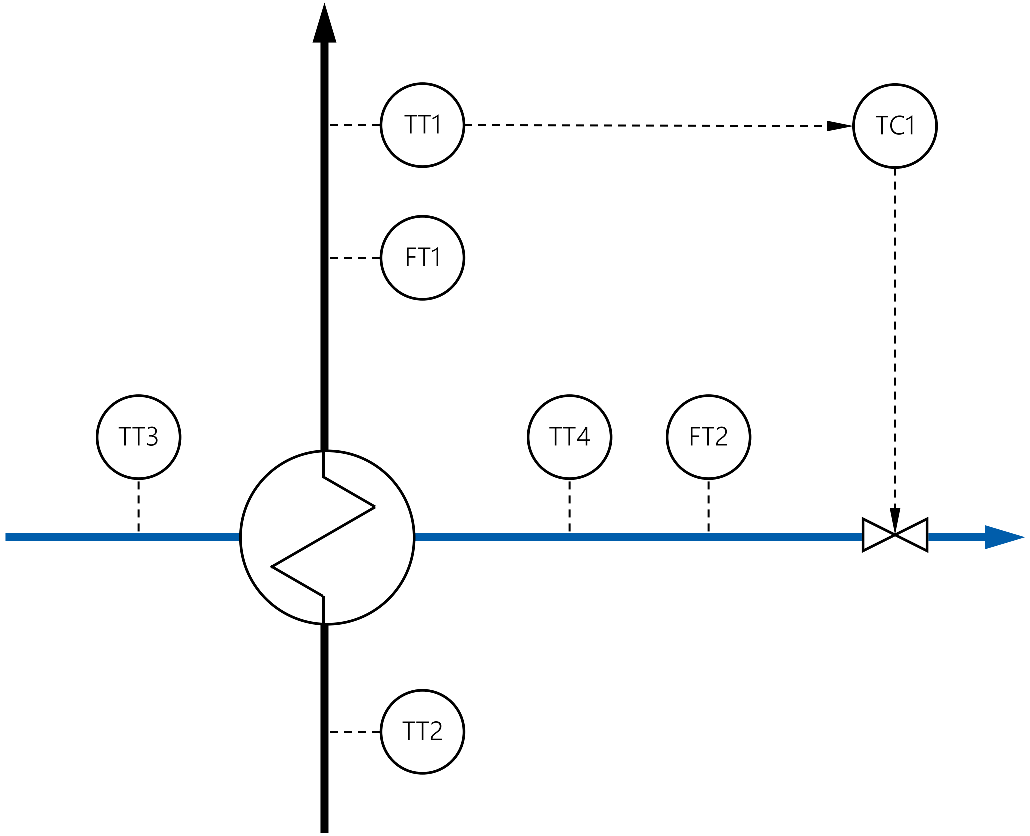
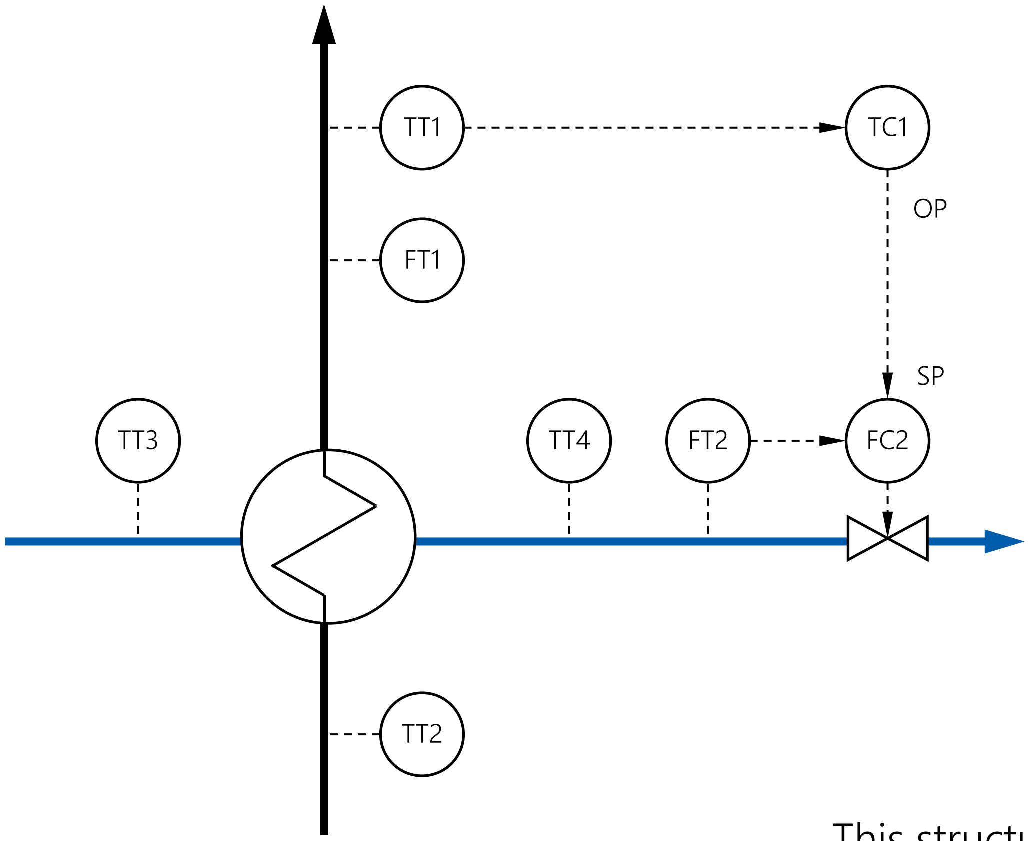


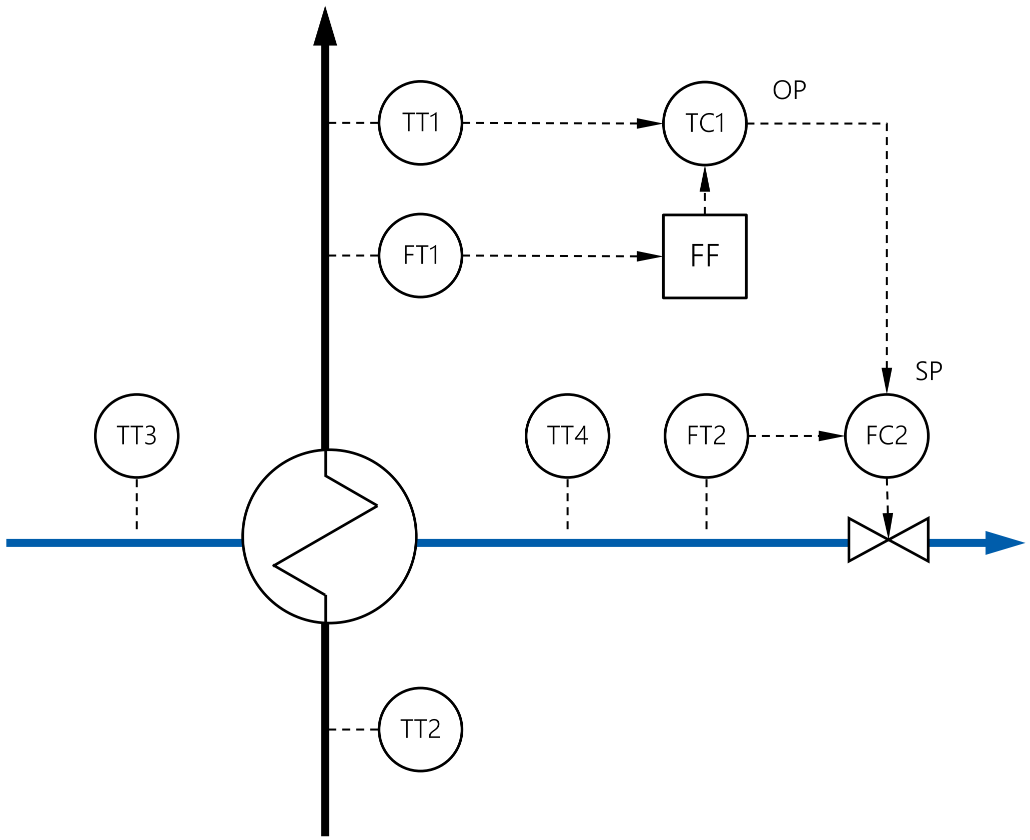
Flow in cascade

Which disturbances are better suppressed by this structure, compared to SISO control?



This structure is good at managing pressure variations in the cooling water header.

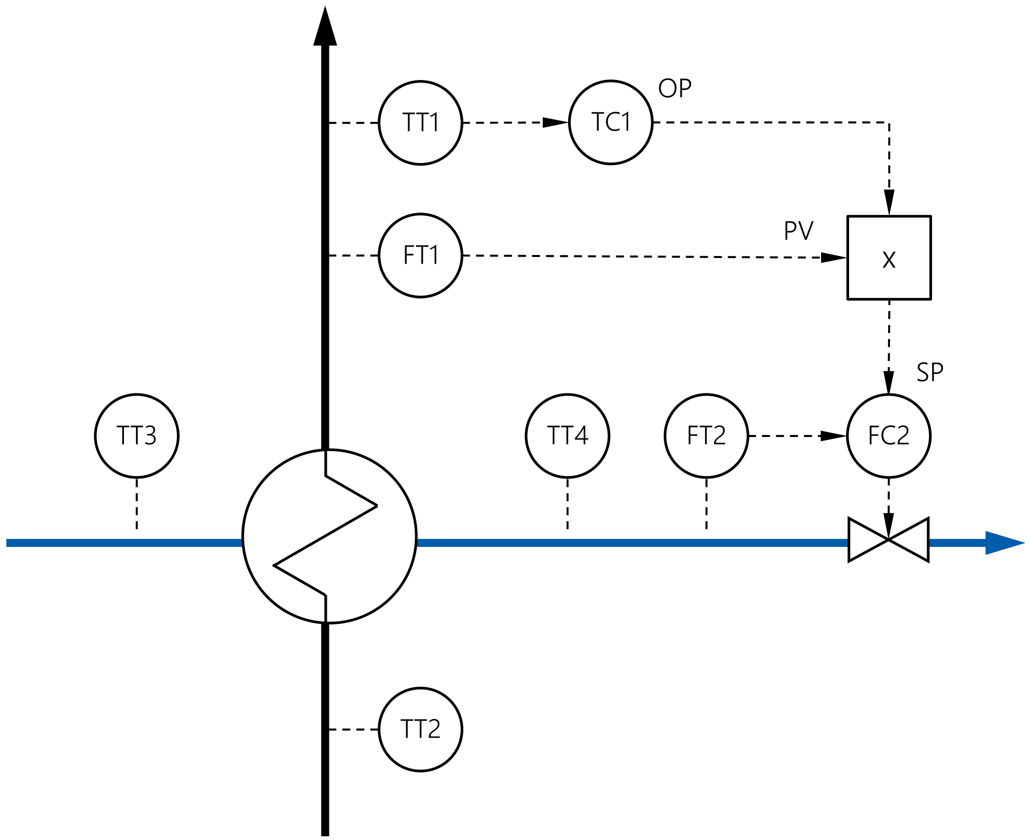
Feedforward from product flow



In which way is this structure superior to the previous one?

Answer: It handles variations in product flow in a better way.

Ratio control: Cooling water vs media flow



Advantage compared to feedforward?

"Variable feedforward gain"
Can be motivated by variations in temp (energy content, enthalpy) in the product stream and/or cooling water
or by heat exchanger nonlinearity (effect of flow change varies with operating point)

Power as manipulated variable: Solves everything

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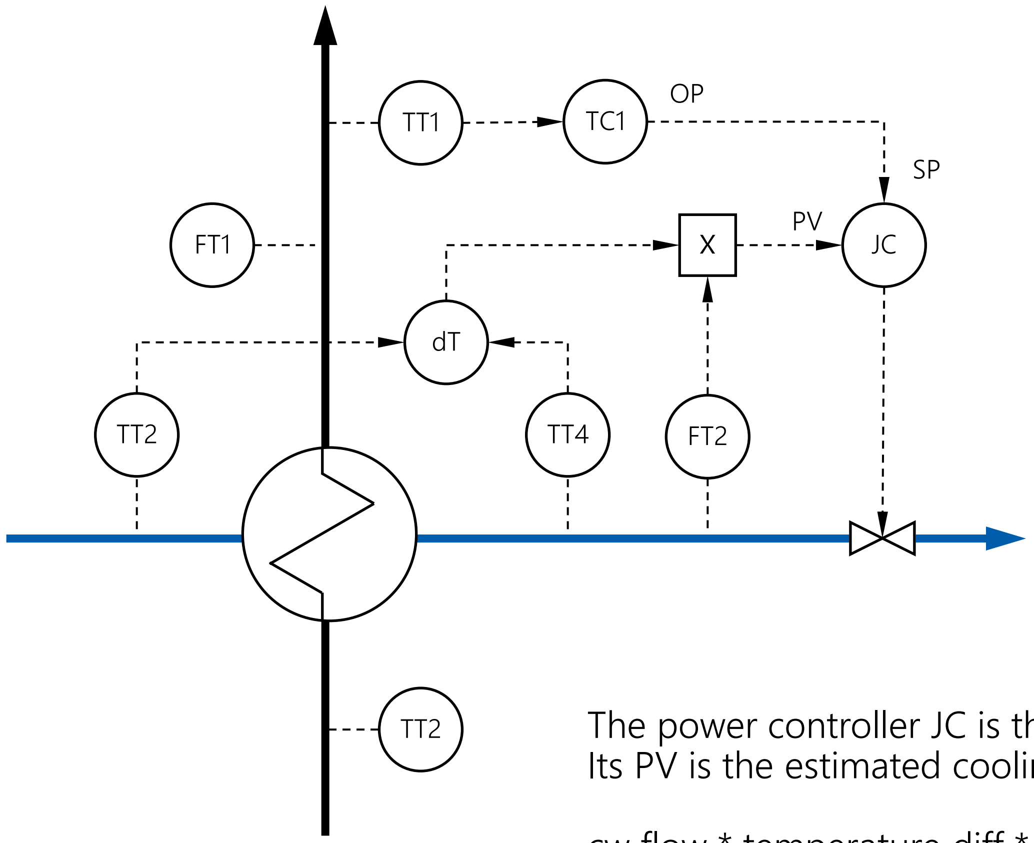

The power controller JC is the slave to the temperature controller.
Its PV is the estimated cooling power ("soft-sensor").

cw flow * temperature diff * heat capacity = power.

Reactor power control: significant improvement

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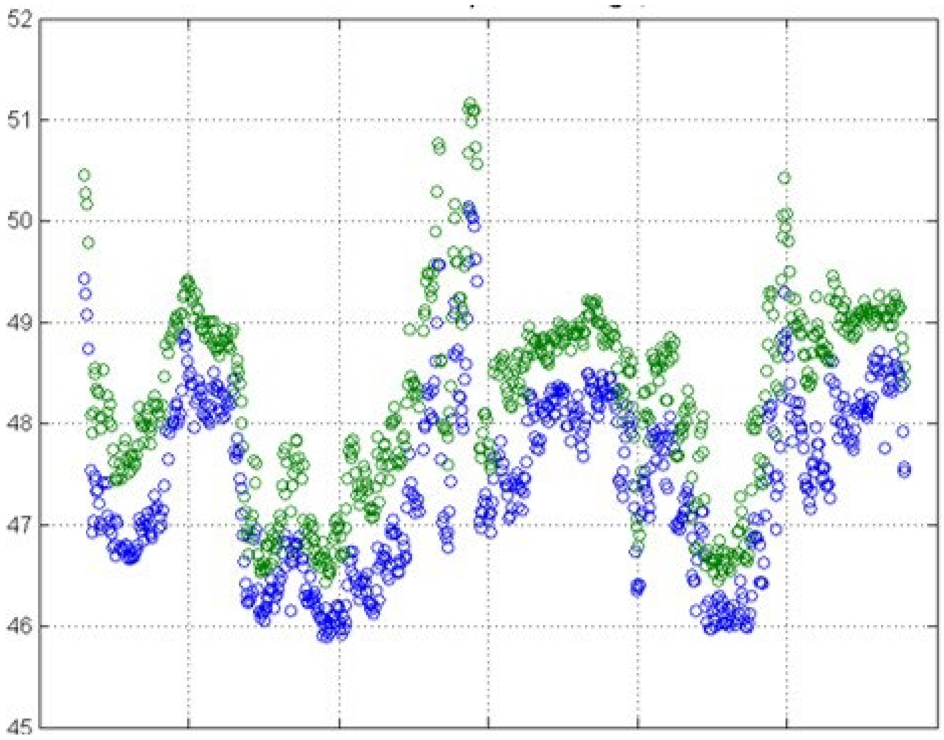
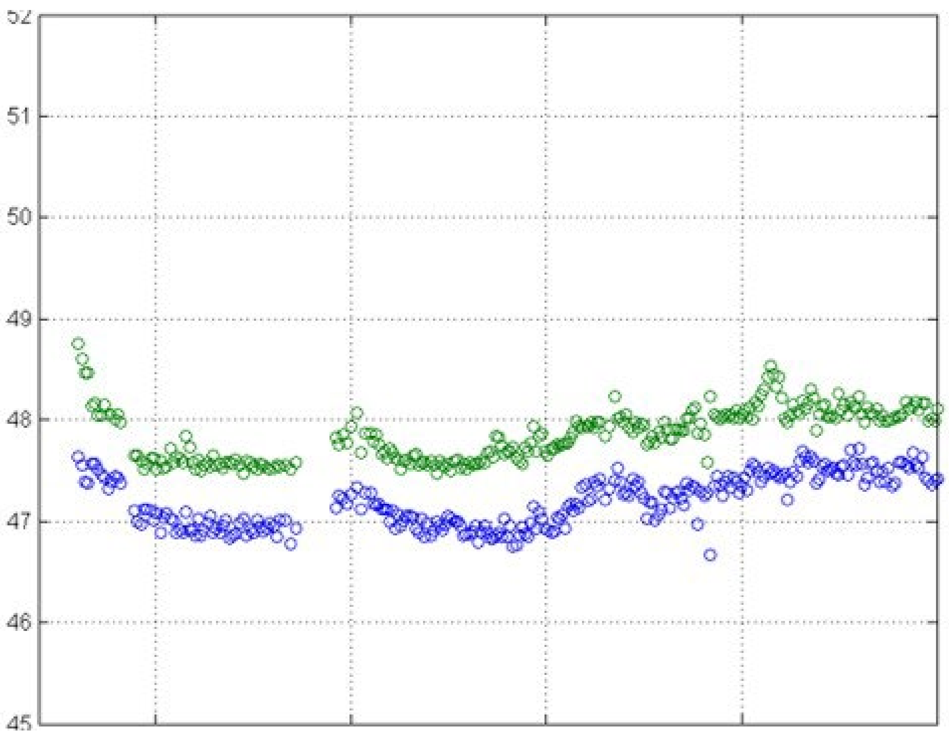
Directly manipulated cooling water valve

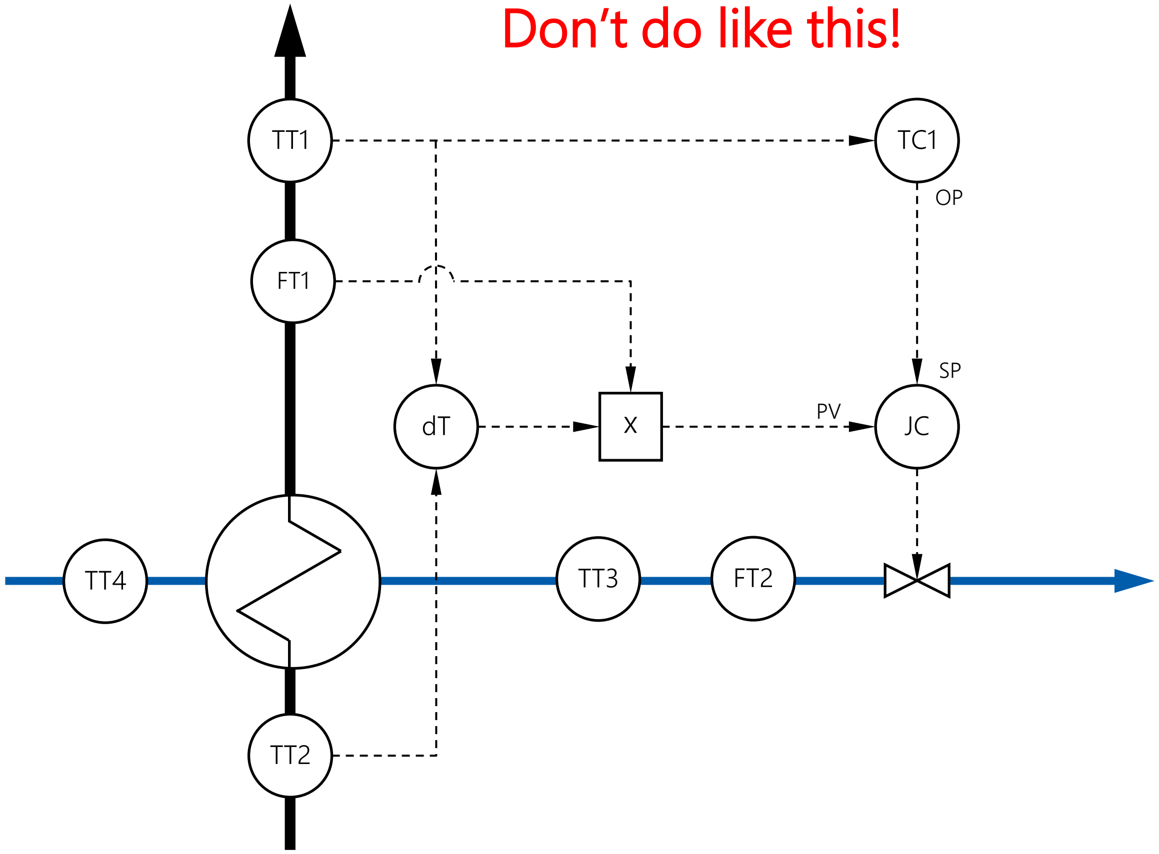


Power control



Don’t estimate power on the non-manipulated side

Don’t do like this!



This soft sensor also gives an estimate of the power.

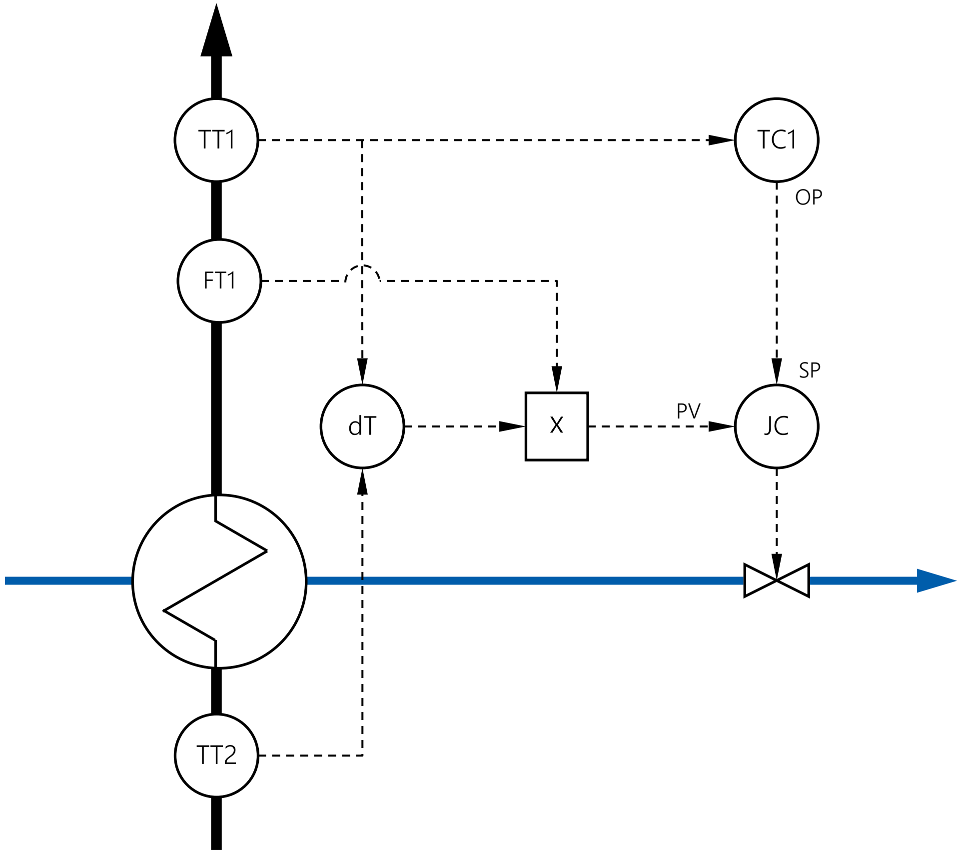
But **often this solution works significantly worse**.
See next slide.

Recommendation:
Use temperatures and flow on the manipulated side to estimate the power.

It is however more common to have full instrumentation on the product side than the cooling side.

The instrument that is most often missing is FT2 in the figure.

Why power should not be estimated on the hot side



Several reasons why this is worse than estimating the power on the manipulated side

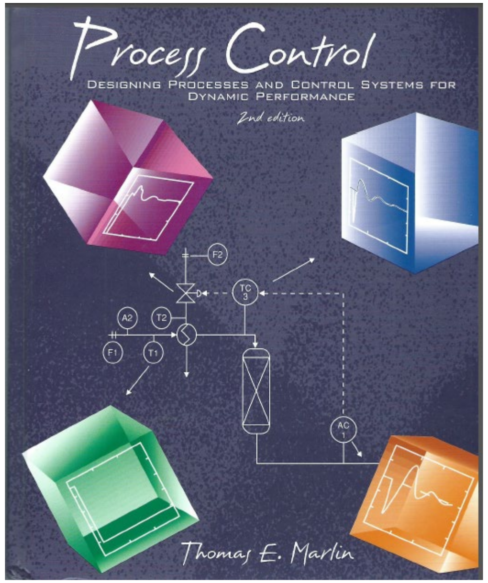
1: The response from MV to PV is much slower if JC.PV is estimated on the hot side because the power controller manipulates the cooling water valve. A slow process is harder to control than a fast process.

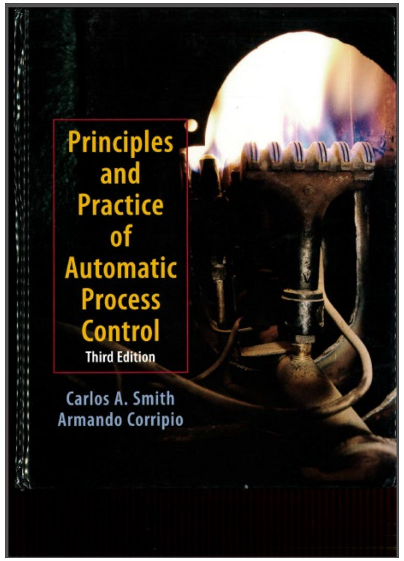
2: Assume that the incoming temp on the hot side (TT2) suddenly increases. Then the estimated power also increases, initially, even though the actual HEX power has not changed.
We get a control deviation against JC.SP and JC reacts by closing the cooling water valve, but it should do the opposite.

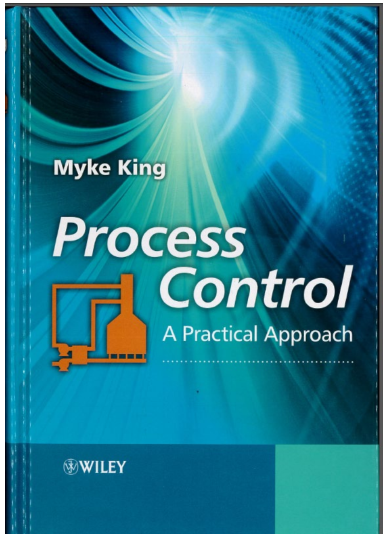
Recommendation:
Use temperatures and flow on the manipulated side to estimate power.

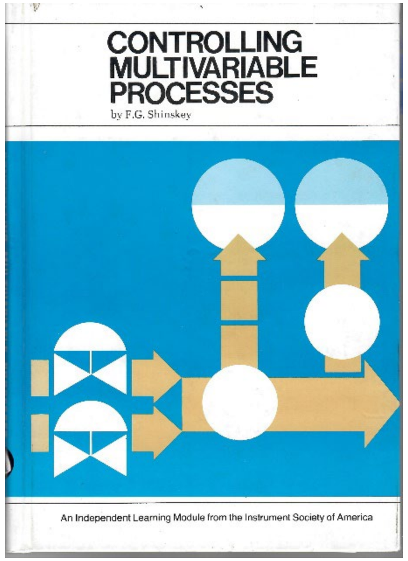
Not clear if you can call JC a slave controller here,
since TT1 is used both by the JC and the master controller TC1.

Some textbooks


Process Control
Designing Processes and Control Systems for Dynamic Performance
2nd edition
Thomas E. Marlin


Principles and Practice of Automatic Process Control
Third Edition
Carlos A. Smith
Armando Corripio


Myke King
Process Control
A Practical Approach
WILEY


CONTROLLING MULTIVARIABLE PROCESSES
by F.G. Shinskey
An Independent Learning Module from the Instrument Society of America