Model Predictive Control including integral transfer functions

SIMATIC PCS 7

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Table of Contents

1 Preface

Objective of the Application

The area of application of the model predictive controller provided in SIMATIC PCS 7 (function block ModPreCon and MPC respectively) is restricted by the following fact: The control algorithm only works for stable processes with a step response settling to a constant value in a finite time horizon.

If the process is not stable or shows an integral action (e.g. level control, position control), the respective sub transfer functions have to be stabilized by slave controllers. The objective of this application note is to describe how to proceed in such cases.

This application note is an extension of the application note “Multivariable Model Predictive Control – the Distillation Column as an Application Example”, containing basic information how to apply the MPC.

The application example considered here shows an MPC with two manipulated and two controlled variables in combination with a simulated process of which the second main control loop shows integral action.

Main Contents of this Application Note

The following issues are discussed in this application:

- The stabilization of unstable processes using P(ID) controllers
- The integration of a subordinated PID controller in an MPC application
- Simulation example

Validity

… valid for PCS 7 V7.0 SP1 or later versions.

2 Introduction

2.1 Basic Principles of Model Predictive Control

A general overview of model predictive control is provided by the White Paper "How to improve the Performance of your Plant using the appropriate tools of SIMATIC PCS 7 APC-Portfolio?"

https://pcs.khe.siemens.com/efiles/pcs7/support/marktstudien/WP_PCS7_APC_EN.pdf

The application note including the basic principles of the MPC can be found here:

http://cache.automation.siemens.com/dnl/zI/zIzMzM1MwAA_37361208_Tools/37361208_MPC_en.pdf

2.2 Stable and Unstable Control Loops

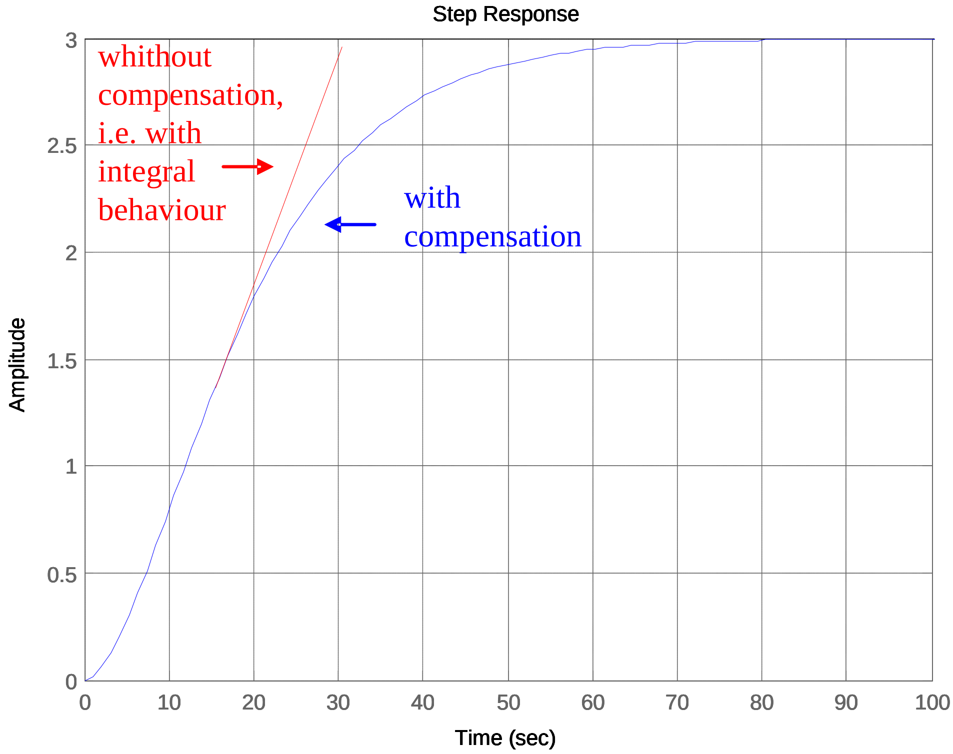
Most of the control loops in process plants show a stable behaviour - after a stepwise change in the manipulated variable the control variable shows a transient behaviour reaching a new steady state after some time. The controlled process is "stable" with respect to systems dynamics, even without a controller.

Example: The temperature of a reactor is increasing after the heating power is increased stepwise. With increasing temperature the heat loss of the reactor to the environment is also increasing, until finally a new equilibrium condition at a higher temperature is reached, where the increased heat loss is equal to the enlarged heating power, and compensates for it.

Thought experiment: Please imagine a reactor with ideal thermal insulation, which means no thermal loss to the environment. Now, if the heating power is increased stepwise starting from the equilibrium condition, the temperature starts to rise. The increase of the temperature is undamped and continuous, as no physical effect in the opposite direction (an increasing heat flow to the environment according to the rising temperature) exists. Therefore, no new equilibrium condition is reached, resulting in an unstable control loop with respect to systems dynamics. This behaviour is called integral action.

There are other forms of instability besides the integrating behaviour, e.g. increasing oscillations. Such behaviours can rather be found in mechanical systems (e.g. the famous inverse pendulum). In process plants, such instabilities if appearing at all, are mostly due to inappropriate controller tunings, and only rarely appear in open loop.

Figure 2-1 Step response of a control loop with compensation (blue) and without compensation (red), i.e. with integral behaviour.



Unstable control loops cannot be stabilized without a controller. Therefore, switching a controller in such a loop to manual mode is not allowed for a longer time.

Hence, the recording of measurement data for the process identification (e.g. for the PID tuner or the MPC configurator) via step experiments in open control loop is not possible. The model type and the control algorithm of the MPC function block are also inappropriate for unstable control loops. Therefore, the unstable part transfer functions have to be stabilized by subordinated slave controllers before the application of the MPC.

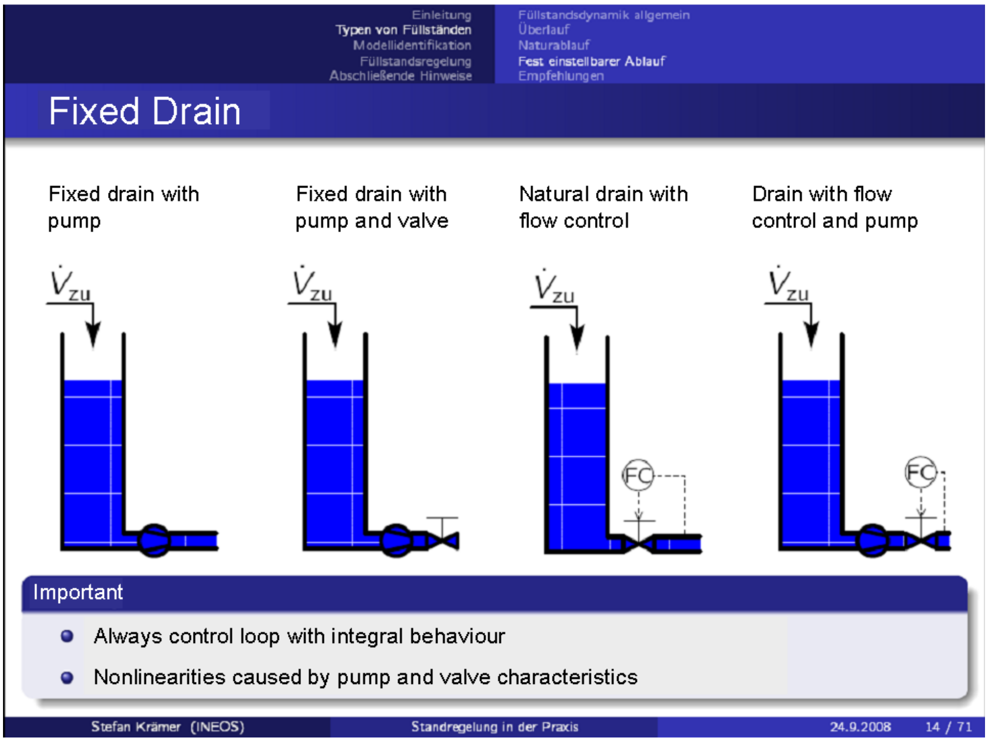
2.3 Examples of Unstable Control Loops

2.3.1 Level Control

If the level of a tank with continuous feed is to be controlled via an adjustable drain as actuator (e.g. pump or valve with or without flow control), the control loop shows integrating behaviour.

An equilibrium condition of the level only exists if the drain is exactly equal to the feed. The level permanently decreases until the tank is empty, if the drain is increased stepwise starting at this equilibrium condition. In contrast the level permanently increases until the tank overflows, if the drain is decreased stepwise starting at the equilibrium condition.

Figure 2-2 Types of level control, taken from Related Literature /1./.

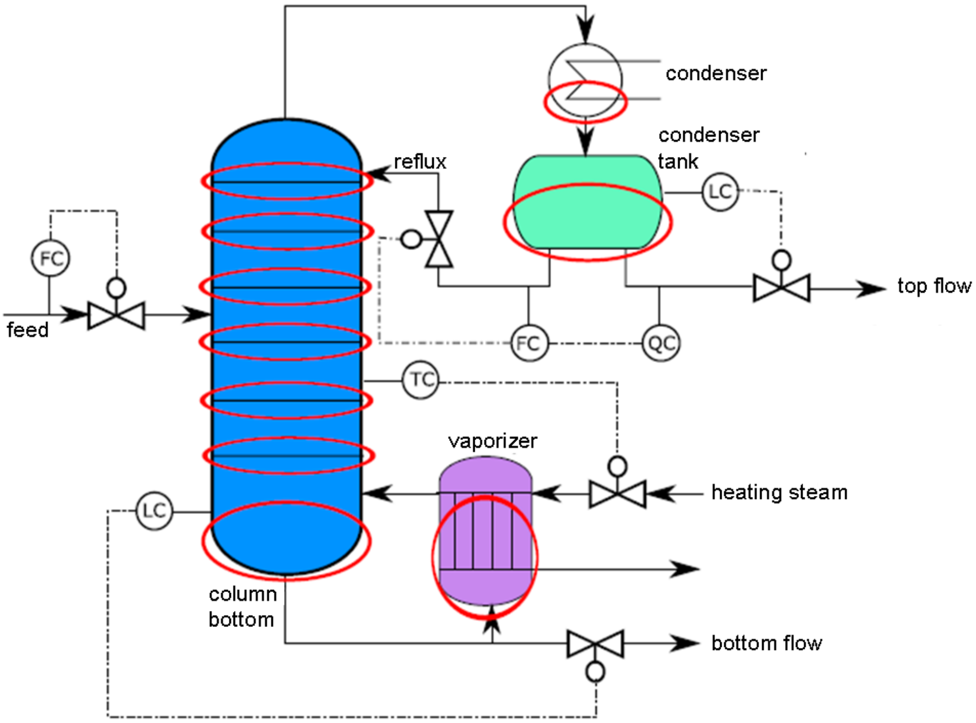


Exception natural drain: If only a valve with constant pressure behind exists in the drain of the tank, the drain flow depends according to the drain formula of Toricelli not only on the valve position (manipulated variable of the controller) but also nonlinearly on the level h itself (proportional to $\sqrt{2gh}$ with acceleration of gravity g). Compensation can be reached after small steps in the valve position in such control loops. The level is decreasing if the valve at the drain is opened a little bit starting at the equilibrium condition. Thus the hydrostatic pressure at the bottom of the tank is decreasing, and accordingly the drain flow is decreasing until a new equilibrium condition is reached. Anyway, the application of the techniques described in the following is helpful for stabilization of a natural drain too, as the mechanism of compensation only works for small steps and cannot be modelled linearly.

In process plants there are many tanks where level control is necessary, e.g. surge drums, separation tanks, stirred tank reactors, column sumps, feed water tanks, feed water tanks. There are different objectives for level control according to the plant context [also see Related Literature /2./]:

- Keep level constant (exactly at the set point) – important for levels directly influencing the process; disturbances are passed through to the output (drain).
- Keep level as small as possible – if “dead volume” and inventory are undesirable.
- Keep level inside specified limits while using the tank as buffer – changes in level are tolerated to achieve a smooth drain flow.

Figure 2-3 Examples of level control (marked in red) in a typical part of plant (distillation column) [also see Related Literature /1./].



2.3.2 Pressure Control in Tanks

In some cases the pressure control in tanks behaves in a similar way as the level control. The control loop shows an integral behaviour, if the manipulated variable is a gas feed (e.g. admission of inerts) and no pressure loss to the environment exists. Typically a separate purge valve exists in such cases to discharge gases. The pressure controller uses a split range function to access either the feed valve or the purge valve.

2.3.3 Position Control

The control loop shows integrating behaviour if the position of mechanical parts is controlled and the speed of the actuator is available as manipulated variable. An "equilibrium condition" without move in the position only exists for a speed equal to zero. The valve actuator is a common example for position control in process plants. However, the valve position controllers are mostly integrated in the corresponding actuators and hence not an issue for the DCS.

2.3.4 Example of Multivariable Control with Integral Part Transfer Function

The level of most chemical reactors has to be kept in certain limits during continuous operation. Therefore the drain flow (rotational speed of pump or set point of subordinated flow controller) is available as manipulated variable and results in an integral part transfer function as described in section 2.3.1. A multivariable control problem including integral part transfer functions results if an additional quality control exists at the same reactor, e.g. with the reactor pressure as manipulated variable (set point for the slave pressure controller). The residence time of the fluid in the reactor as well as the progress and the result of the reaction (the product quality) are dependant on the drain flow, which is the manipulated variable of the level controller.

3 Stabilization of Unstable Control Loops

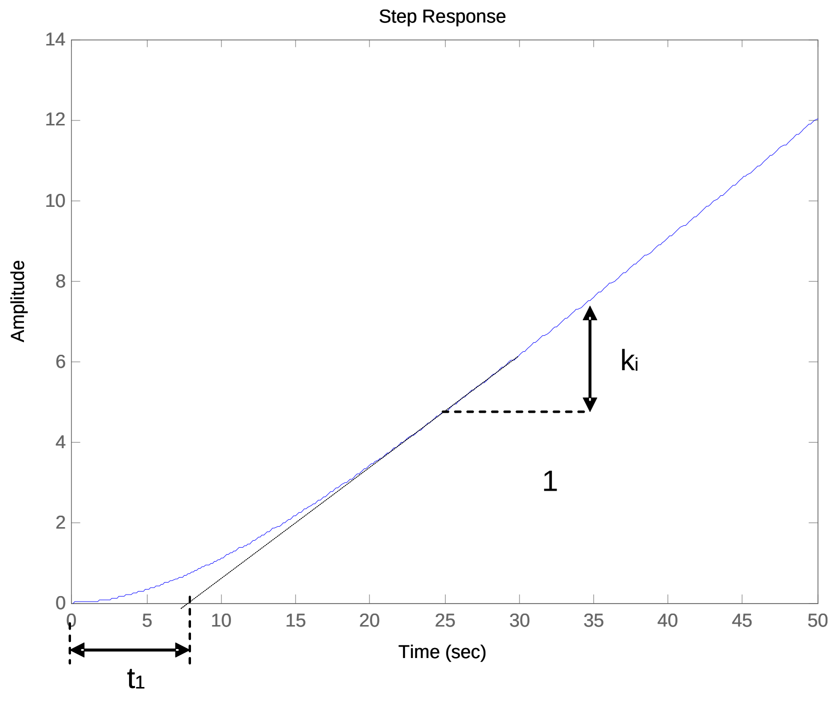
Regarding the stabilization of unstable control loops, integral processes and monotone unstable or oscillating unstable processes have to be distinguished.

In general only an analysis in frequency domain is helpful for oscillating control loops. As an example, displacements of unstable poles to the stable domain can be examined using root locus analysis [also see Related Literature /3./]. These oscillating control loops are a common issue in the context of mechanical systems (spring-damper-systems, elas-tical roboter arms) but can rarely be found in process plants. Oscillations in process plants can rather be attributed to malfunctions of slave control loops, e.g. in the valve position controllers.

In the following only the stabilization of integral processes will be discussed, due to the practical relevance in process engineering. A proportional-only controller is sufficient to stabilize integrating processes, as confirmed by systems dynamic considerations (e.g. root locus). Thus the problematic interaction of the integral part of a PI controller with the integral part of the plant is avoided. However, persistent control deviations caused by disturbances at the input of the process have to be accepted, if no integral action is used in the controller. Example: The proportional-only controller is not able to hold the level exactly at its set point if the feed is varying.

3.1 Manual Parameterization of a Proportional-only Controller for Integrating Processes

Figure 3-1 Unit-step response of an integrating process



The behaviour of an integrating process $g_i(s) = \frac{k_i}{s(t_1 s + 1)}$ can be described by two parameters:

- The maximal gradient k_i of the response to a unit-step (of height one)
- The delay time t_1 needed by the process to reach its maximal gradient after a step in the manipulated variable (intersection point of the tangent with the base line in Figure 3-1)

The transfer function of the closed loop including a proportional-only controller $k(s) = k_p$ (k_p is the proportional gain) is

$$g_{cl}(s) = \frac{g_i(s)k(s)}{1 + g_i(s)k(s)} = \frac{1}{\frac{t_1}{k_p k_i} s^2 + \frac{1}{k_p k_i} s + 1}$$

Thus the closed control loop has unity gain (the actual process value is equal to the set point in steady state, if no disturbance at the input occurs) and two poles at

$$s_{1/2} = -\frac{1}{k_p k_i} \pm \sqrt{\left(\frac{1}{k_p k_i}\right)^2 - 4\frac{t_1}{k_p k_i}}\ .$$

Both poles are real, if the (absolute value of) gain k_p of the controller is chosen such that

$$k_p < \frac{1}{4 t_1 k_i}\ .$$

Hence, an asymptotic stable control loop is ensured. A considerably smaller value is a good starting point for a stable controller parameterization and a following computer-based PID tuning, even if the specific values of the process are not known exactly.

If the process is uncritical, an adequately small gain can be chosen arbitrarily and used as starting point. You can increase this starting value iteratively until first indications of oscillations in the control loop become visible.

NOTE The sign of the controller gain must be negative, if the sign of the controlled process k_i is negative too (open drain valve -> level decreases)!

3.2 PID Tuner

The PCS 7 PID tuner can be used for integral processes without problems if at least one stable controller parameterization is already available. The following hints are helpful for this:

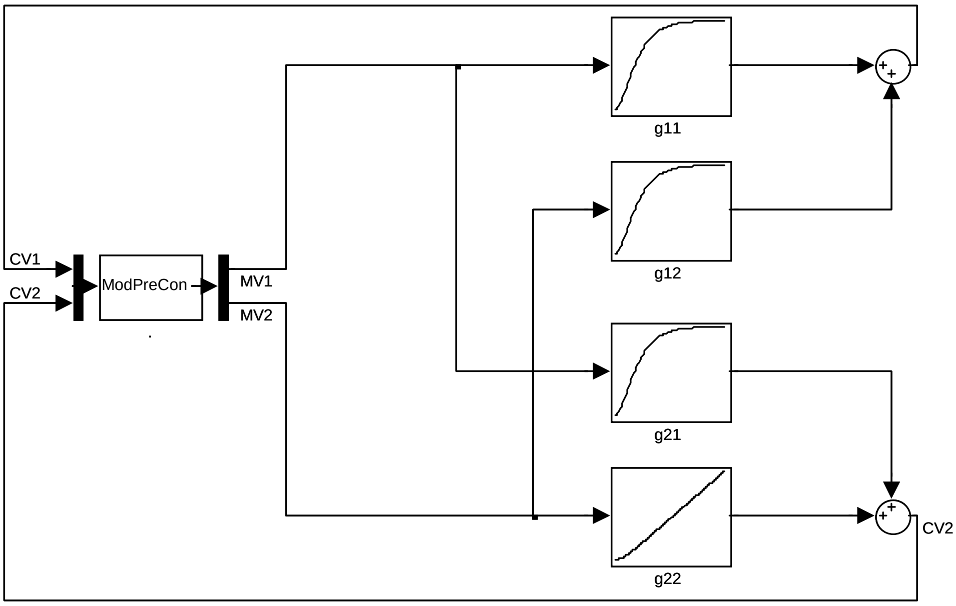
- Fix the check mark „With integral action in the process“
- Excite the process with a step in the set point in closed loop in automatic mode
- Chose the proportional only controller as controller type, at least, if you want to use this controller as slave controller in cascades (e.g. subordinated to a MPC) or if you are not interested in exact set point tracking.
- Set the MC_Offset to the value of the manipulated variable needed to reach the typical operating point, if you know it, in order to avoid the persistent control error at least in this operating point.

4 Configuration of MPC with Slave Controller

4.1 Starting Point

The starting point is the standard connection of the MPC with its actuators. The structure of the following example corresponds to the control of product quality (CV1) and level (CV2) in a reactor as mentioned in section 2.3.4. However, this example is not a realistic simulation of a real reactor.

Figure 4-1 Original signal flow chart of an MPC with 2x2 process, where the main transfer function g22 shows an integral action

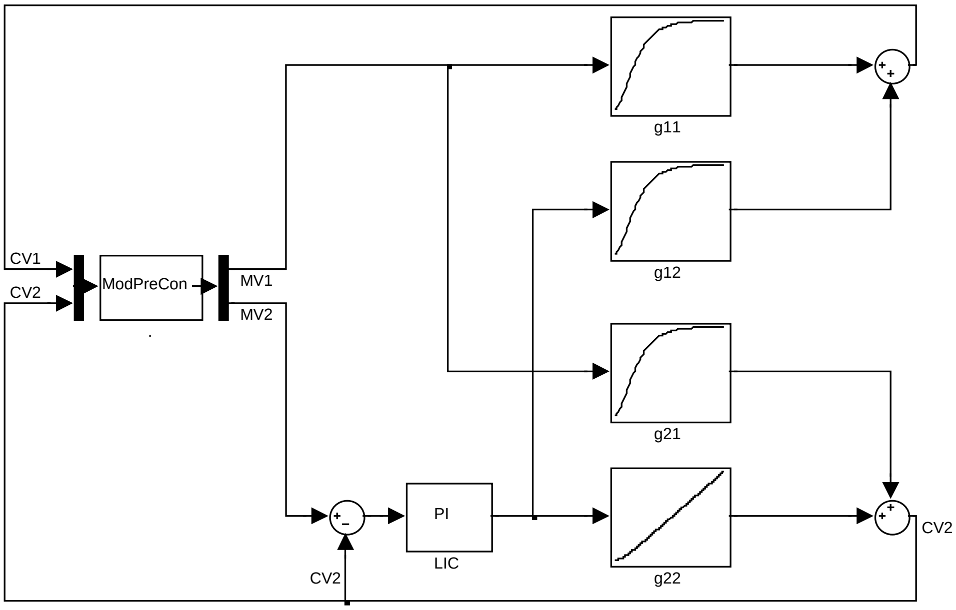


4.2 Connection in CFC

Now, a stabilizing PID controller is inserted for the integral sub process. In principal this control structure is a cascade consisting of an MPC as master controller and the stabilizing proportional-only controller as slave. The external set point SP_EXT of the slave controller is linked to the corresponding manipulated variable of the MPC.

The equality of the controlled variable of the master controller (i.e. the corresponding MPC control channel) and the controlled variable of the slave controller is the only special feature in the present case.

Figure 4-2 Signal flow chart of MPC with subordinated stabilizing PID controller for the integral main transfer function g22

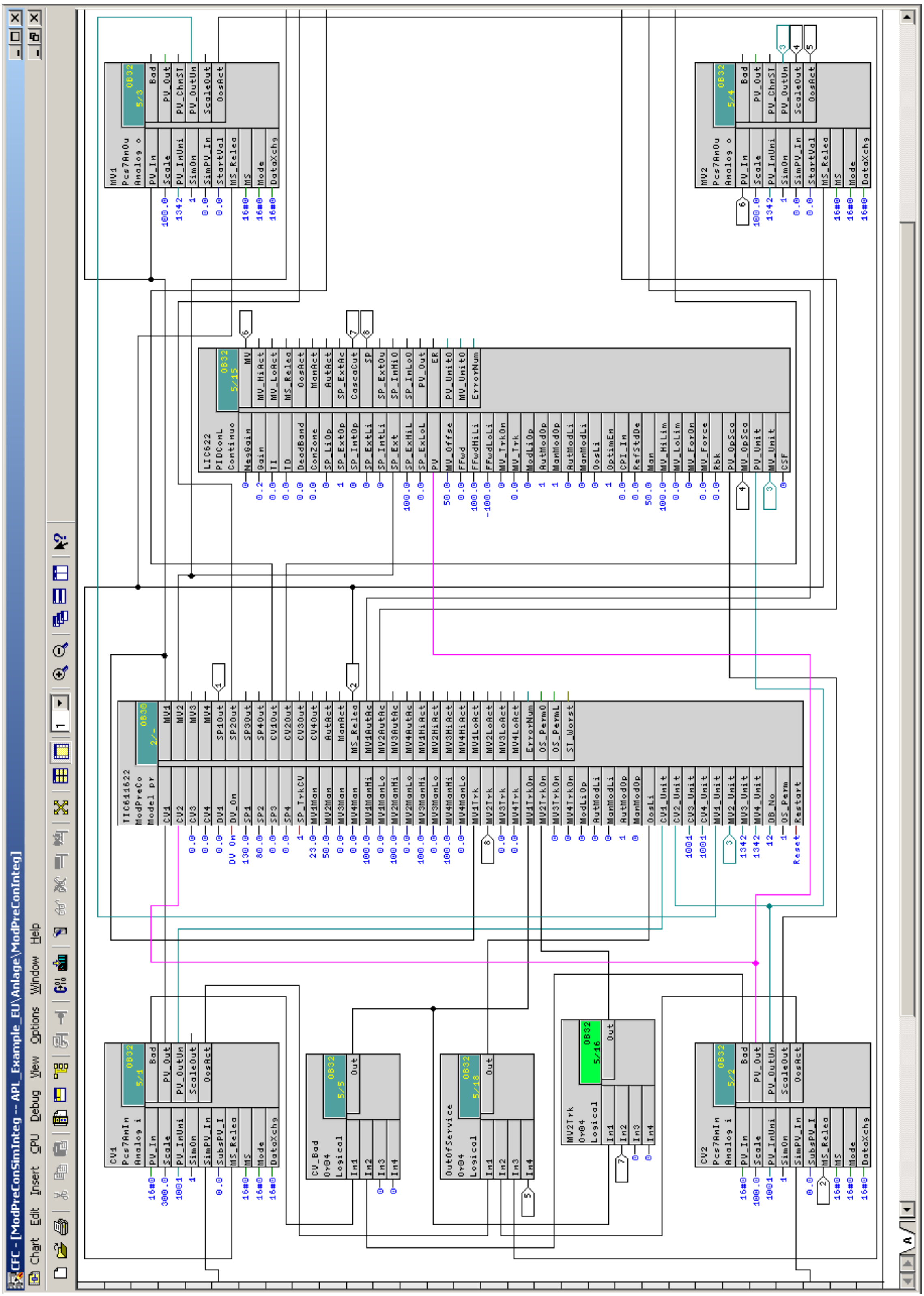


The slave controller stabilizes the control variable CV1 in general. The integral effect of the main transfer function g22 is compensates as well as the integral effect of the coupling transfer function g21. The influence of MV2 on CV1 is also modified by the slave controller due to g12.

All general notes on the configuration and commissioning of cascade controls are relevant for this case (see Figure 5-1):

- To get a correct anti windup calculation of the master controller, the range of the manipulated variables of the master controller (respectively the corresponding MPC channel) must be equal to the range of the external set point of the slave controller (PID.SP_ExtHiLim… SP_ExtLoLim). Typically the MV limits for automatic mode MViHiLim…MViLoLim are set tighter than the ones for manual mode MViManHiLim…MViManLoLim in an MPC. Hence, the limits for manual mode are set equal to the limits of the set point of the slave controller and the ones for automatic mode are set even tighter only if necessary.
- The master controller must be set to “tracking mode”, if the slave controller is not in cascade mode (automatic mode with external set point) but in any other mode (e.g. manual or automatic mode with internal set point) with no reaction to instructions by the master controller (announced by PID.CascaCut= true). The “tracking mode” must also be activated if a bad status of measurement data at the master controller is detected. An OR-combination of both conditions is passed to the binary input MPC.MV2TrkOn. To ensure a bumpless switching back to cascade mode, the manipulated variable of the master controller MPC.MV2Trk is linked to the current set point PID.SP of the slave controller.
- The cycle time of the slave controllers in cascades must be at least as fast as the cycle time of the master controller. In the present case this is ensured automatically: the slave P(ID) controller runs in a standard fast cycle of the automation system (typically 1s), while the MPC is moved to a slow cycle specific to the application after the model identification.

Figure 4-3 Connection of MPC and slave controller



4.3 Commissioning

The parameterization of the controller and the commissioning is done „from interior to exterior" as in any cascade control. First the slave controller is tuned (see chapter 3) and switched to automatic mode. Afterwards the slave controller is switched to cascade mode and the master controller is parameterized. While tuning the master controller please consider that the whole inner closed control loop of the slave controller is the controlled process of the master controller. Therefore the adjustable parameters of the master controller are not independent of the tuning of the slave controller. The step experiments for the identification of the MPC models can only be executed after the linking of the stabilizing slave controller, as the additional proportional-only controller affects the sub transfer functions g22, g21 and g12 (see Figure 4-2).

A dead band can be used in the slave controller, if the level is not to be exactly controlled to its set point. The controller has no reason to interfere with the process as long as the controlled variable is inside the dead band. The control deviation seems to be zero for the controller. Therefore, the control signal can be smoothed to avoid valve wear, and variations of the drain flow can be reduced to obtain a smooth feed for the downstream process components. The dead band should be adjusted before the measurement data for the MPC configurator is recorded, as the dead band influences the behaviour of the slave control loop. A dead band of at least the same size must also be specified for the corresponding control channel of the master controller. Of course, the master controller is not able to reach the set point more precisely as it is allowed by the dead band of the slave controller.

5 Simulation Example

The simulation example was generated from a copy of the plant section ModPre-Con of the APL_Example_EU, by introducing an additional integral block after the transfer function Proc662.

Figure 5-1 Modified process simulation of the example project; the inserted integrator is marked in blue

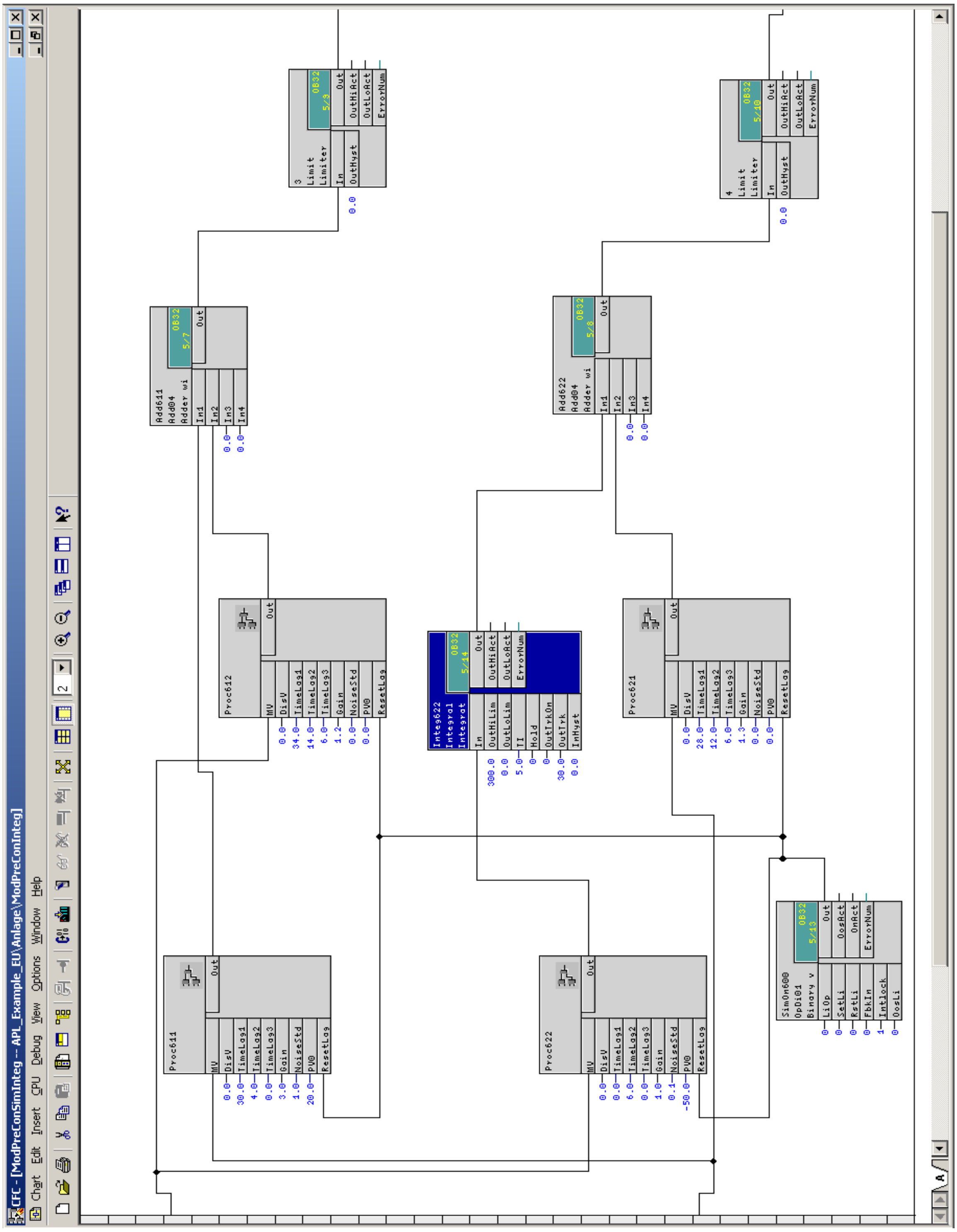
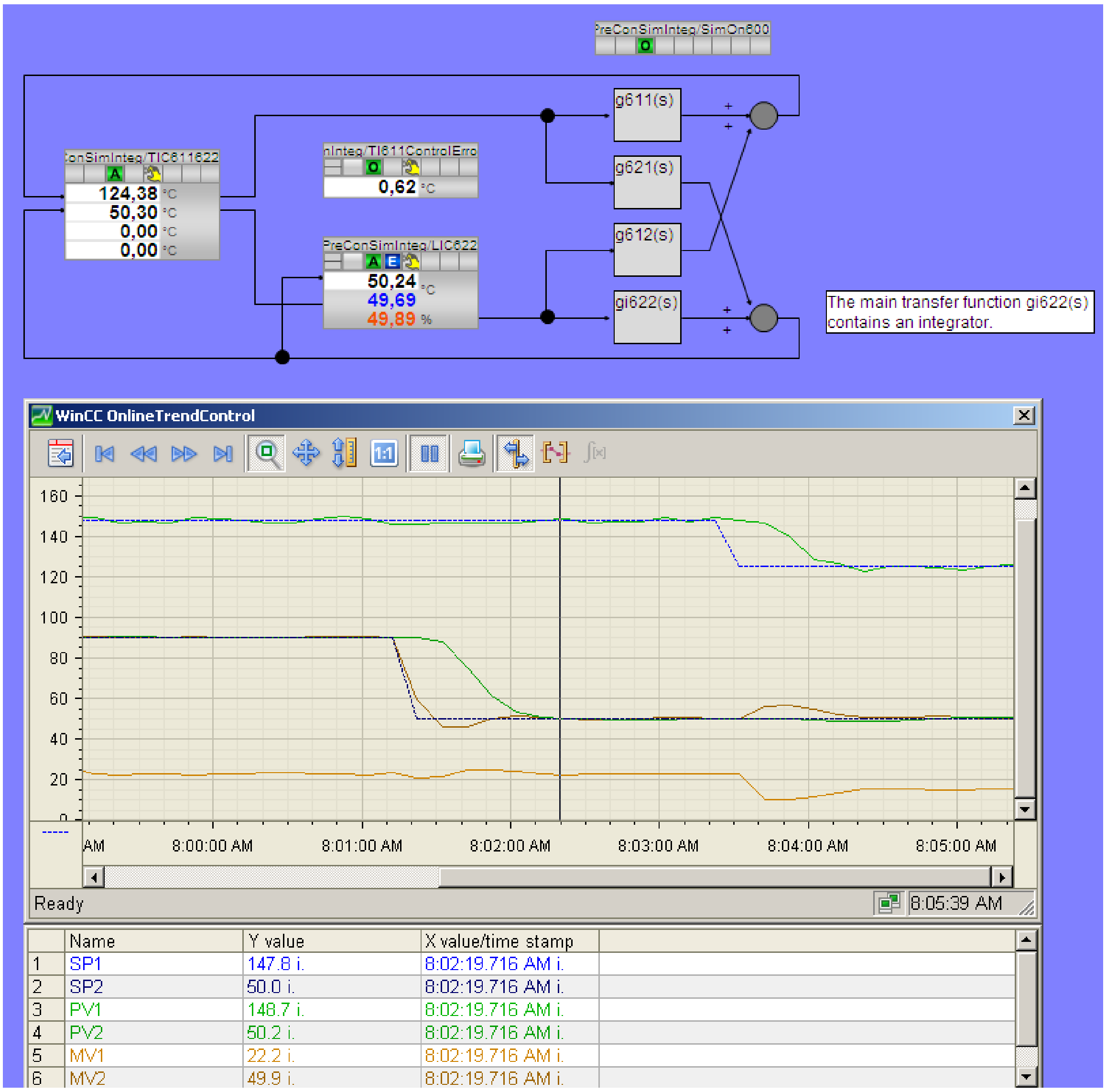


Figure 5-2 OS picture of the example project



Despite the interaction between both manipulated variables MV1 and MV2 and the integral action of MV2 on PV2, both control variables PV1 and PV2 of the master controller can be controlled to their given set points independently of each other, which is a success of the described control concept. A “crosstalk” between the interacting control loops can mostly be avoided, e.g. the level PV2 (dark green) is only moved minimally during a set point step in SP1 (light blue). Specifics: Not only the current value of the level PV2 (dark green) is reaching the set point SP2 of the MPC (dark blue dashed) but also MV2 of the MPC (dark brown) in steady state, as MV2 is simultaneously the set point of the slave level controller.

6 Conclusion

The area of application of the model predictive controller embedded in SIMATIC PCS 7 is extended clearly by the described stabilization of unstable sub control loops with the help of a slave proportional-only controller. A typical application is an MPC where one controlled variable is the level of a tank, reactor, etc.

7 Related Literature

7.1 Bibliography

This list is not complete and only represents a selection of relevant literature.

Tabelle 7-1

	Titel
/1/	Krämer, S. Auslegung von Standreglern in der verfahrenstechnischen Praxis. Dechema-Seminar „Prozessregelungen – von den Grundlagen zu Advanced Control“, Sep. 2008.
/2/	Cheung, Tak-Fai und William L. Luyben Liquid-Level Control in Single Tanks and Cascade of Tanks with Proportional Only and Proportional-Integral Feed-back Controllers. Ind. Eng. Chem. Fund., 18(1):15–21, 1979.

7.2 Internet Link Specifications

This list is not complete and only represents a selection of relevant information.

Tabelle 7-2

	Subject	Titel
\1\	Referenz auf den Beitrag	http://support.automation.siemens.com/WW/view/de/42200753
\2\	Siemens I IA/DT Customer Support	http://support.automation.siemens.com
\3\		http://en.wikipedia.org/wiki/Root_locus

8 History

Tabelle 8-1

Version	Date	Modifications
V1.0	04/2010	First version