

**SLOVAK UNIVERSITY OF TECHNOLOGY IN  
BRATISLAVA  
FACULTY OF CHEMICAL AND FOOD TECHNOLOGY**

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**STABILISATION OF COLUMN FEED USING APC  
MASTER THESIS**

**2017**

**Bc. Martin Mišenko**



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Study program: Automation and Information Engineering in Chemistry and Food Industry

Study field: 5.2.14. Automation

Workplace: Faculty of Chemical and Food Technology

Thesis supervisor: prof. Ing. Miroslav Fikar, DrSc.

Consultant: Ing. Karol Ľubušký

**Bratislava 2017**

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## MASTER THESIS TOPIC

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Student's ID: 61616  
Study programme: Automation and Information Engineering in Chemistry and Food Industry  
Study field: 5.2.14. Automation  
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Topic: **Stabilisation of Column Feed using APC**

Language of thesis: English

Specification of Assignment:

Non traditional usage of APC for control of complex of tanks at BCDU6 Slovnaft Plc. There are two reflux drums, from which liquid is routed to surge drum. Liquid in surge drum is a feed for a next column. Inlet flows to reflux drum are disturbances and they fluctuate remarkably. Fluctuation is transferred to column feed and this disturbs operation of all columns after and thus product quality deteriorates. Aim of control is to use capacity of these drums to stabilise feed from surge drum as much as possible. Improve existing traditional flow/level controllers using MIMO APC strategy.

Tasks:

1. Create dynamic model of controlled system using MATLAB.
2. Test basic strategies MATLAB simulation.
3. Setup APC controller using Profit Suite.
4. Implement APC to control given tank system at BCDU6 at Slovnaft Plc.

Length of thesis: 50

Selected bibliography:

1. Mikleš, J. – Fikar, M. *Process Modelling, Identification, and Control*. Berlin Heidelberg: Springer Berlin Heidelberg New York, 2007. 480 s. ISBN 978-3-540-71969-4.
2. M. King, *Process Control: A Practical Approach*, Wiley, 2011.

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# Abstract

Liquid tanks are quite common devices in refinery industry. A cascade control is ordinarily used to control the level in tanks, however, it controls only separate tank and it does not take interactions between tanks into account. When it is necessary to coordinate control between tanks, Advanced Process Control (APC) is suitable to implement. APC is multivariable control, so it can treat the whole system of tanks as one complex system. The aim was to propose and implement APC control for tanks system at BCDU6 unit in Slovnaft Plc, to mitigate fluctuations in output flow from the system.

The first part of this work deals with mathematical modeling and basic control of the given tanks system in Matlab-Simulink. In the second part, we are implementing the APC controller, which controls the given tank system in BCDU6 unit. We use Honeywell software Profit Suite in this step.

**Key words:** Honeywell; APC controller; liquid tanks

# Abstrakt

Zásobníky kvapaliny sú bežnou súčasťou rafinérskych zariadení. Na riadenie výšky hladiny sa zvyčajne používa kaskádová regulácia, ktorá však riadi iba jeden konkrétny zásobník a neberie do úvahy interakcie s ostatnými zásobníkmi. Ak chceme koordinovať riadenie medzi viacerými zásobníkmi, je vhodné použiť Advanced Process Control (APC) riadenie. APC je viacrozmerové riadenie, takže nám umožňuje riadiť systém zásobníkov ako jeden celok. Cieľom práce bolo navrhnúť a implementovať APC riadenie pre systém zásobníkov na prevádzke AVD6 v Slovnaft a.s., ktoré bude čím viac tlmiť výkyvy na výstupnom prietoku z riadeného systému.

Práca sa v prvej časti zaoberá matematickým modelovaním a základným riadením zásobníkov v Matlabe - Simulinku. V druhej časti implementujeme a vyhodnotíme APC regulátor, ktorým riadime sústavu troch zásobníkov na prevádzke AVD6. Používame pritom software Profit Suite od spoločnosti Honeywell.

**Kľúčové slová:** Honeywell; APC regulátor; zásobníky kvapaliny

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# Abbreviations

BCDU6	Bratislava Crude Distillation Unit 6
PID	Proportional Integral Derivative Controller
SP	Setpoint
K	Controller gain
$T_s$	Time constant
$T_i$	Integral time constant
DCS	Distributed Control System
FI	Flow Indication
LC	Level Control
FC	Flow Control
LHA	Level High Alarm
LLA	Level Low Alarm
APC	Advanced Process Control
RMPCT	Robust Multivariable Predictive Control Technology
CV	Controlled Variable
MV	Manipulated Variable
DV	Disturbance Variable
PV	Process Variable
SISO	Single Input Single Output
PDS	Profit Design Studio
FIR	Finite Impulse Response
URT	Unified Real Time
PSRS	Profit Suite Runtime Studio



# Introduction

Liquid tanks are common devices used in all chemical industry. We focus on petrochemical industry, specifically on Bratislava refinery Slovnaft Plc. There is a production unit called Bratislava Crude Distillation Unit 6 (BCDU6). BCDU6 contains atmospheric and vacuum distillation parts. Among the vessels in atmospheric part, there are three horizontal cylindrical tanks (T1, T2, T3), which are subject of our work. Outputs from tanks T1 and T2 are routed into T3, which is a feed buffer for a redistillation column. Unfortunately, output flows from T1 and T2 fluctuate, especially the second one. This fact causes fluctuation of T3 feed flow to the redistillation column. This disturbs heavily operation in the column. Our aim is to decrease the feed fluctuation as possible.

The first part of this work deals with the synthesis of a mathematical model of treated tanks, which is necessary for controller design. We have three horizontal cylindrical tanks with same construction but different parameters. This part also provides verification of derived model in MATLAB – Simulink. There were designed several types of control systems including averaging level control, the percentage of working volume control and controller with gain scheduling. We also reproduced Slovnaft cascade control.

The second part is about APC controller design. At first, we need to define variables and design model matrix of our sub – processes. Data for identification were prepared in MATLAB and then exported to Profit Design Studio. After identification, we started with controller implementation in Profit Suite Runtime Studio. Instead of raw level control, we used a percentage of working volume for control. Designed controller has been switched on in Slovnaft and some another improvement were done.

# 1 Bratislava Crude Distillation Unit 6

The first step in crude processing in an oil refinery is distillation. Distillation separates crude oil to several fractions. It is physical process based on different boiling temperatures of desired fractions. It means that hydrocarbons already present in crude are separated and no chemical changes are intended. Its products do not satisfy requirements for final products and various refinery units treat them. Slovnaft refinery has two crude oil distillation units – BADU5 and BCDU6. Since our work is oriented on BCDU6, we will deal only with this one in following parts of work.

## Technology Description

Since BCDU6 technology is quite complex, we will focus only on control systems and its upstream processes. The first step in crude distillation is desalination. Oil is preheated in a heat exchange system and mixed with wash water. Water is added to dissolve the salts and clear out mechanical impurities. A mixture of oil and water enters desalter tanks, where hydrocarbons and water are separated.

Desalted crude oil is preheated and enters preflash column. The main task of preflash column is to separate the lightest hydrocarbon fractions from oil and thus unload atmospheric furnace and column. C1 overhead product is called preflash heavy naphtha. Bottom product of C1 is called preflash crude oil. Preflashed crude oil is transferred to the atmospheric column where oil is divided to atmospheric heavy naphtha, kerosene, light gas oil, heavy gas oil and atmospheric residue. Preflash and atmospheric heavy naphtha are mixed in T3 tank. Heavy naphtha from T3 is redistilled in redistillation column C4.

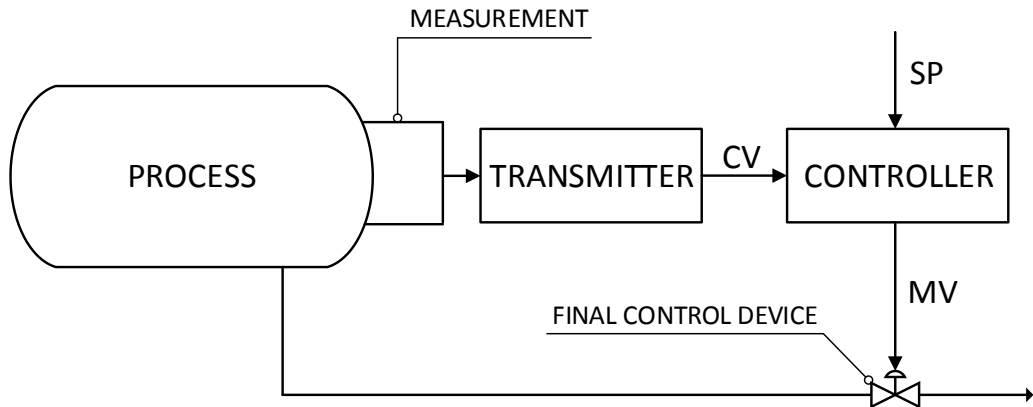
With a basic knowledge of related technology, we can focus on control system. At first, we will explain basic principles and terminology of basic automatic control.

# 1.1 Automatic Control

Process control involves manipulation of process variables to some desired value. Many of process variables are dependent on another variable but they are controlled independently. We know two objectives in control systems:

1. In a case of set point change, the control system has to achieve new set point with the process variable.
2. If there is any disturbance in the system, the control system has to reject this disturbance and maintain process variable on the set point.

Most common type of control system is a feedback control loop (Fig. 1). A process deviation must occur in order for a control action to be made. The controller is a brain of the control loop which receives the value of process variable from measurement from the process. This value compares with the set point and calculates the error. Then controller sends a signal to a final control device.



**Fig. 1. Feedback control loop.**

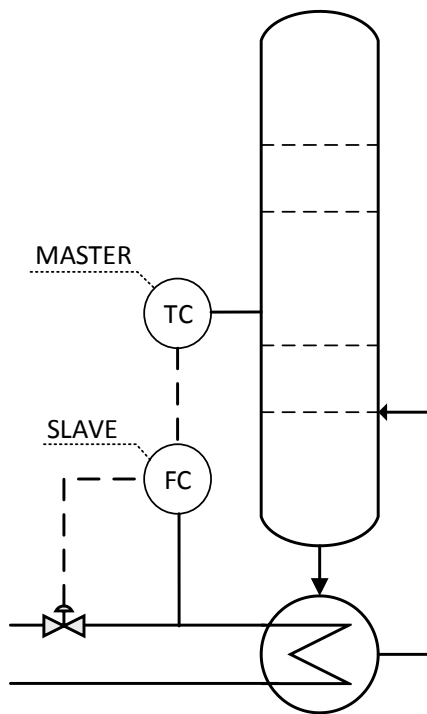
PID controllers are used to reacting to process changes automatically. They should bring process variable back to steady state after some deviation occurs. PID contains three parts:

- Proportional (P)
- Integral (I)
- Derivative (D)

Honeywell uses an ideal form of PID controllers:

$$u(t) = K_c \cdot \left[ e(t) + \frac{1}{T_i} \int_0^t e \, dt + T_d \frac{de}{dt} \right] \quad (1)$$

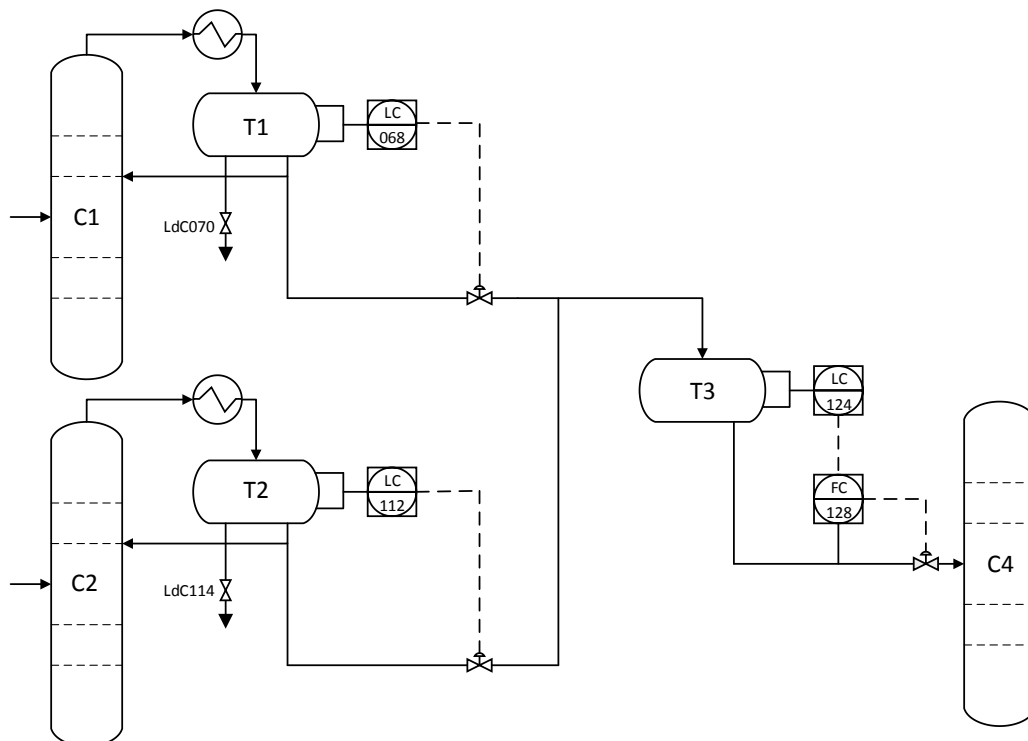
A better solution to decrease the impact of disturbances is to use cascade control. Usually, there are two controllers. First of them (master) controls main process variable like temperature and calculates a setpoint for the secondary controller (slave) which controls manipulated variable like the flow of steam (Fig. 2). Secondary controller is usually P controller and it should be faster than the master controller to deal with the disturbances (Bakošová, Fikar, 2012).



**Fig. 2. Cascade control of distillation column.**

## 1.2 C4 Feed Control

We have three handlers to affect C4 feed: C1 head control, C2 head control and T3 control as shown in Fig. 3. In following parts, we will describe all handlers more closely.



**Fig. 3. C4 feed control.**

### **C1 heavy naphtha:**

Cooled heavy naphtha with gases from preflash column enters to reflux tank T1. Gases which are not condensed are routed into refinery gas storage. Reflux tank - T1 separates water from heavy naphtha by built partition (see Chapter 2). Separated water flows through valve LV070 to a chemical canalization automatically. Heavy naphtha from T1 to T3 is controlled by T1 heavy naphtha level controller LC068. This stream flow is measured by FI076. The flow of the second stream of heavy naphtha – C1 reflux, is controlled by C1 reflux controller FC051 in a cascade from TC052 (temperature on the head of the column)

### **C2 heavy naphtha:**

Cooled mixture of heavy naphtha and water from atmospheric column C2 enters tank T2 where water is separated and flows through valve LV114 to the chemical canalization. Heavy naphtha

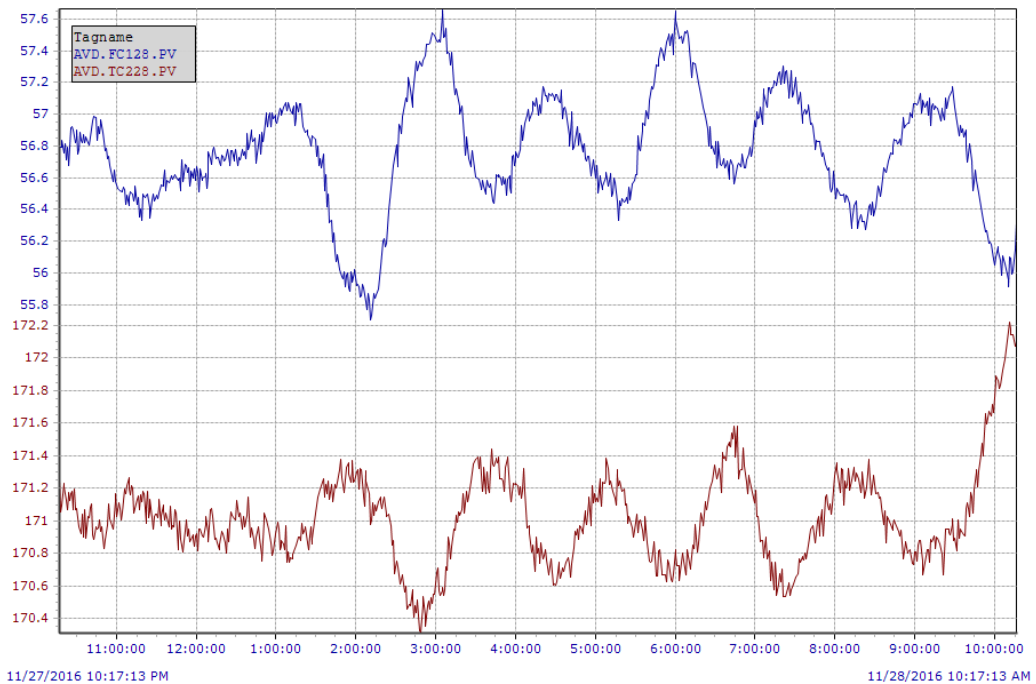
from T2 to T3 is controlled by T2 heavy naphtha level controller LC112 controller and measured by FT118.

### T3 control:

Heavy naphthas from preflash and atmospheric columns are mixed in tank T3. A mixture of heavy naphtha (feed to C4) is controlled by flow controller FC128 in a cascade from T3 heavy naphtha level controller LC124. Heavy naphtha from T3 goes through exchangers to redistillation column C4 on 27th tray. Products from C4 column are a gas to low pressure gas storage, C5/C6 fraction, light heavy naphtha and heavy naphtha.

### C4 feed control issues:

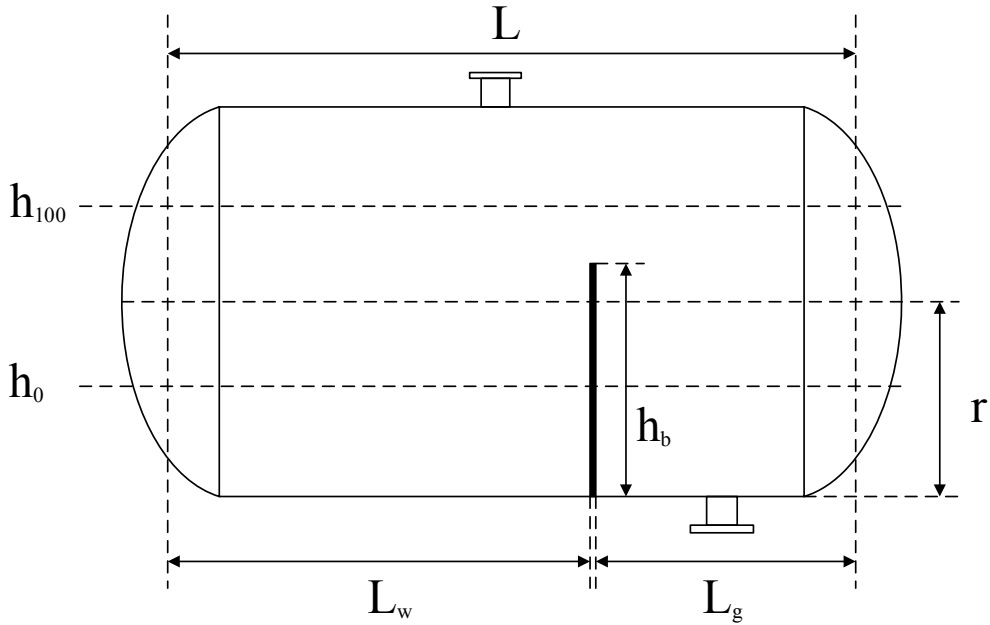
Heavy naphtha flow from T2 to T3 oscillates significantly, for reasons which cannot be resolved. Due to the current control setup, these fluctuations are transferred downstream to the C4 column. Fluctuation of feed flow to C4 column causes fluctuation of temperature, which has a negative impact on quality of heavy naphtha. See Fig. 4, where the blue line is feed flow (t/h) and the red line is temperature on 24th tray ( $^{\circ}\text{C}$ ). The purpose of this work will be to propose and implement control system to mitigate given fluctuations.



**Fig. 4. Fluctuations of feed and temperature.**

## 2 Modeling of the Tank System

All three tanks (T1, T2 and T3) have a similar construction. The difference is only between construction parameters values. In Fig. 5 we can see the scheme of the horizontal tank. Construction parameters are in the table below. The tank has rounded ends, but we omitted this for simplification.



**Fig. 5. Tank scheme.**

In a table below there are values of construction parameters for T3:

**Table 1. Tank parameters.**

Parameter	Description	Value	Unit
$L$	the length of the tank	5.803	m
$L_w$	the length of water side	3.375	m
$L_g$	the length of heavy naphtha side	2.428	m
$r$	the radius of the tank	1.184	m
$h_b$	the height of the barrier	1.400	m
$h_0$	height of 0% level indication	0.675	m
$h_{100}$	height of 100% level indication	1.725	m

## 2.1 Derivation of Model

We need more parameters to derive a model:

**Table 2. Model parameters.**

Parameter	Description	Unit
$q_{in}(t)$	input flow to tank	m <sup>3</sup> /h
$q_{out}(t)$	output flow from the tank	m <sup>3</sup> /h
$h(t)$	the height of level in the tank	m
$\rho$	density	t/m <sup>3</sup>
$\dot{m}_{in}(t)$	input mass flow to tank	t/h
$\dot{m}_{out}(t)$	output mass flow from the tank	t/h

Mass balance:

$$\{input\ flow\ to\ tank\} = \{output\ flow\ from\ tank\} + \{accumulation\ of\ liquid\ in\ tank\}$$

$$q_{in} = q_{out} + \frac{dV}{dt} \quad (2)$$

The volume of the tank depends on the height of level and the level varies with time. We can write:

$$\frac{dV(h(t))}{dt} = q_{in}(t) - q_{out}(t) \quad (3)$$

with initial condition  $h(0) = h^s$ .

Instead of volumetric flows we rather use mass flows:

$$\frac{dh(t)}{dt} = \frac{\dot{m}_{in}(t) - \dot{m}_{out}(t)}{\frac{dV(h)}{dh} \rho} \quad (4)$$

The volume of the horizontal tank can be calculated as (King, 2011):

$$V = \left[ r^2 \cos^{-1} \left( \frac{r-h}{r} \right) - (r-h) \sqrt{2rh - h^2} \right] L \quad (5)$$

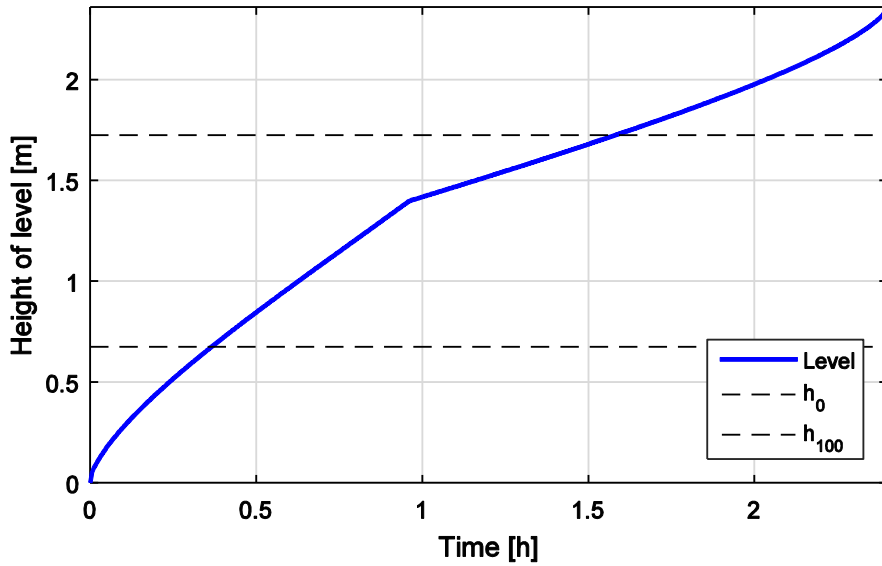
Finally, we have a mathematic model:

$$\frac{dh(t)}{dt} = \frac{\dot{m}_{in}(t) - \dot{m}_{out}(t)}{2L\sqrt{-h(h-2r)}\rho} \quad (6)$$

This calculation assumes a nonlinear relationship between volume and level indication. We explain this argument in a small experiment. Let's say  $h(0) = 0$  m and input mass flow  $\dot{m}_{in} = 5$



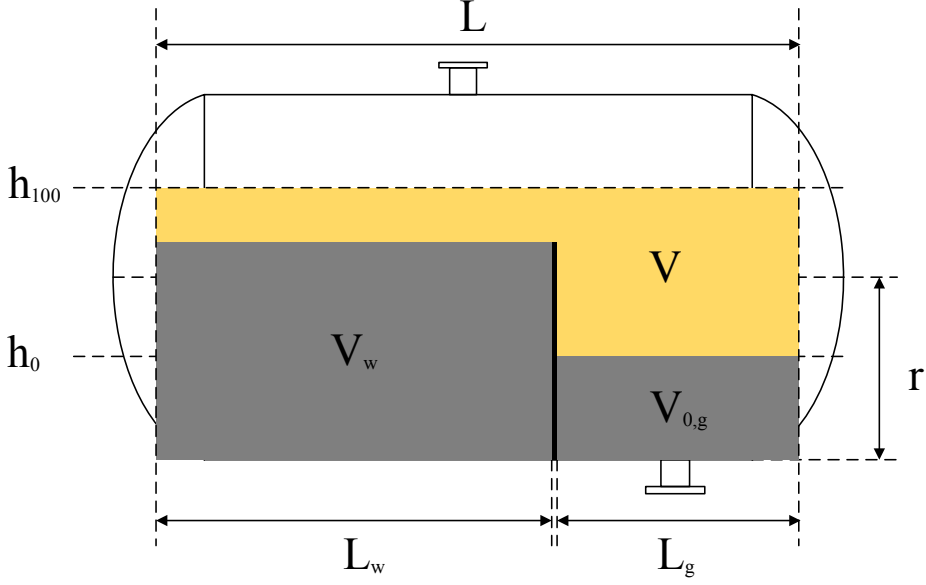
t/h. There is no output flow. After circa 2.4 hours the tank will be fully filled, which can be seen in Fig. 6. Increasing of level in time is nonlinear.



**Fig. 6. Nonlinear behavior of level indication.**

Nonlinear behavior is partially removed by the location of measuring sensors of the level. As can be seen in Fig. 6, main nonlinearity is not measured, it occurs under and over level indicators.

There is another way to remove nonlinearity. Instead of the height of the level, we can use a percentage of working volume for control level. We need to calculate total measured and usable volume of the tank. For better illustration, all necessary volumes are shown in Fig. 7.



**Fig. 7. Volumes to calculate.**

Firstly we calculate  $V_{0,g}$ , (King, 2011):

$$V_{0,g} = \left[ r^2 \cos^{-1} \left( \frac{r - h_0}{r} \right) - (r - h_0) \sqrt{2rh_0 - h_0^2} \right] L_g \quad (7)$$

The volume of water side is constant:

$$V_w = \left[ r^2 \cos^{-1} \left( \frac{r - h_b}{r} \right) - (r - h_b) \sqrt{2rh_b - h_b^2} \right] L_w \quad (8)$$

The working volume will be:

$$V = \left[ r^2 \cos^{-1} \left( \frac{r - h_{100}}{r} \right) - (r - h_{100}) \sqrt{2rh_{100} - h_{100}^2} \right] L - V_{0,g} - V_w \quad (9)$$

Current measured volume  $V_m$  in the tank which varies with time we calculate using equation (9), but there are some changing parameters because of built barrier:

If  $h \leq h_b$

$$V_m = \left[ r^2 \cos^{-1} \left( \frac{r - h}{r} \right) - (r - h) \sqrt{2rh - h^2} \right] L_g - V_{0,g} \quad (10)$$

Else

$$V_m = \left[ r^2 \cos^{-1} \left( \frac{r - h}{r} \right) - (r - h) \sqrt{2rh - h^2} \right] L - V_{0,g} - V_w \quad (11)$$

Percentage of working volume we calculate as follow:

$$100 \frac{V_m}{V} \quad (12)$$

Fig. 8 describes the linear behavior of increasing percentage of working volume.

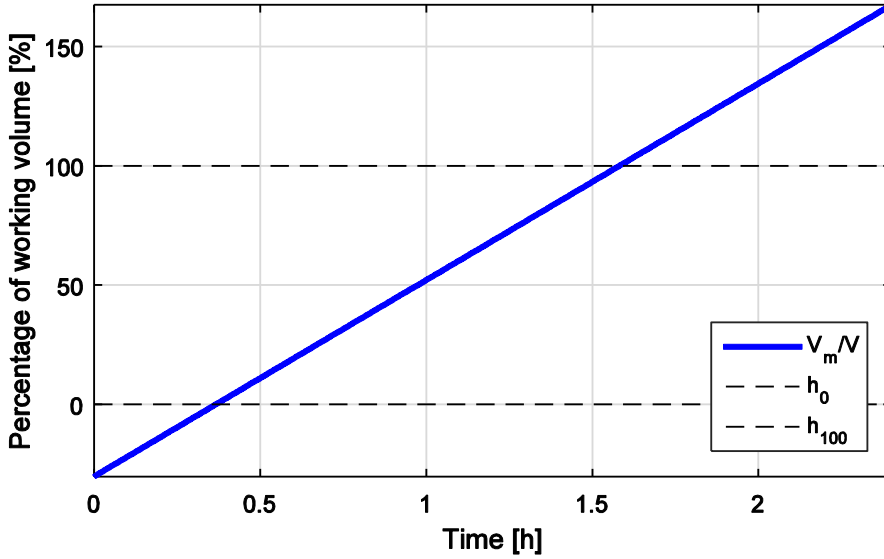


Fig. 8. Linear behavior of percentage of working volume.

## 2.2 Model Validation

Validation checks the accuracy of the model's representation of the real system. In our case, we used data from Slovnaft to check the validity of our system. We used given input and output flows and checked the height of level in T3. Level obtained from the model was compared with given level from Slovnaft.

Ideally, input to tank is equal to output from tank

$$\dot{m}_{out} = \dot{m}_{in,1} + \dot{m}_{in,2} \quad (13)$$

but data from Slovnaft do not satisfy the equation (13) as can be seen in Fig. 9.

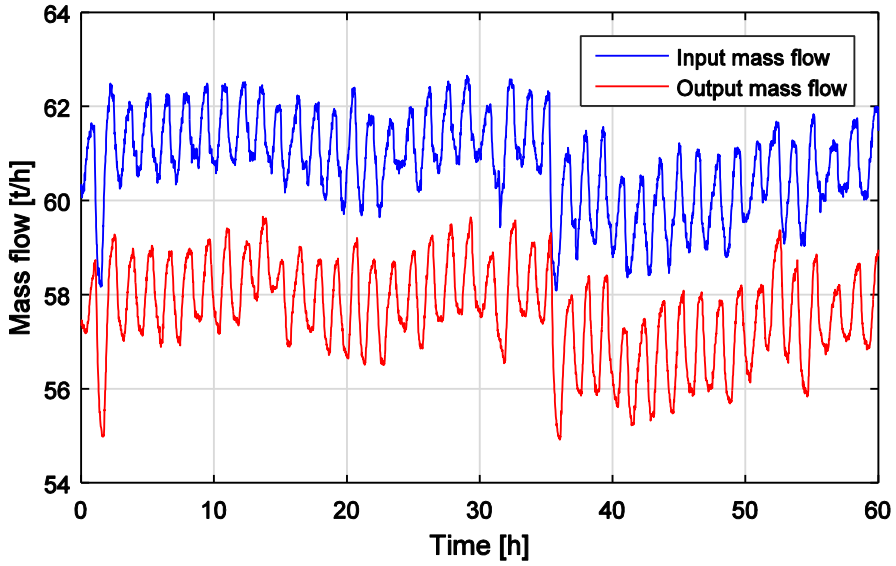


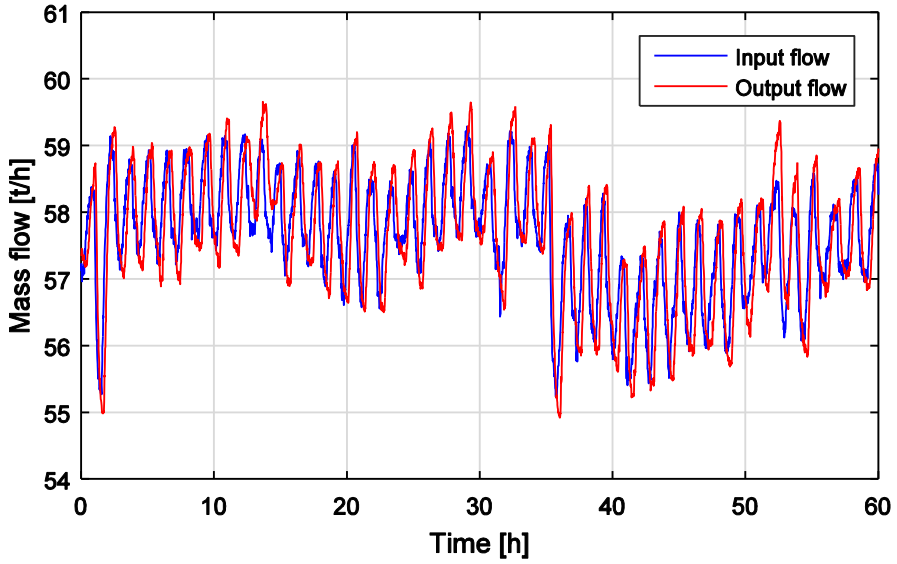
Fig. 9. Input and output flow comparison.

Based on process data, this is caused by a proportional error on sensors of mass flow of output from T1 and T2, which sum is input to T3. By using input presented above ( $\dot{m}_{in,1} + \dot{m}_{in,2}$ ), the T3 is filled in a few minutes, because input is always bigger than output.

If our input and output mass flows should be equal, we need to multiply outputs flows from T1 and T2 by constants calculated by regression of process data:

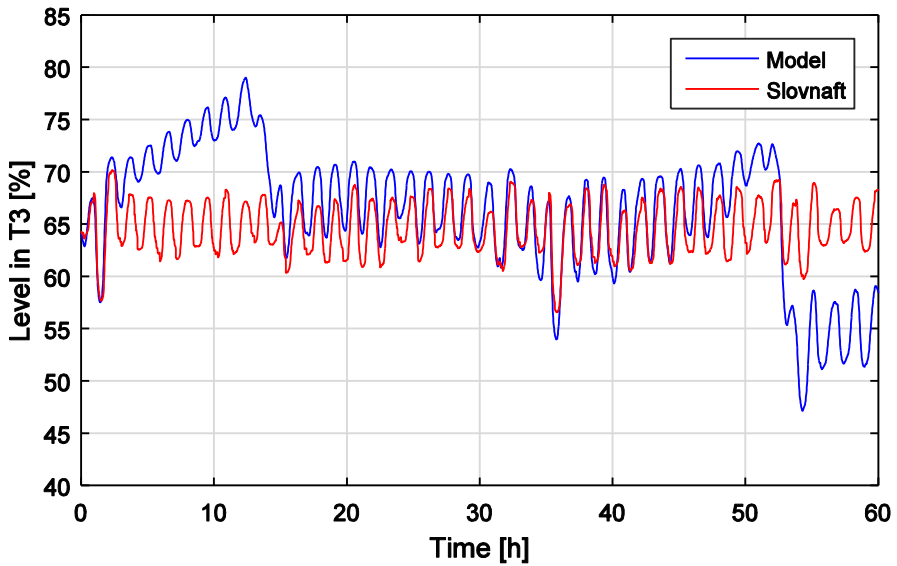
$$\dot{m}_{out} = \dot{m}_{in,1} \cdot 1.0303 + \dot{m}_{in,2} \cdot 0.8846 \quad (14)$$

After this correction output is approximately the same as input (Fig. 10).



**Fig. 10. Input and output flow comparison after correction.**

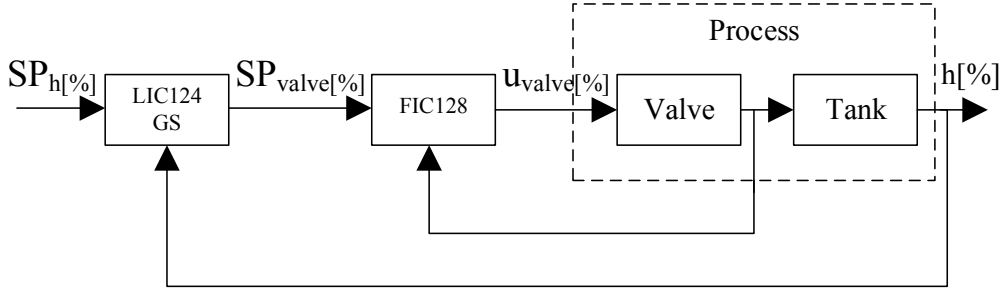
Now we can validate our model by comparing level from the model and from Slovnaft data, which is shown in Fig. 11. We succeeded with type of oscillations but it is unstable process, so every small change can lead to divergence as you can see in Fig. 11 until 15th hour or from 53rd hour.



**Fig. 11. Model and Slovnaft level comparison.**

### 3 Slovnaft Control System

In order to compare our proposals with the current solution, we created a simulation of the current control system. Slovnaft uses cascade structure to control level in T3 as is shown in Fig. 12.



**Fig. 12. Cascade control in Slovnaft**

Both controllers work with sampling time of 15 seconds and have following structure:

$$C(z) = K \cdot \left( 1 + \frac{T_s}{T_i} \frac{1}{z-1} \right) \quad (15)$$

with parameters:

**Table 3. PID parameters.**

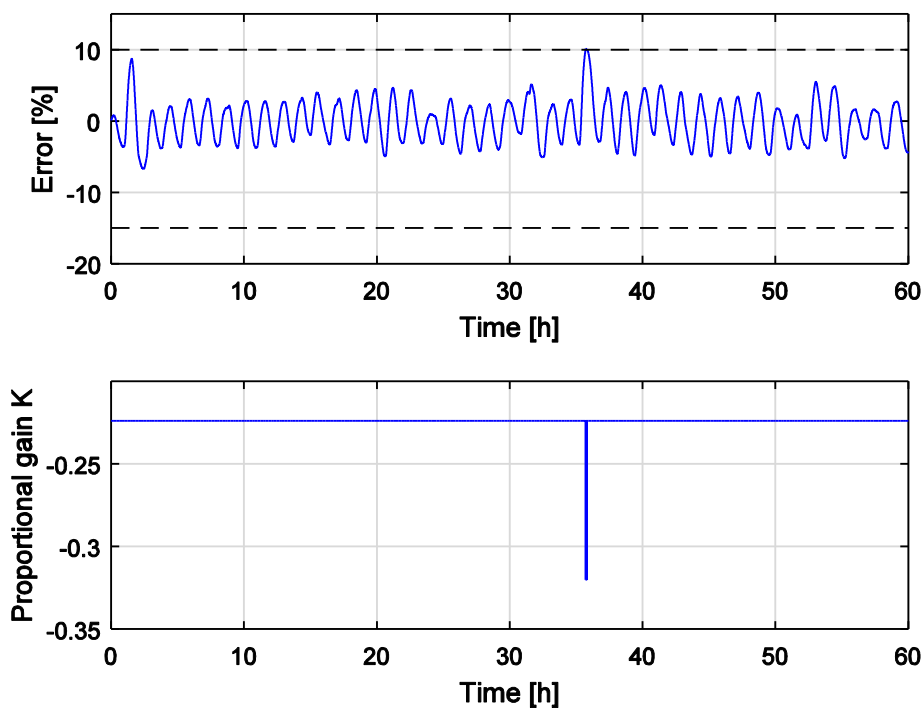
	LC124	FC128
<b>K</b>	0.400	0.476
<b>T<sub>i</sub> [min]</b>	80.00	1.35
<b>T<sub>s</sub> [s]</b>	15	15

Primary controller LC124 controls the height of level in T3 by calculating setpoint for secondary controller. In DCS this controller calculates output (OP) in the range 0-100%, but output flow is in the range 0-80 t/h. There are two options how to handle this fact. We can implement the conversion of OP in Simulink scheme, or we can divide  $K$  by process gain correction:

$$correction = \frac{100}{80} = 1.25 \quad (16)$$

This controller contains gain scheduling, which means it has two zones – fast and slow zone. The slow zone is defined 10% over and 15% under the setpoint where controller gain is

multiplied by factor 0.7. Gain scheduling has been achieved in Simulink using „IF“ block. Fig. 13 shows, how controller switches between proportional gains.



**Fig. 13. Gain scheduling.**

The secondary controller FC128 controls output flow by operating with a valve on output from T3. Setpoint for this controller generates primary controller. Performance under PID controller with Slovnaft parameters is shown in Fig. 14.

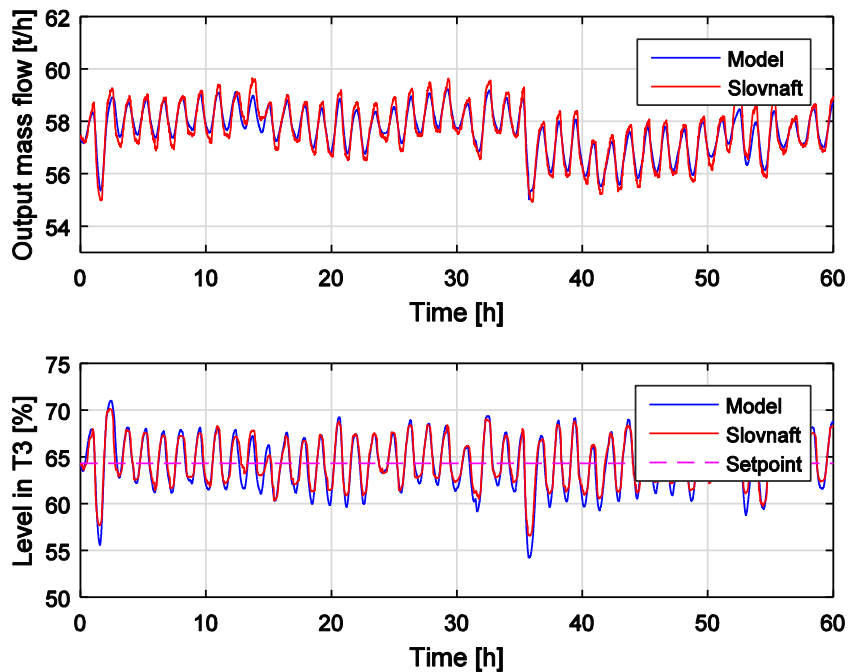


Fig. 14. Model and Slovnaft PID control performance.

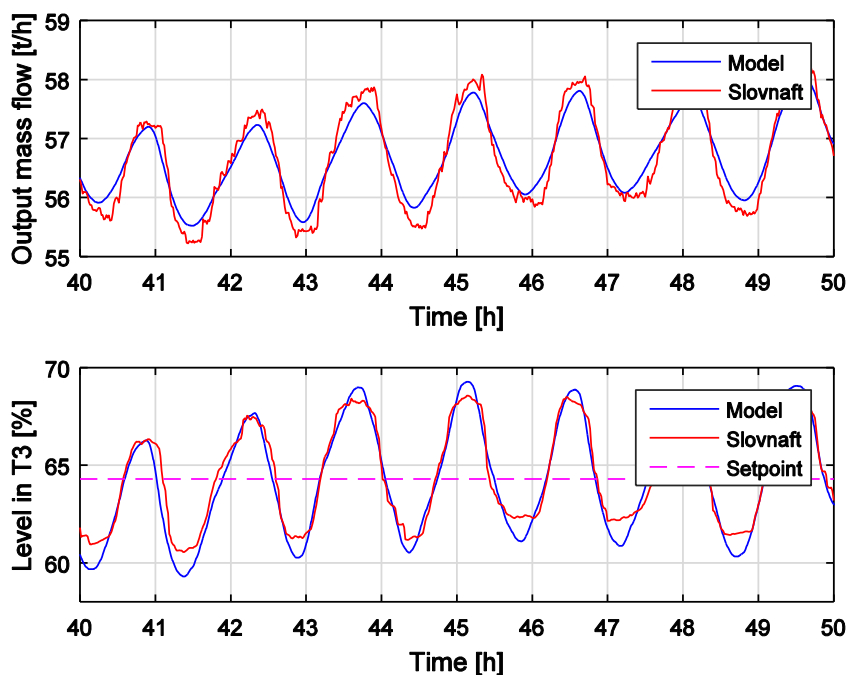


Fig. 15. Detail of control performance.

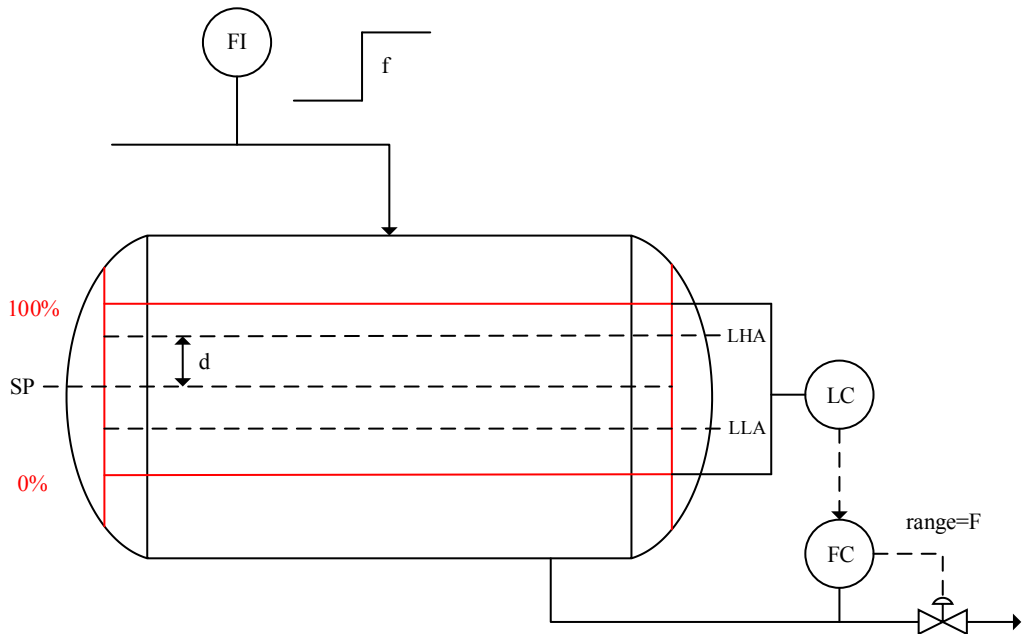


# 4 Level Control

The aim of this section of work is to design a controller which minimizes oscillations of the output flow from T3. In this chapter, we will try to propose such controller in the traditional way, ie. not using APC. There are two methods to design level controllers, tight and averaging level control. Tight control is for situations when it is more important to hold the level close to setpoint than maintaining a steadily manipulated flow. We use the second method, averaging control when it is more important to keep manipulated flow as steady as possible. This can be achieved by using all working volume of the tank without violating level alarms.

## 4.1 Parameters for Controller Design

The first step is to determine some necessary parameters as shown in Fig. 16.



**Fig. 16. Parameters for controller design.**

Working volume ( $V$ ) was determined in Chapter 2.1. Normally expected flow disturbance ( $f$ ) can be obtained from historical data. This disturbance is shown for example in Fig. 14. The smaller value of  $f$  should be chosen when all capacity of the tank is not used. To define how much of tank capacity can be used, we need parameter  $d$  which is a maximal deviation from setpoint to low and high-level alarm. The last parameter is  $F$ , defined as maximum output flow. All parameters are in Table 4 below.

**Table 4. Parameters for controller design.**

Parameter	Value	Unit
f	2.500	t/h
V	6.072	m <sup>3</sup>
SP	57.50	%
LHA	75.00	%
LLA	40.00	%
d	17.50	%
F	80.00	t/h

## 4.2 Averaging Level Control

We calculate the smallest possible controller gain (King, 2011):

$$K_{\min} = \frac{100f}{Fd} \quad (17)$$

As can be seen in Fig. 17, the level does not activate alarms. It was only one model disturbance, but we need return level to SP considering the next disturbance. For this case, we add integral action and modify controller gain as (18)

$$K_c = \frac{80f}{Fd}, \quad T_i = \frac{Vd}{12.5f} \quad (18)$$

Controller uses parallel form:

$$C(z) = K_c + \frac{K_c T_s}{T_i} \frac{1}{z - 1} \quad (19)$$

Input flow was increased as a step change. Manipulated flow was increased as slowly as possible and it took about 2 hours. Fig. 18 shows the performance of averaging control with designed controller.

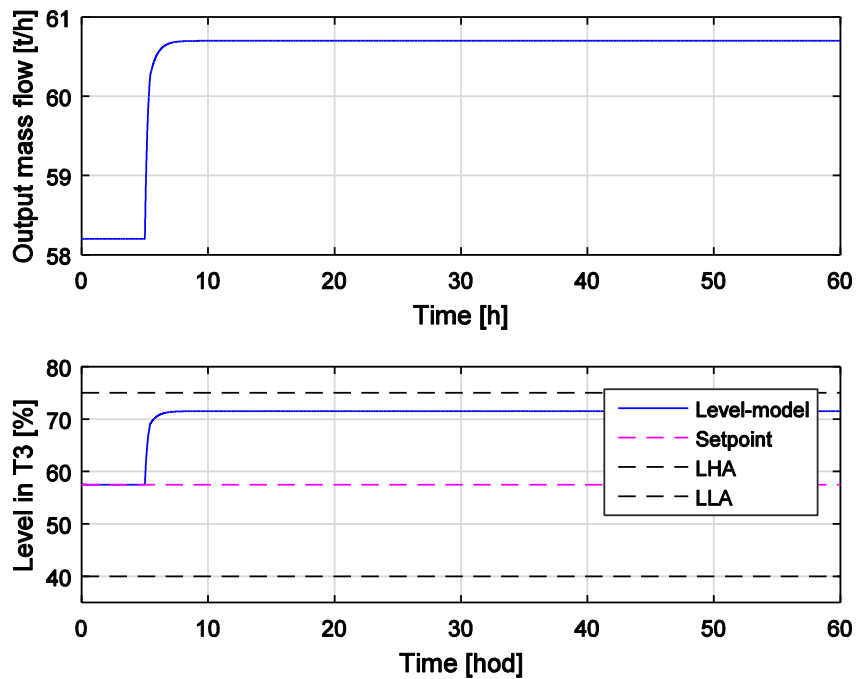


Fig. 17. Proportional control only.

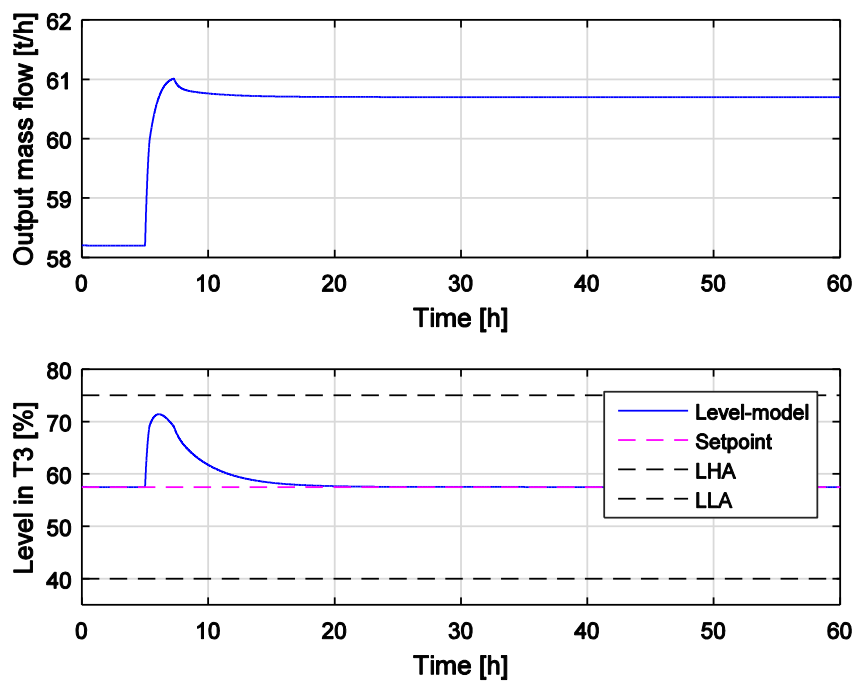


Fig. 18. Averaging level control.

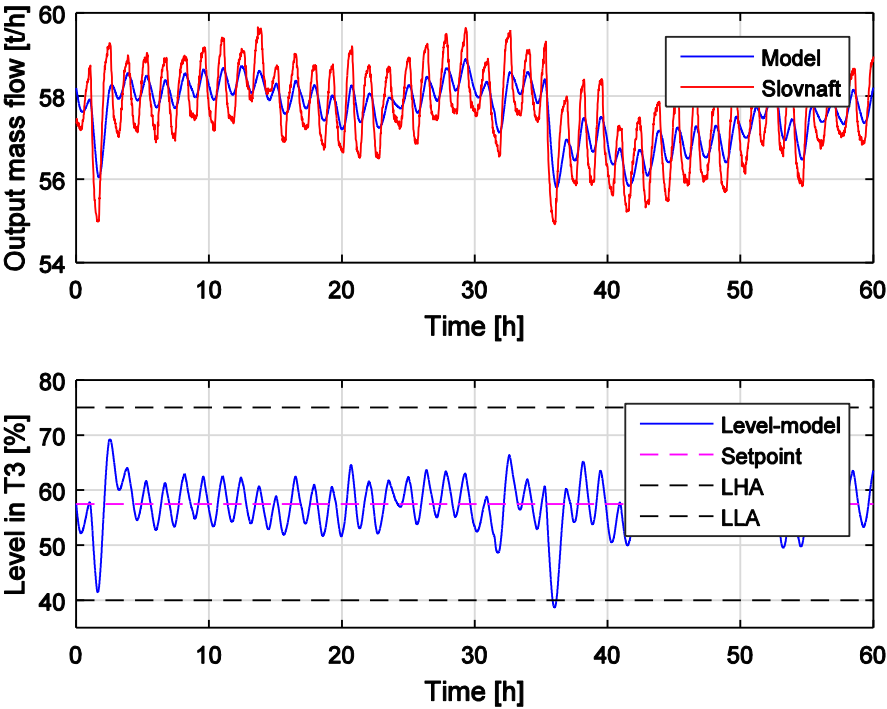
For better control performance, controller parameters were tuned. Both, calculated and tuned constants are in Table 5,

**Table 5. Calculated and tuned controller constant.**

Constant	Calculated	Tuned
P	-0.14	-0.08
I	-0.04	-0.10

where  $P = K_c$  and  $I = K_c/T_i$ .

As can be seen in Fig. 19 manipulated flow oscillates less using averaging control than using cascade in Slovnaft.



**Fig. 19. Averaging control with real data.**

## 4.3 Averaging Control with Gain Scheduling

In this section, we control level in T3 using a controller with gain scheduling. The first step was the design of discrete PI controller in the form:

$$C(z) = P \cdot \left( 1 + I \cdot T_s \frac{1}{z-1} \right) \quad (20)$$

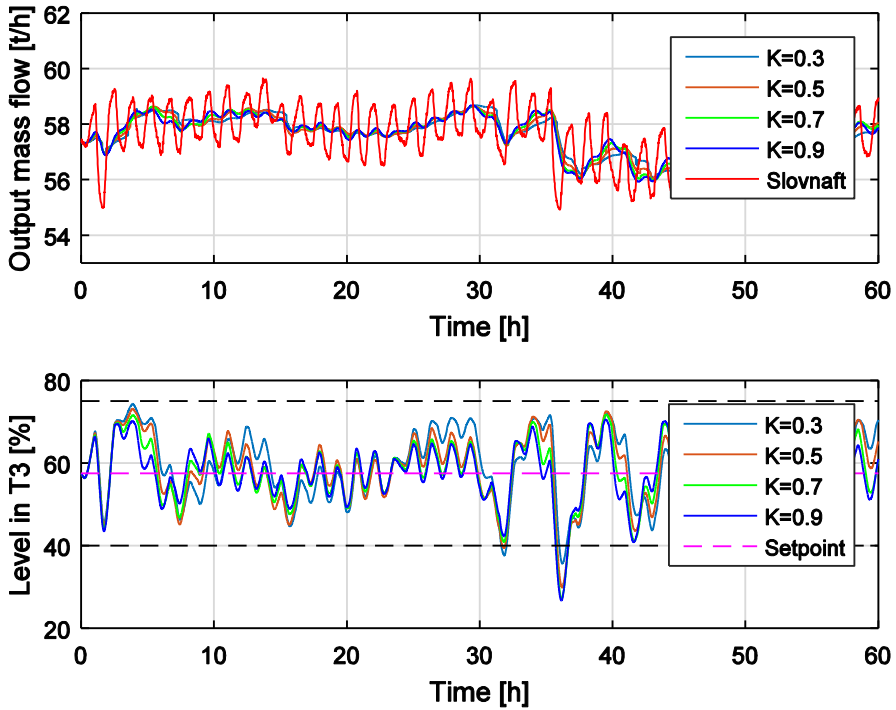
with parameters:

**Table 6. Designed PI controller.**

Constant	Value
P	-0.03
I	-2.00

We have same two zones like in Slovnaft control system. The slow zone is defined 10% over and 15% under the setpoint where controller gain is multiplied by factor K.

Level control with several values of constant K can be seen in Fig. 20.



**Fig. 20. Averaging level control with gain scheduling performance.**

Detail of control performance:

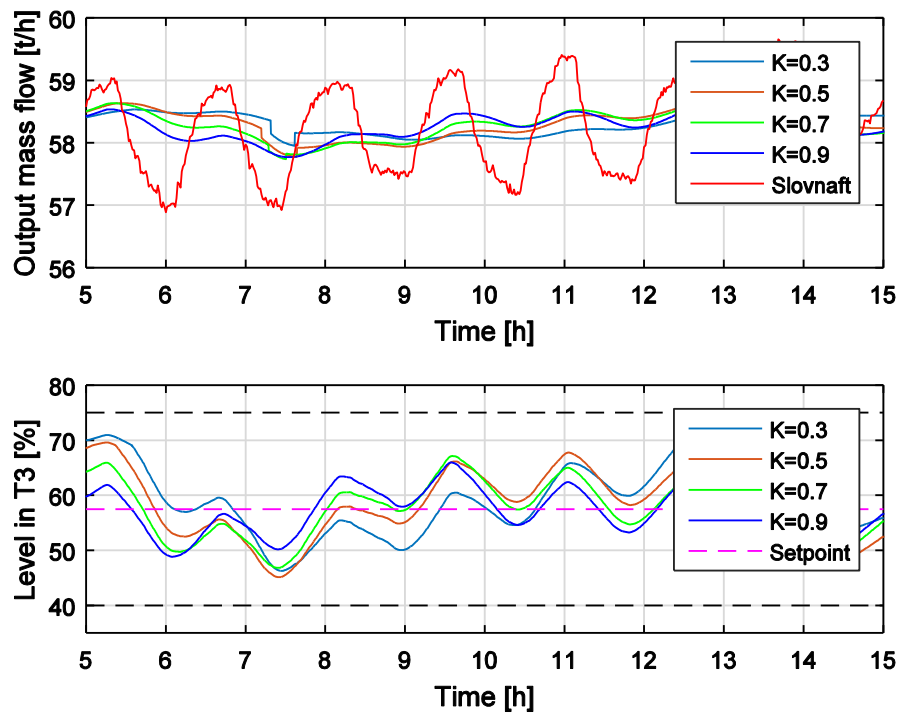


Fig. 21. Detail of control performance.

# 5 Percentage of Working Volume Control

## Control

This chapter is focused on the percentage of working volume control described in 2.1. Requested height of level in the tank is recalculated on the percentage of working volume of the tank and this is setpoint for the controller. At first, we tried it using Slovnaft control scheme. Results are in Fig. 22 and Fig. 23.

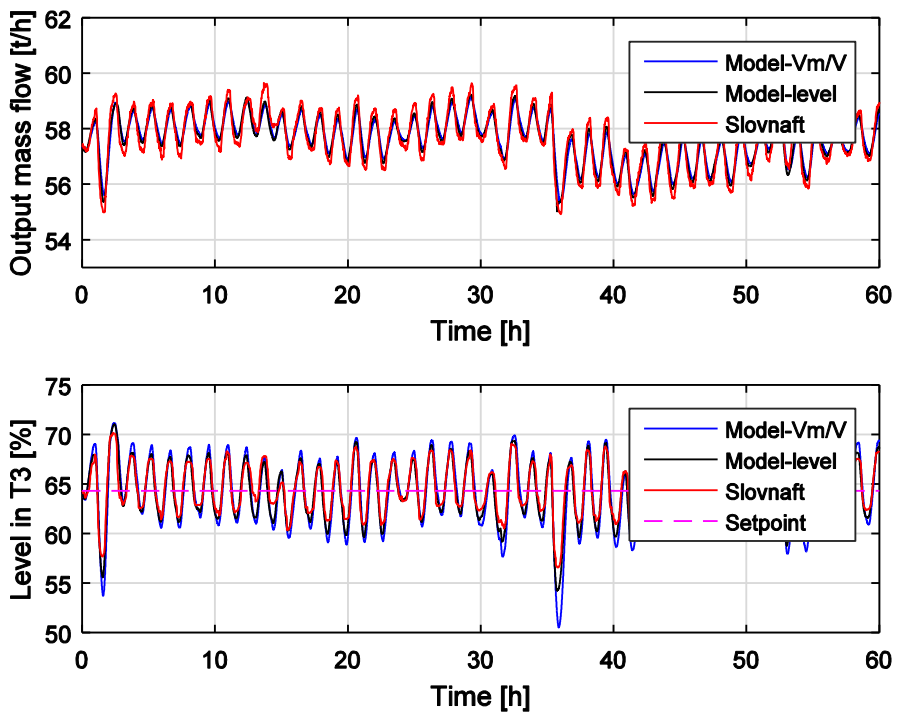


Fig. 22. The performance of percentage of working volume control – Slovnaft.

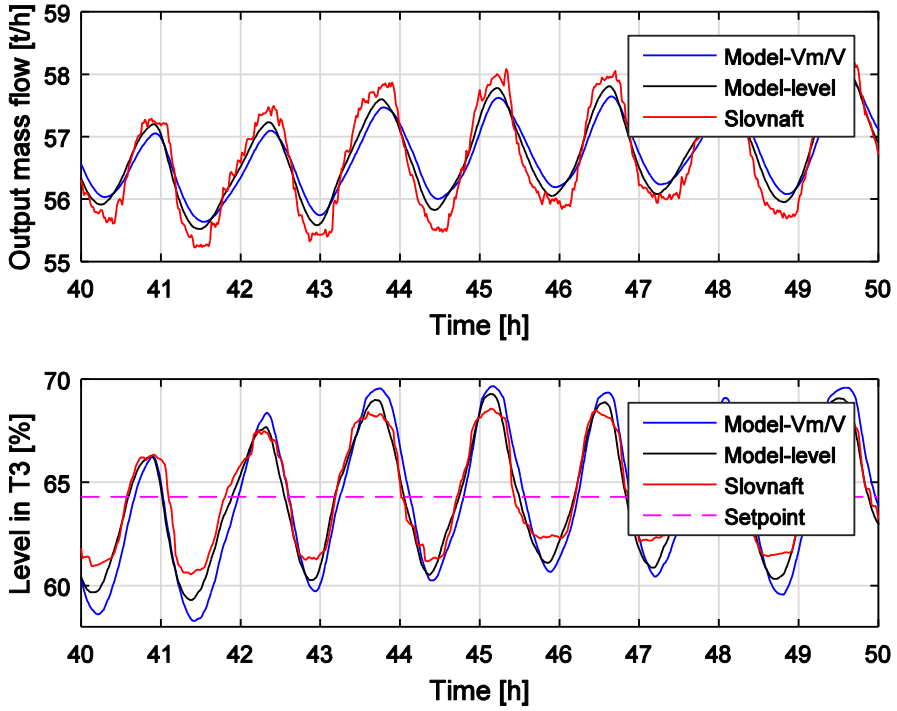


Fig. 23. Detail of control performance.

We also used designed averaging controller from chapter 4.2 and controlled percentage of working volume. By comparing level control and percentage of working volume control we can see that there are only minimal improvements using second mentioned method. The reason is simple. Main nonlinearity which occurred in level indication was removed by level sensors placement. Comparison of these two methods is in Fig. 24 and Fig. 25. A varying percentage of working volume is in Fig. 26.



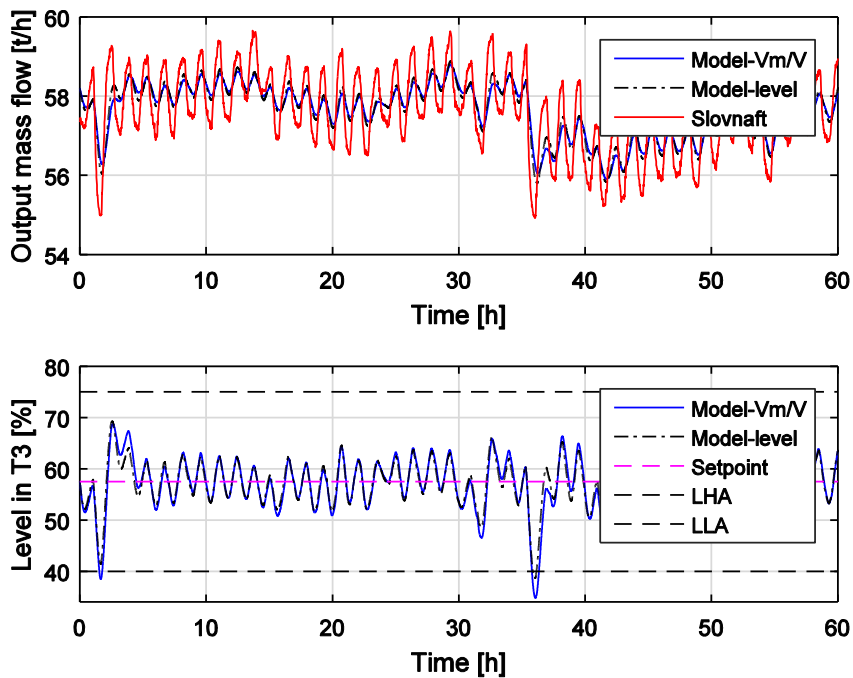


Fig. 24. The performance of percentage of working volume control – averaging control.

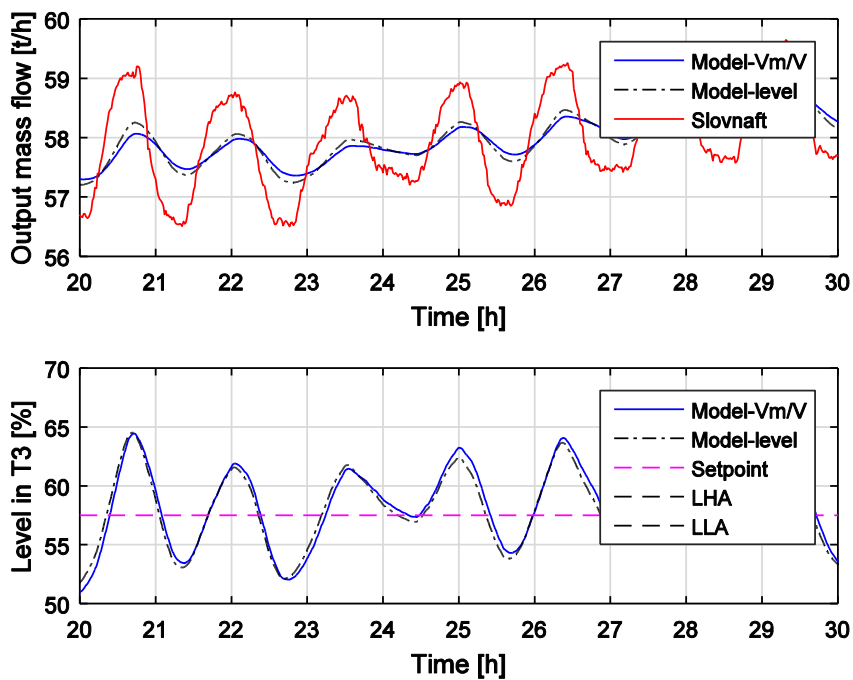
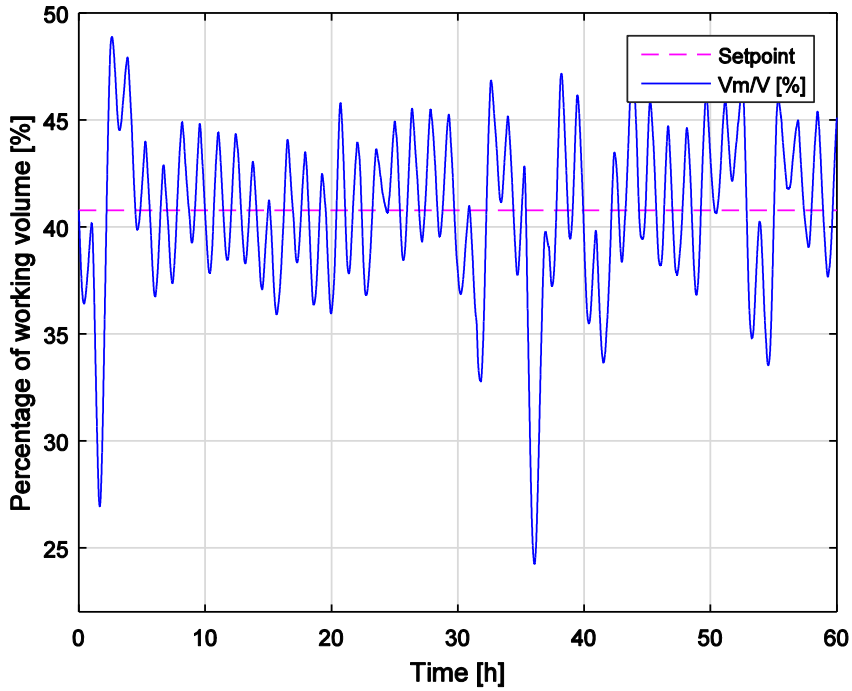


Fig. 25. Detail of control performance.



**Fig. 26. Percentage of working volume.**

To compare which method gives us the smallest oscillation of output flow from T3, we calculated the sum of squares:

$$\sum_{i=1}^K (\dot{m}_{i,\text{out}} - \text{ref}_i)^2 \quad (21)$$

where  $\text{ref}$  is reference curve vector shown in Fig. 27 and  $K$  is a number of elements of reference curve. Calculated values are in Table 7 below. Using averaging level control was a significant improvement against Slovnaf controllers. Little improvement has been achieved using percentage of volume instead of level.

**Table 7. Control methods comparison.**

Control method	Sum of squares
Slovnaf controllers	9684
Averaging control - level	2212
Averaging control – percentage of volume	1749

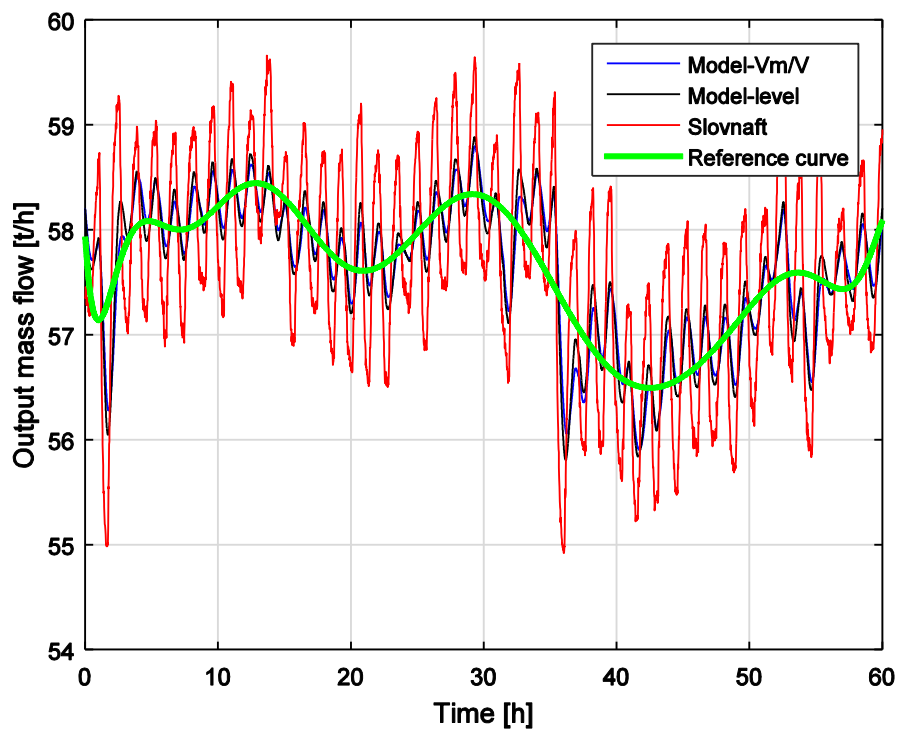


Fig. 27. Reference curve.

# 6 Advanced Process Control - RMPCT

Advanced process control (APC) is a wide range of techniques implemented in industrial process control systems. One of them is multivariable model predictive control. Honeywell has developed his own application for control of multi-input, multi-output interactive industrial processes called Robust Multivariable Predictive Control Technology (RMPCT). (Honeywell, 01/2012)

Profit controller (RMPCT) provides control and optimization subject to time varying constraints. This controller contains a dynamic model of the process from which it predicts future behavior and determines, how controller's outputs should look like to achieve setpoints of all process variables. Sometimes it is sufficient when process variables are within constraints. RMPCT can achieve good control results even though there are errors in the process model. (Honeywell, 10/2012)

## 6.1 Controller Variables

We know three types of variables which uses Profit controller to control the system:

- **Controlled Variables (CVs):** controller tries to keep these variables on a setpoint or within a range, but in both cases, the constraints have to be satisfied.
- **Manipulated Variables (MVs):** these variables are changed by the controller to achieve required results with CVs. Of course, all MVs can be changed only within constraints.
- **Disturbance Variables (DVs):** these variables are measured, but they are not controlled. They may come from upstream processes and affect CVs. Prediction of the effects of DVs on CVs can help keep CVs within constraints. (Honeywell, 10/2012)

Using a SISO controller there is only one CV and one MV. The controller looks only at one loop and does not care about other loops in a system. With RMPCT there are multiple MVs and multiple CVs. The controller views all variables together as a system and considers effects of all interactions between MVs and CVs.

## 6.2 Models

Model of the system that uses Profit Controller is built from dynamic sub-processes models which contain information about behavior between independent variables MVs, DVs and CVs. Sub-process models are null when there is no physical connection between MV or DV and CV.

Profit controller (RMPCT) uses a generic form of sub-process model, which contains a number of coefficients whose values we need to find to identify the model. We want to achieve predicted responses as close as possible to actual process responses.

## 6.3 Objective Function

To keep CVs on a setpoint or within a range, we need sufficient degrees of freedom. Degrees of freedom is a number of MVs which are not at a limit minus number of CVs which have setpoints or which are out of the limits. If the number of degree of freedom is positive or zero, all constraints of CVs can be satisfied. If the number is negative it is impossible to achieve CV's setpoint or keep CVs within a range.

There are some cases when we have more CVs than MVs and it is possible to achieve setpoints or keep CVs within the range. In these cases, it should be there a sufficient number of CVs which have only a range instead of the setpoint. (Honeywell, 01/2012)

As an objective function, we consider the linear or quadratic function of any CVs and MVs or of all of them.

General form:

$$\min J = \sum_i b_i \cdot CV_i + \sum_i a_i^2 \cdot (CV_i - CV_{0,i})^2 + \sum_j b_j \cdot MV_j + \sum_j a_j^2 \cdot (MV_j - MV_{0,j})^2 \quad (22)$$

where:

**b** is the linear coefficient

**a** is the quadratic coefficient

index **i** belongs to the CVs

index **j** belongs to the MVs

**CV0, MV0** are the desired values

In our case, we consider only quadratic function, because we have setpoint. In the case of control between limits, we would use the linear function.

# 7 Identification and Controller Design

The major aim of this thesis was to propose and implement APC which will coordinate flows and levels in given tanks systems in order to mitigate redistillation column feed fluctuations. Following chapters describe steps of implementation of given APC and its evaluation.

## 7.1 Data Preparation

For the identification of tanks system using Honeywell Profit Design Studio, we need to get model data from step responses. These step responses were obtained from Matlab – Simulink simulations. Before running any simulation or identification we need to classify variables to controlled, manipulated and disturbance variables as you can see in Fig. 28.

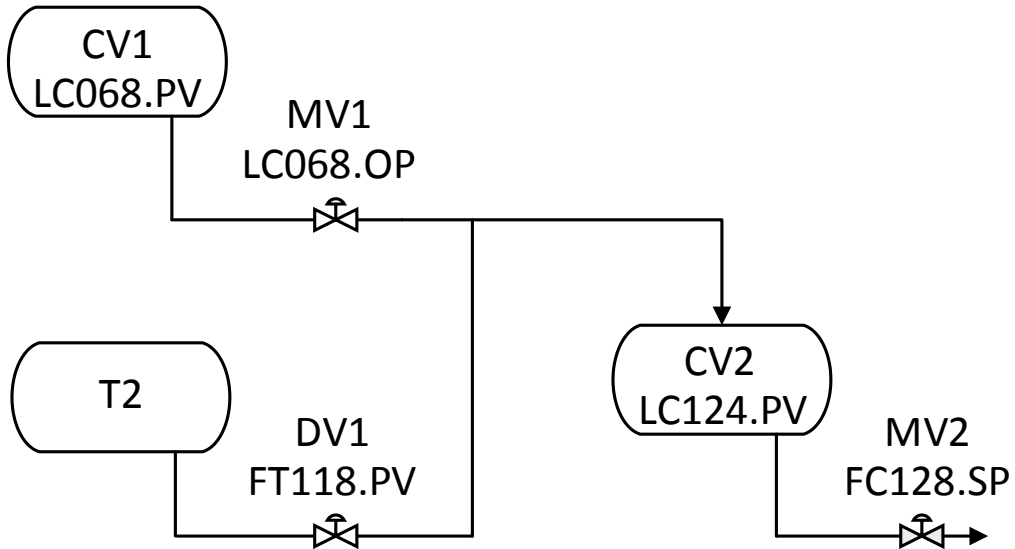


Fig. 28. Classification of variables for APC.

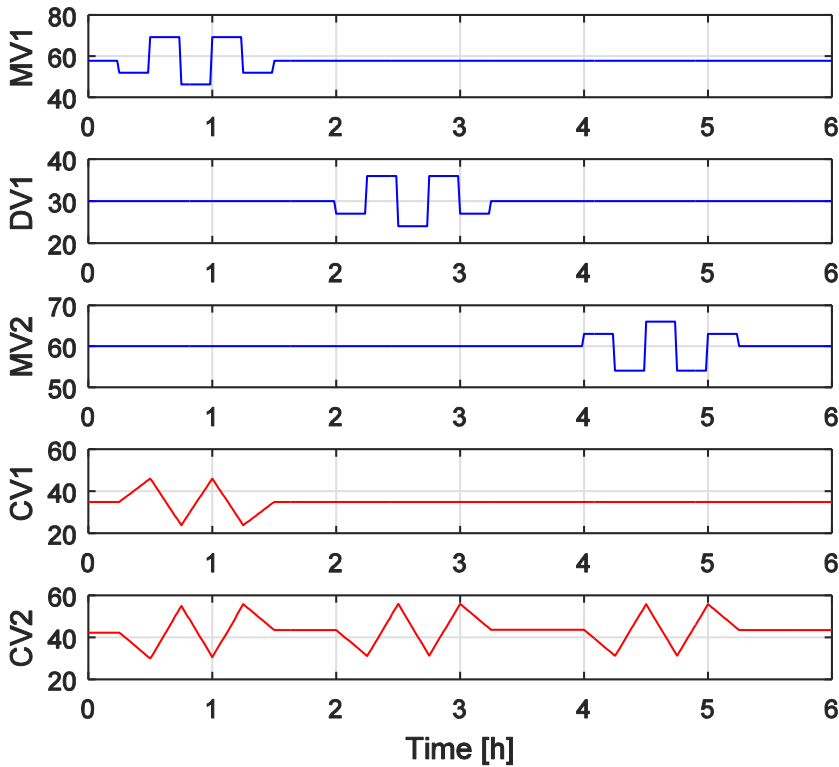
We have two controlled variables (CVs) – percentage of working volume in T1 and T3, two manipulated variables (MVs) – valve opening from T1 and output flow from T3 and one disturbance variable (DV) – output flow from T2. Last mentioned disturbance variable can not be manipulated variable because the valve is sticking. We use a percentage of working volume CVs instead of raw levels, since this ensures us linear model between MVs and level CVs. Instead of measured output flow from T1, we use a percentage of opening of the valve. Between valve

opening and flow through there is a linear relationship calculated from LC068.OP vs FI076.PV process data:

$$\text{Output flow} = 0.52 * \text{Percentage of opening}$$

After classification of variables, we can achieve step responses. Since the percentage of working volume (CVs) have linear models to our MVs and we had all necessary data, we could calculate responses and models theoretically. However in order not to skip Profit Suite identification we decided to use Matlab-Simulink simulation to get step test data.

The time period for step responses data is one minute. Step responses were made in respect to not violating alarms in the tank. Results are shown in Fig. 29. We made steps up and down due to unstable process. The process will not stabilise by itself after one step change.



**Fig. 29. Step responses.**

The data were given into Excel template in the form shown below.

	A	B	C	D	E	F	G
1	Date	Time	LC068.PV/Var/CV/T1 level/% T3 level/% ut flow/% low/TpHR ow/TpHR				
2	11/03/2016	5:00:00	34,92410125	42,2461	57,69231	60	30,00
3	11/03/2016	5:01:00	34,92	42,25	57,69	60,00	30,00
4	11/03/2016	5:02:00	34,92	42,25	57,69	60,00	30,00
5	11/03/2016	5:03:00	34,92	42,25	57,69	60,00	30,00
6	11/03/2016	5:04:00	34,92	42,25	57,69	60,00	30,00

Fig. 30. Data preparation for PDS.

## 7.2 Identification

Now we can identify system using Honeywell Profit Design Studio (PDS). We open the PDS and choose **File => New => Model Dev. File** (Fig. 31).

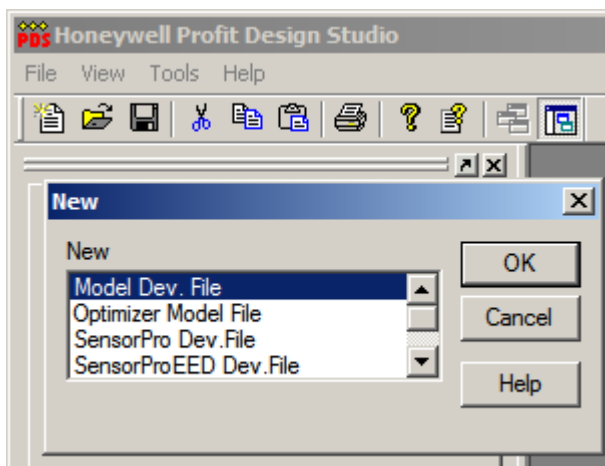


Fig. 31. Creating anew model.

Mark all data in created Excel file including headings and select **Data operations => Paste from clipboard**. After this, we can see the window shown below (Fig. 32).



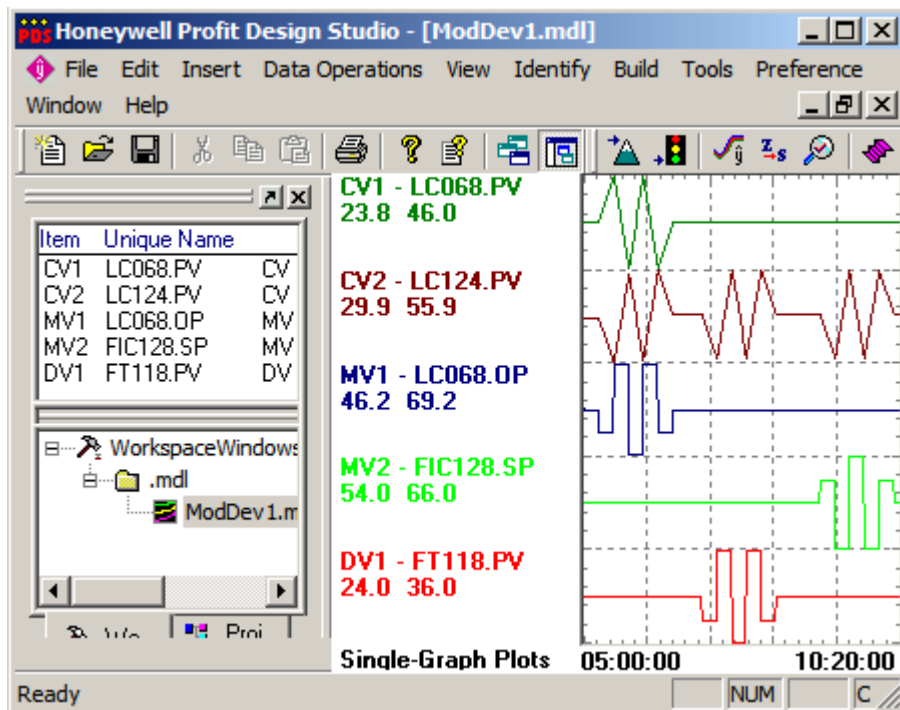


Fig. 32. Data imported in PDS.

The output from identification is a CVs - MVs/DVs model matrix. We use Finite impulse response (FIR). FIR step response models are obtained by integrating the finite impulse response coefficients. These models represent the response of a dependent variable (CV) to a step change made to an independent variable (MV or DV). Select **Identify** => **Set Overall Options** and mark **Finite Impulse Response (FIR)**. In this window, we can also specify the number of trials. The trials are partial models in case if we need only a part of all dynamic. For better illustration see Fig. 33 from another process identification.

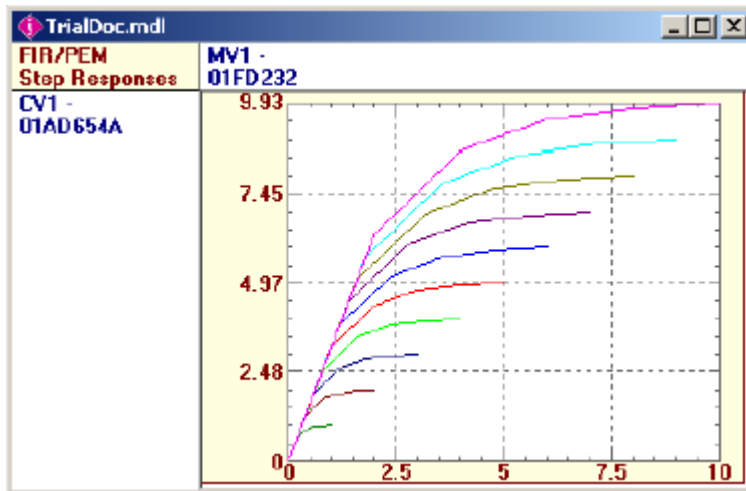


Fig. 33. Trials.

The number of the trials is a number of continuous models for one given CV – MV/DV pair. In our case, we chose three trials with parameters shown in Fig. 34.

To configure FIR select **FIR Setup**. Using drop-down list box we can configure maximal settling time for each trial. We can also define a number of coefficients to represent the curvature of the step response curve.

Selecting the **Options** button will display more options. All displayed parameters we leave as default except one. Deselect **Penalize Oscillations**.

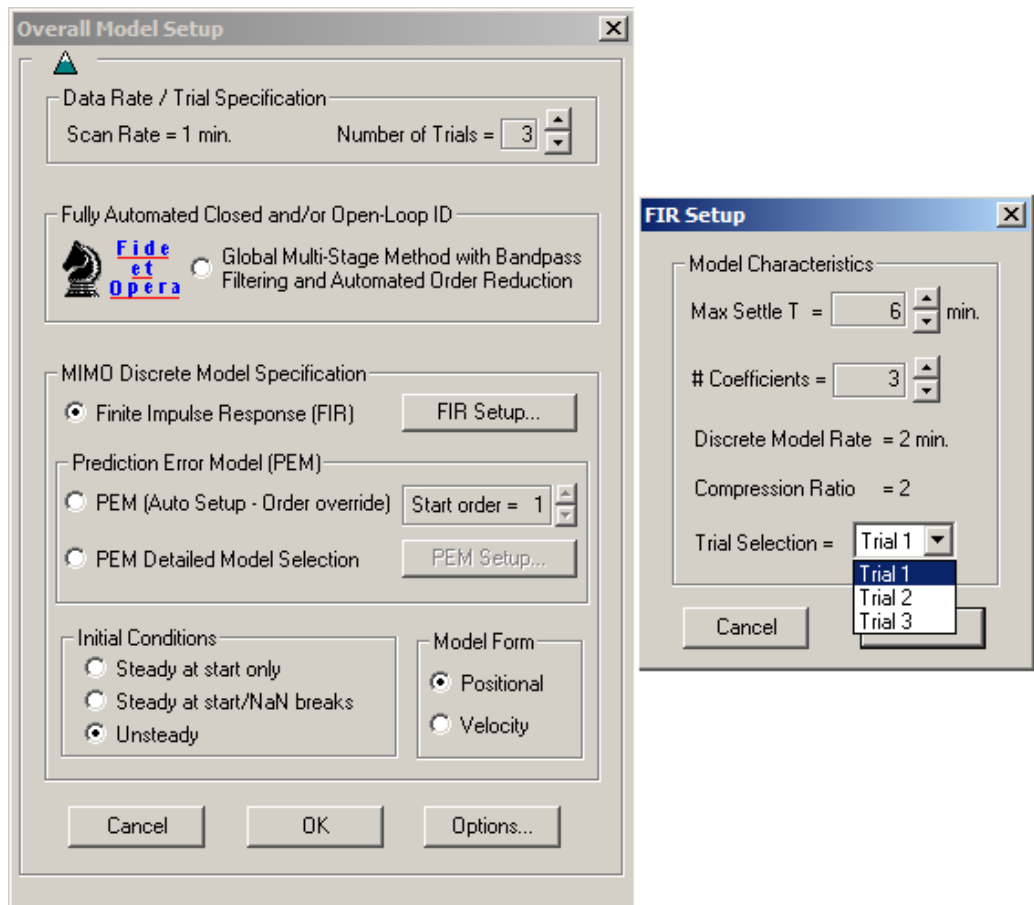


Fig. 34. Overall parameters.

After setting overall options the identification can begin. Select **Identify => Fit FIR/PEM/CLid Models**. Double click on sub-process will open dialog box **Options per Submodel** (Fig. 35). If there is no physical way, how independent variable – MV/DV can affect CV, mark **Null Sub-Process**. In all other sub-process, we mark **Integrating Sub-Process**.

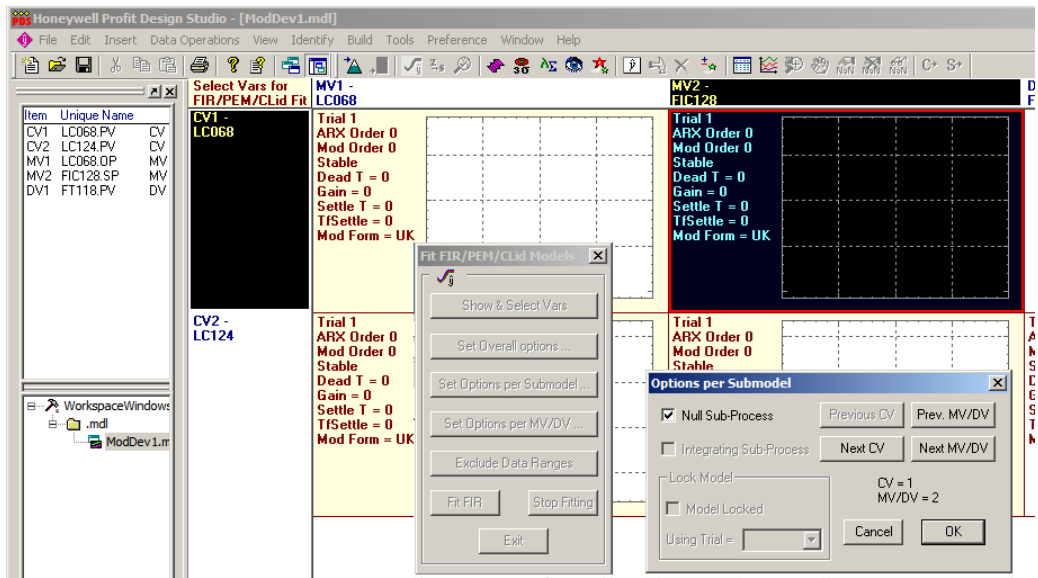


Fig. 35. Fit FIR/PEM/CLid Models.

After setting all necessary options we can select **Fit FIR** (Fig. 36). It can be seen all three trials for every sub-process.

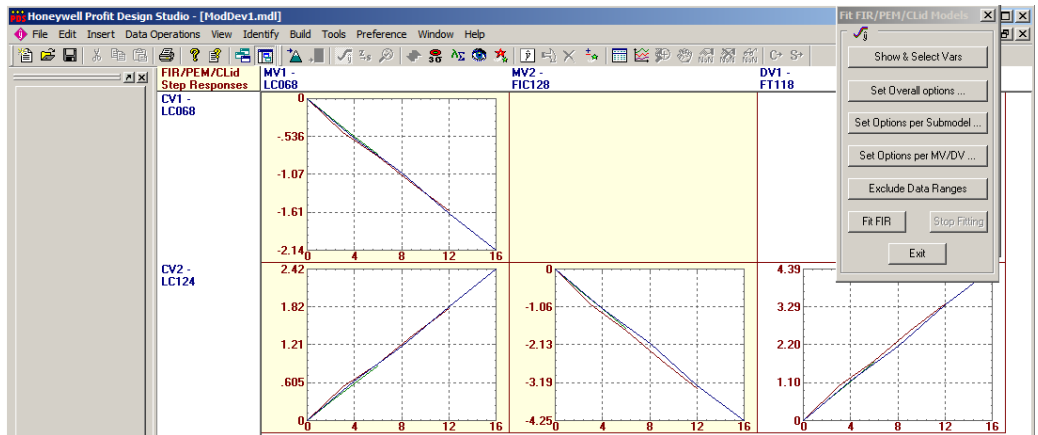


Fig. 36. Fitted FIR.

Select **View** => **Model summary**, mark all models and click **Edit** => **User2Final**.

Select **Identify** => **Fit Parametric Models** and mark Laplace method (Fig. 37). By clicking on sub-process we can select final trial and edit identified transfer function. Finally, we have identified model matrix of tanks system (Fig. 38).

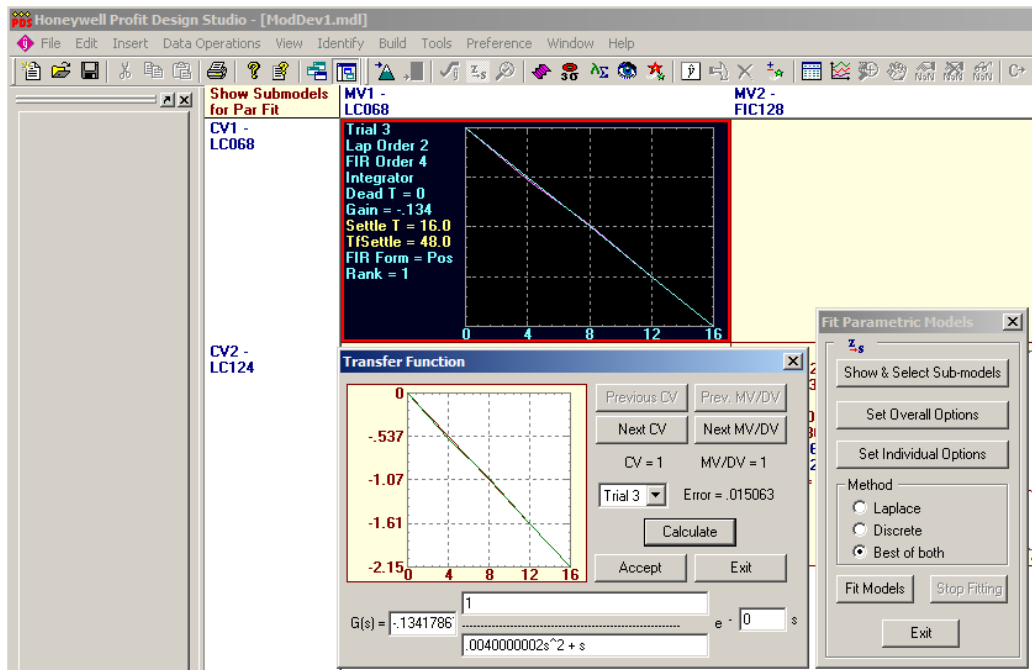


Fig. 37. Editing sub-processes.

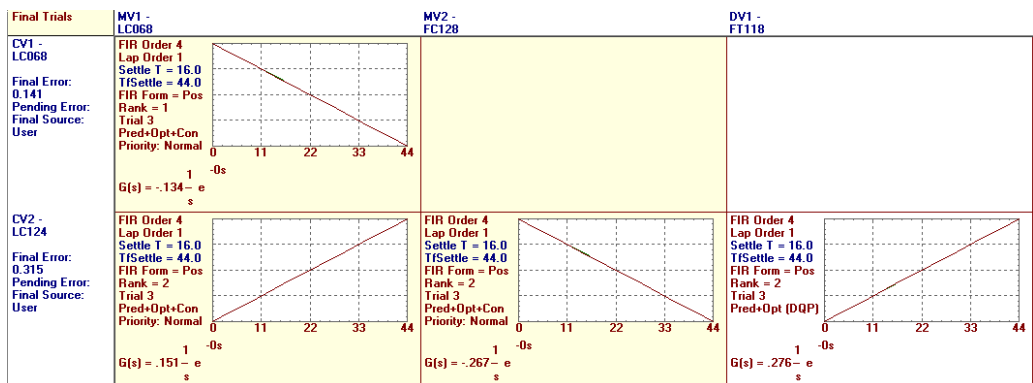
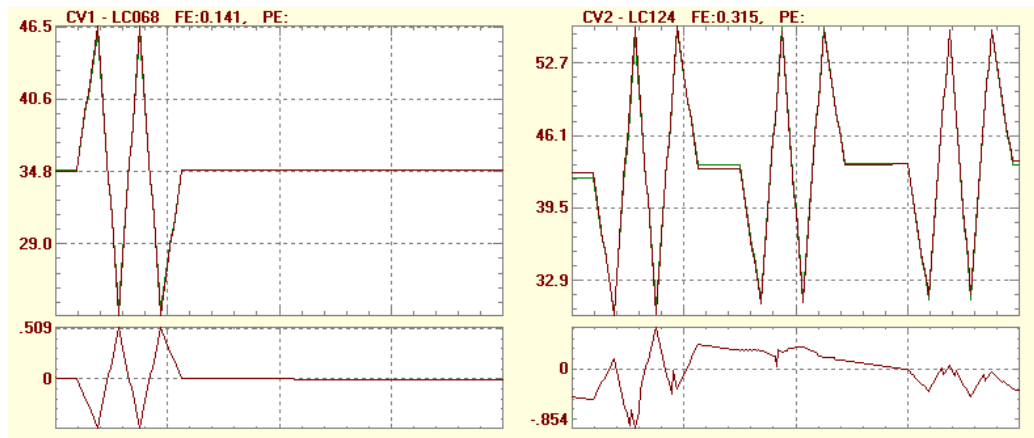


Fig. 38. Final trials.

We can illustrate how accurate is our identification by comparing measured CVs with predicted CVs from identified models: **Identify** => **Select Final Trials**, enable **Store Predictions** and select **Plot Predictions**. The result is in Fig. 39.



**Fig. 39. CVs predictions.**

The red line represents measured step responses and the green line is predicted step responses. As it was mentioned before, it was possible to get models theoretically. Since models are linear and we know capacity of the tank, slope of the integrator is easily calculated as the ratio of step change per minute to the capacity of the tank. Models obtained by calculation and by identification were in coincidence, so based on this and prediction check we can consider, that models were identified correctly.

## 7.3 Building Controller

Next step is building the controller. At first, we need to build Unified Real Time Platform (URT Platform). This platform serves to implement real-time APC applications. In PDS we choose **View => Model Summary**, mark all models and select **Edit => User2Final**. Now we can start with the controller: **Build => Controller**. It is necessary to generate all setting files choosing options shown in Fig. 40. Generating setting files. Fig. 40.

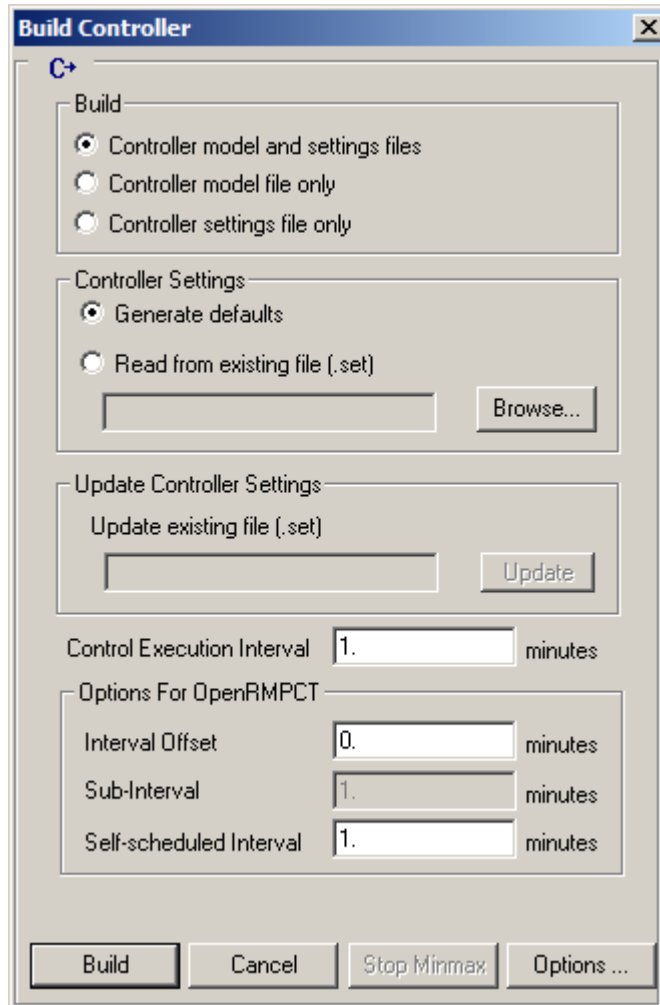
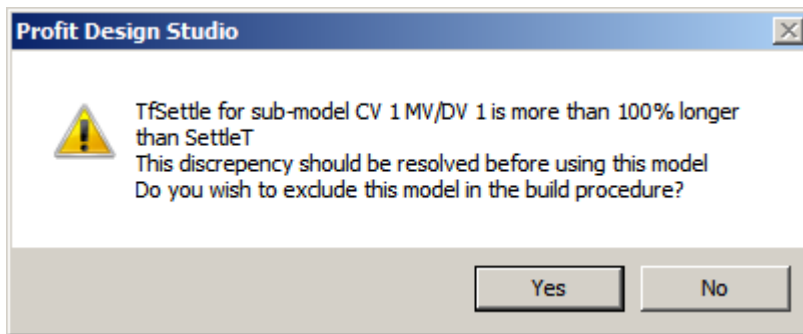


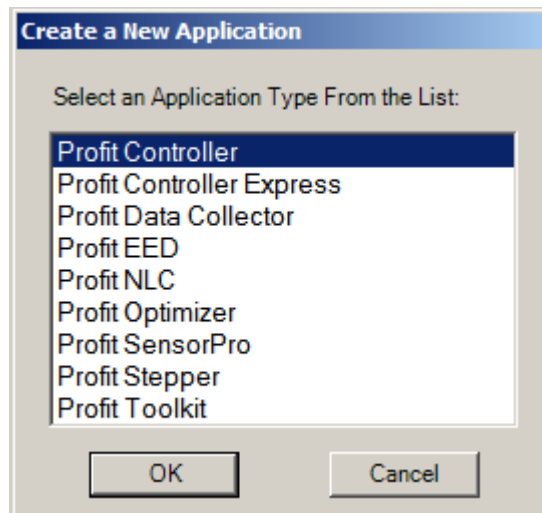
Fig. 40. Generating setting files.

Select **Build**, if an error window appears, just click **No** (Fig. 41).



**Fig. 41. Error window.**

PDS generates setting files. These files we move to C:\ProgramData\Honeywell\URT\Platforms. Open Profit Suite Runtime Studio (PSRS): **File => New => Profit Controller => Ok** (Fig. 42).



**Fig. 42. Create a New Application.**

We describe general information shown in Fig. 43, choose .xm .xs files generated in the previous step and .mdl file with our model and click **Ok**.



**New Profit Controller Application** [?] [X]

**General Info**

Name: BCDU61L1

Desc: Controller

Area: URT

**Application Files**

xm file: JHoneywell\URT\Platforms\BCDU61L1.xm ...

xs file: \Honeywell\URT\Platforms\BCDU61L1.xs ...

xp file: ...

☒ mdl file: Honeywell\URT\Platforms\BCDU61L1.mdl ...

☐ xml file: ...

ccf file: ...

**Import Controller Tunings from PSES Simulation**

SimSet: ...

urt file: ...

**Extension Options**

☐ Use Profit Simulator ☐ No BLC

**Default DCS Box & OPC Server for Process Connections**

DCS Box: Experion PKS C-Series ▼

Server: HWHsc.OPCServer ...

OK Cancel

**Fig. 43. General information.**

In a **Controller** section, we describe information as shown in Fig. 44. Again we choose .mdl file with our model.

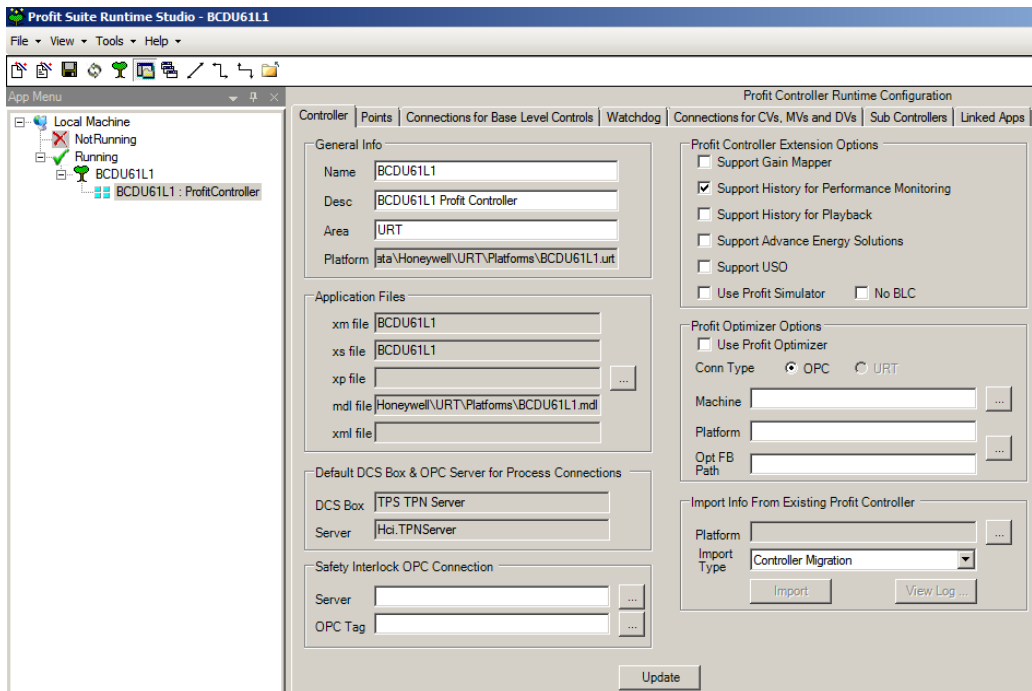


Fig. 44. Controller.

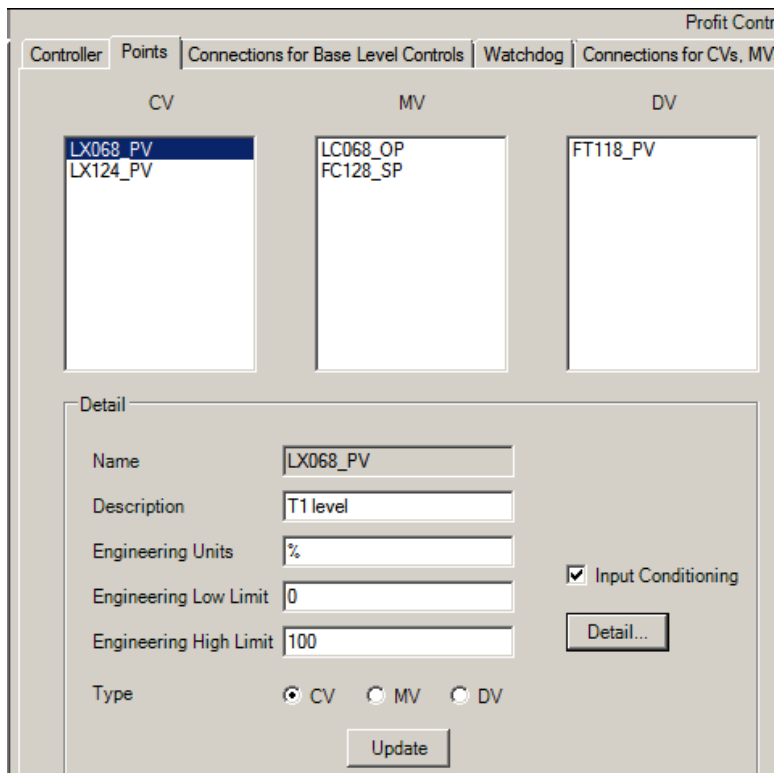


Fig. 45. Points.

Section **Points** (Fig. 45) is about variables properties. We define engineering units and range of the variable. Instead of the level, we use a percentage of working volume to calculate output from the controller. This is a reason why we named our CVs LX instead LC. Part of the **Points** section called **Detail** we set it as in Fig. 46.

**Input Conditioning Configuration**

**Input Validation**

☐ Enable Validation

High Limit: 100

Low Limit: 0

ROC Up Limit: 100

ROC Down Limit: 100

Freeze Tolerance: 0

Hold Time (minutes): 0

Freeze Hold Time (minutes): 0

Max Reset Time (hours): 24

☐ Clamp Input

**Input Filtering**

☐ Enable Filter

Filter Time (minutes): 0

**Asynchronous CV Processing**

☐ Enable New Value Monitoring

OK Cancel

Fig. 46. Details.

**Connections for Base Level Controls** is about what to do with MV's controller in a case of shutting down the APC. MV1 switches to the automatic and MV2 switches to the cascade control (Fig. 47).

Profit Controller Runtime Configuration

[Controller](#) | 
 [Points](#) | 
 [Connections for Base Level Controls](#) | 
 [Watchdog](#) | 
 [Connections for CVs, MVs and DVs](#) | 
 [Sub Controllers](#) | 
 [Linked Apps](#)

Var	BLC Name	BLC Template	OPC Server	Prim...	
MV 1	LC068_OP	HW_LCN_Ctrl-PMan_Shed-OAuto_TPNServer	Hci.TPNServer	Primary	
MV 2	FC128_SP	HW_LCN_Ctrl-PAuto_Shed-OCas_TPNServer	Hci.TPNServer	Primary	
CV 1	LX068_PV	HW_LCN_CV_PV_TPNServer	Hci.TPNServer	Primary	
CV 2	LX124_PV	HW_LCN_CV_PV_TPNServer	Hci.TPNServer	Primary	
DV 1	FT118_PV	HW_LCN_DV_PV_TPNServer	Hci.TPNServer	Primary	

Base Level Control Detail

Variable:

BLC Name:

DCS Box Type:  WD LCN Unit:  Max Point Chars:

BLC Template:

OPC Server:

Target:

**Fig. 47. Connections for Base Level Controls.**

Now we have built **Watchdog**, which is a kind of small program to check a communication between APC and our target. We save it like URT Platform, which is ground for APC coordination.

## 7.4 Controller Calculations

We start **URT Explorer** and in a left side, there are all platforms (Fig. 48). If our platform is not activated we can do it by double click on **BCDU61L1** => **Start Platform**.

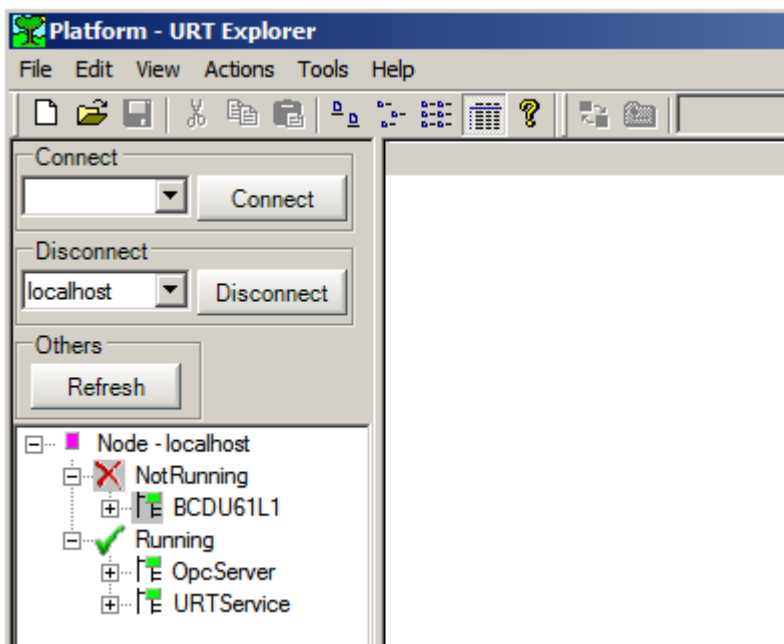


Fig. 48. URT Explorer.

Double click on **BCDU61L1** =>expand the menu =>right click on an **OpcDAClient** => **Append New Item** => **Data Item** => **node list** and we name it **Calc** (Fig. 49). In this block, we will write calculations between level and percentage of working volume.

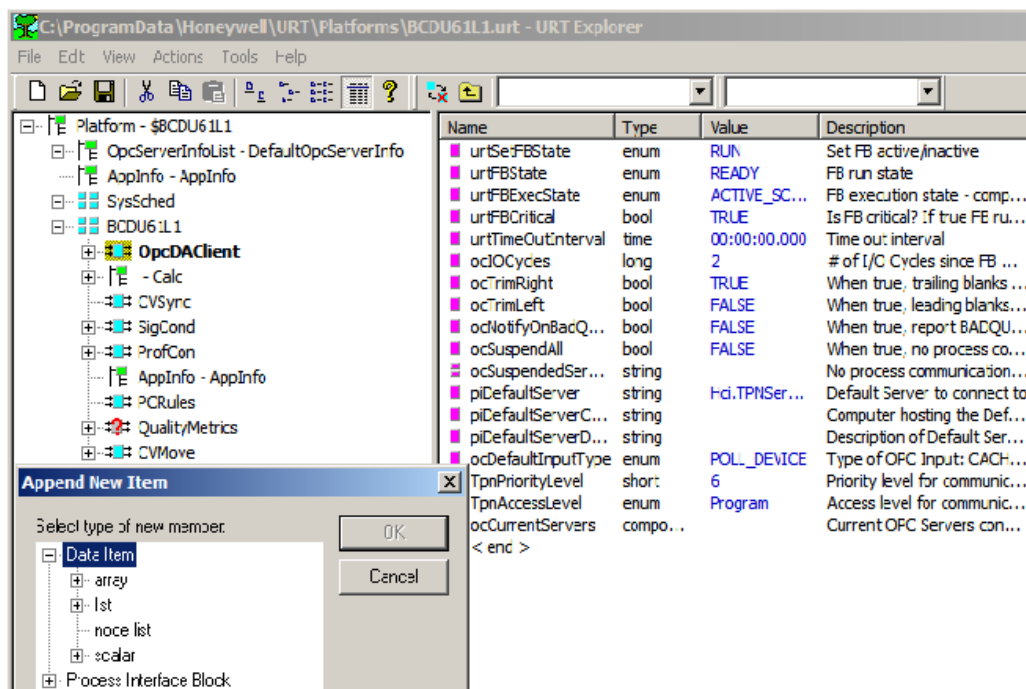
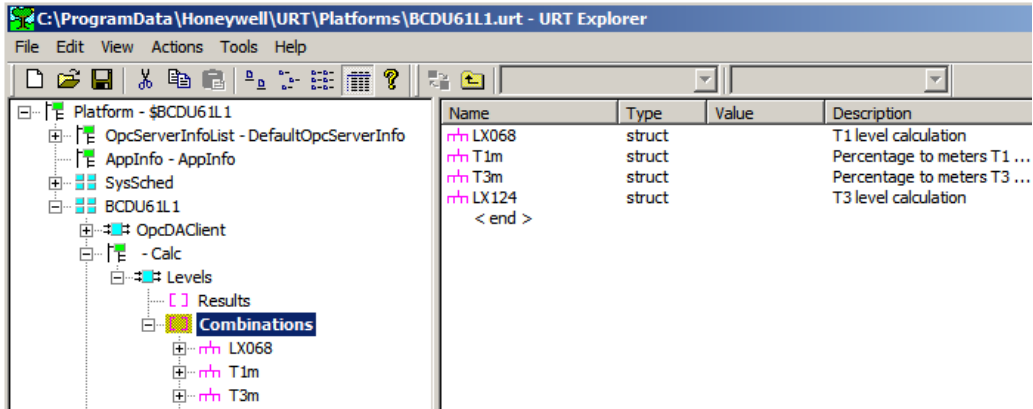


Fig. 49. New item.

Right click on **Calc** => **Append New Item** => **Profit Suite Block** => **Profit Toolkit** => **Variable Combinations** and we name it **Levels**. Now we have block for calculations. We need four calculations. Two of them are for the level in percentage to level in meters calculation and the rest two are for calculation of the percentage of working volume from a level in the tank. Right click on **Combinations** => **Properties** => **Value** => **Working size = 4**. The result is in Fig. 50.



**Fig. 50. Combinations block.**

At first, we need to define T1m and T3m. Inputs to these blocks are values of levels in tanks T1 and T3 in percentage. The output from these blocks are levels in meters. I will describe the setup of combinations blocks on LX068 calculation. The result from this block is the percentage of working volume in tank T1. The input to this block is level in meters, which is also result from T1m block. Basically, we need to insert calculations from Chapter 2.1, equations (9),(10)(11). At first, we define a number of variables necessary for the calculation. **LX068** => **Number of Variables** => **Value** => **Working value =8**, (Fig. 51). Variables are described in Table 8, (Fig. 52).

**Table 8. Calculation variables.**

Name	Description
IV1	Level in the tank
IV2	Radius of the tank
IV3	Length of the tank
IV4	Length of gasoline side
IV5	Volume of unuseful gasoline part
IV6	Volume of water side
IV7	Total useful volume
IV8	Height of barrier

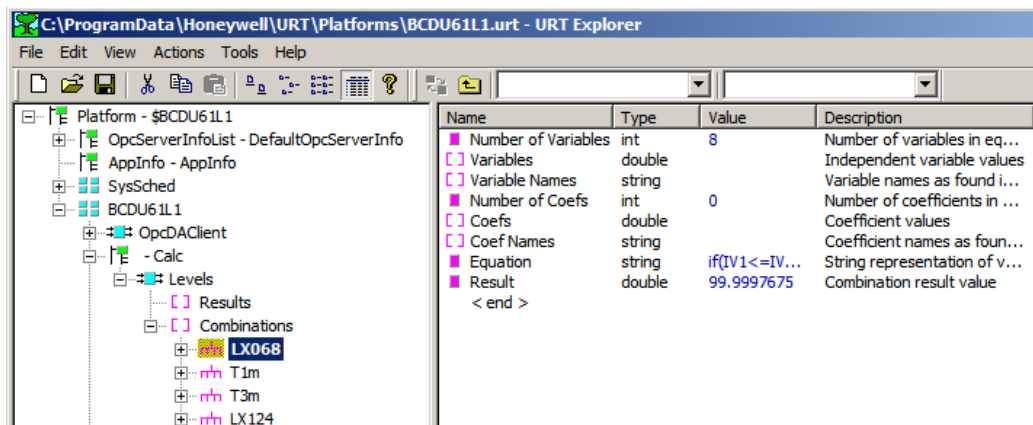


Fig. 51. T1 level calculation block.

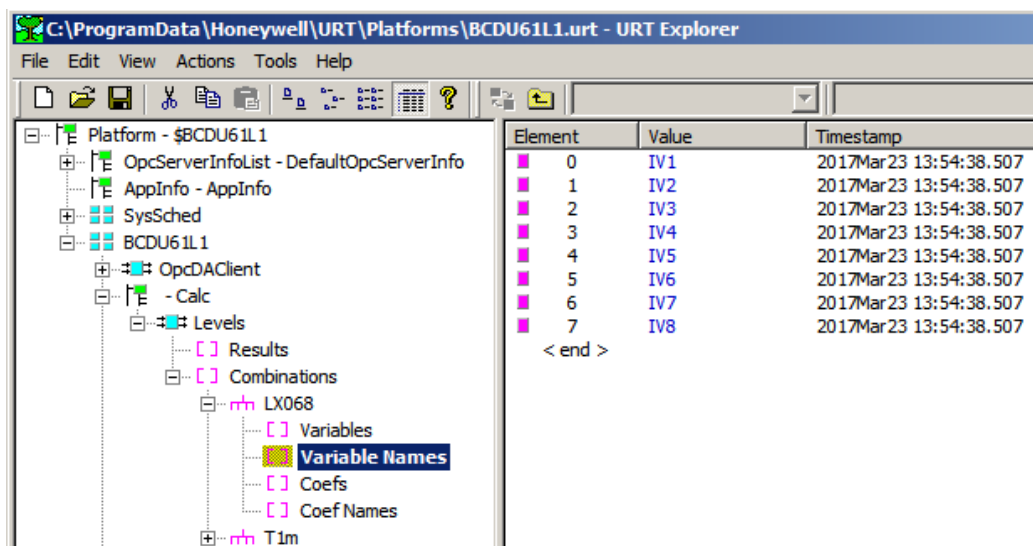


Fig. 52. Variable names.

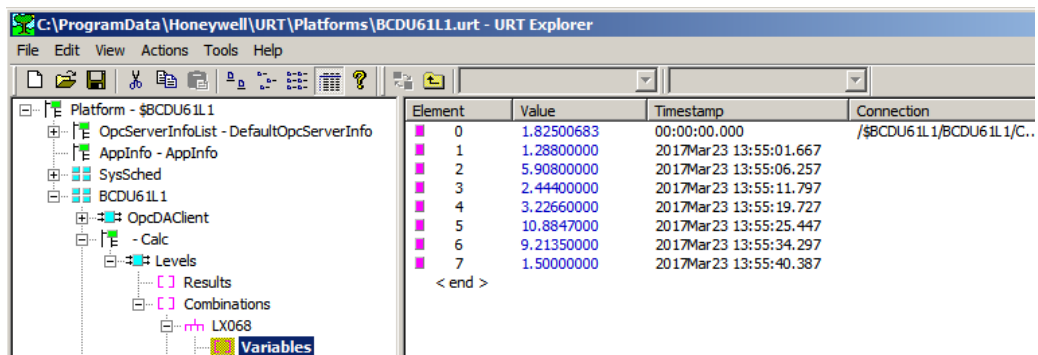


Fig. 53. Variable values.

After defining variable names, we define their values. Only the first variable IV1 (level in meters) is changing at a time and it is result from T1m block. We have to connect this variable with the corresponding result. Right click on **Variables** => **Properties** => **PerElem IN Con** and choose Type: URT and a Target is the location of T1m result shown in Fig. 54.

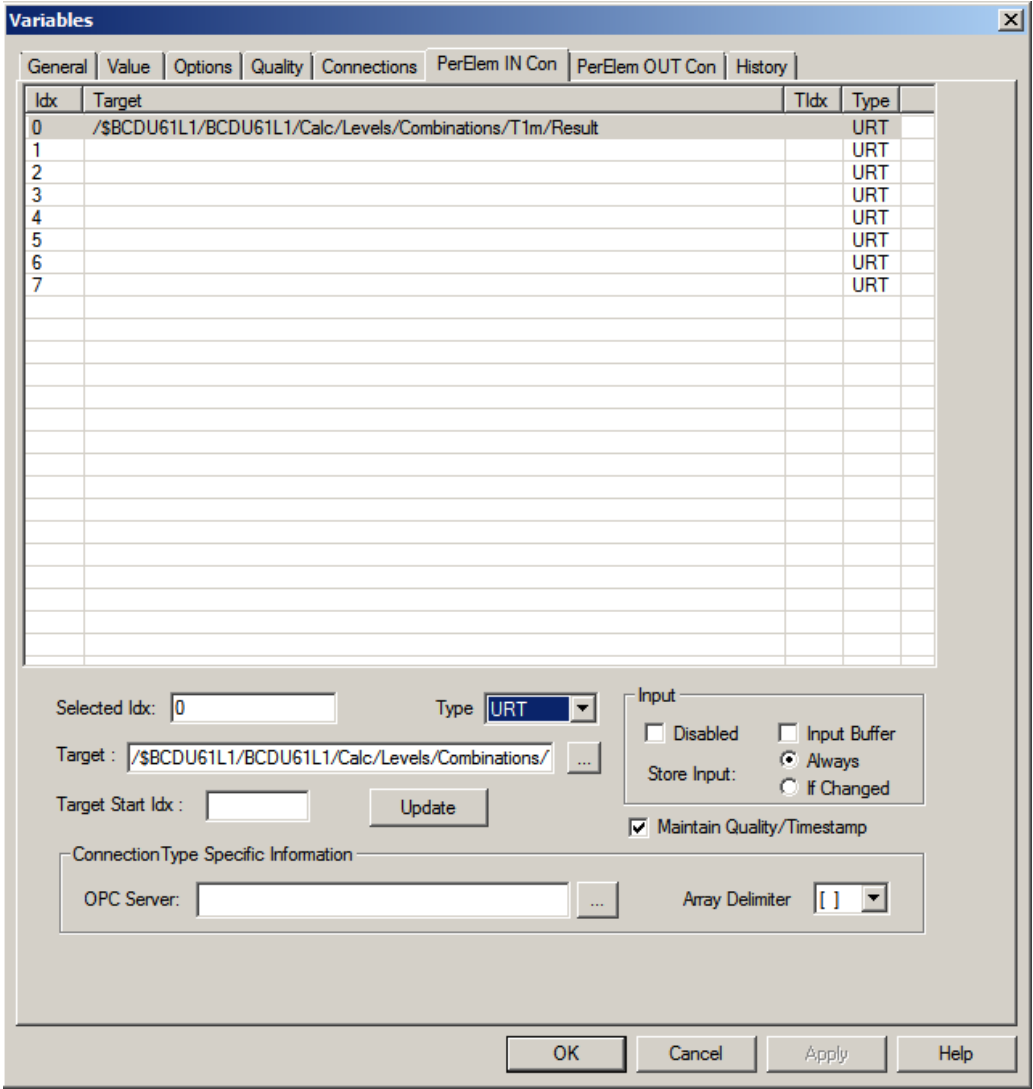
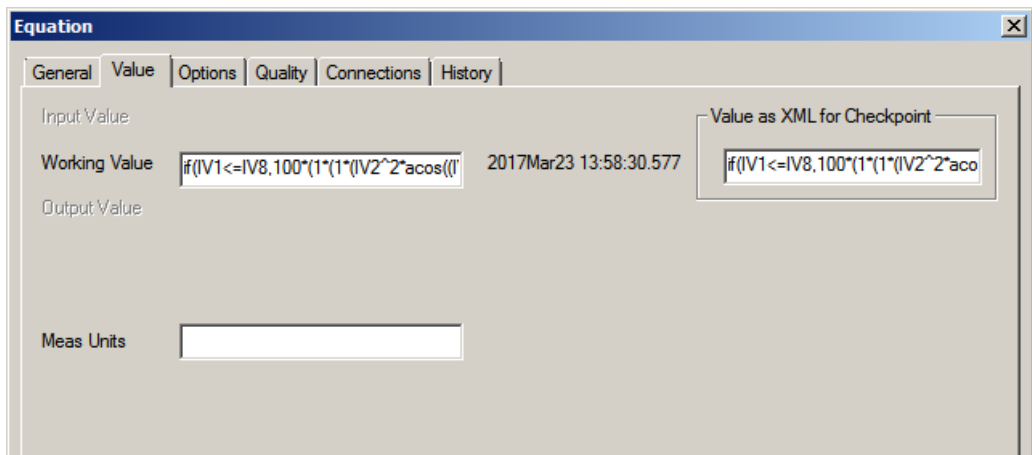


Fig. 54. Variable connection.

The last but very important step is to define an equation for calculation with decision rule described in 2.1. **Variables** => **Equation** => **Value** => **Working value**. In this box, we write all decision rules with two equations (Fig. 55).





**Fig. 55. Equation.**

The result from this block is the percentage of working volume in tank T1. The same approach is used for T3 level calculation.

## 8 Controller Configuration and Strategy

Part of the package that we used – Profit Suite is Profit Suite Operator Station (PSOS). This software is used for interaction between operator and APC. Here we set controller strategy by changing of optimization coefficients and limits. Our aim was to ensure, that T2 heavy naphtha flow fluctuation will be compensated by T1 heavy naphtha flow while levels will be inside limits. At first, we set some parameters for CVs as can be seen in following pictures.

CV #	CV Name	CV Description	Status	Value	SS Value	Low Limit	High Limit
1	LX068.PV	T1 level	GOOD-C	38.943	37.423	32.000	48.000
2	LX124.PV	T3 level	GOOD-C	37.757	37.216	10.000	70.000

**Fig. 56. CV Summary.**

As we mentioned before, instead of level we use a percentage of working volume for control. All values in Fig. 56 are a percentage of working volume. In this table, we set limits for levels. The wider range is allowed for CV1, the bigger capacity for flow compensation is possible. This setup is based on the range accepted by operators.

CV #	CV Name	CV Description	Linear Obj Coeff	Quadratic Obj Coeff	Desired CV Value	Delta Soft Low Limit	Delta Soft High Limit	Error Tolerance
1	LX068.PV	T1 level	0	1.0000	40.000	0	0	0
2	LX124.PV	T3 level	0	1.5000	38.231	0	0	100.00

**Fig. 57. CV Optimize.**

We use control to desired value for levels so we set quadratic objective coefficients, according to their priority and linear coefficients are equal to zero. The higher coefficient for CV2, the higher is speed of compensation. Coefficient 1.5 was set to get quite conservative but robust compensation for all types of fluctuations.

We set penalization for T3 outlet flow movement and levels control priorities. This configuration should ensure compensation of T2 heavy naphtha flow by T1 heavy naphtha flow. However the result was, that APC controlled both levels on the setpoint because of steady state optimization. That was against our idea that we need to fluctuate level in T1 in reverse with T2 level to minimize fluctuation in T3. The solution was to change T3 outlet flow to DV and thus removing one degree of freedom so both levels setpoints cannot be achieved and APC has to choose which level it will control with one MV (output flow from T1). T3 level is controlled by the ordinary cascade.

MVs were configured in following way:

MV #	MV Name	MV Description	Status	Value	Move	SS Value	Low Limit	High Limit	Mode	.
1	LC068.OP	T1 HN output flow	ON	80.898	-0.0726	80.729	35.000	85.000	RMPC	C

**Fig. 58. MV Summary.**

MV #	MV Name	MV Description	Linear Obj Coeff	Quadratic Obj Coeff	Desired MV Value	Delta Soft Low Limit	Delta Soft High Limit	MV Optimization Speed
1	LC068.OP	T1 HN output flow	0	0	0	0	0	NORMAL

**Fig. 59. MV Optimize.**

Limits were set according to process data and manual limits. In the previous setup MV2 weight was set, to penalize its movement. However, the strategy was changed and MV weight is not set.

DVs configuration is defined here:

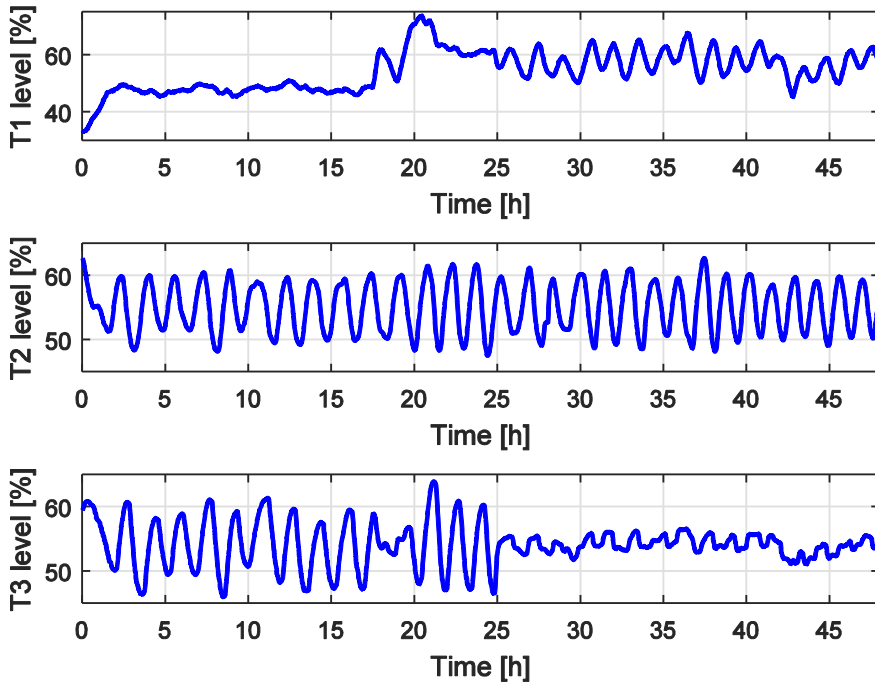
DV #	DV Name	DV Description	Status	Value
1	FC128.SP	T3 HN output flow	GOOD-C	61.191
2	FT118.PV	T2 HN output flow	GOOD-C	34.585

**Fig. 60. DV Summary.**

There are no limits to set, what comes from nature of DVs.

# Discussion

After few hours of running APC controller, we have results to show. In Fig. 61 it can be seen the behavior of levels in tanks before and after the APC was turned on. It was turned at time of 25 hours.



**Fig. 61. Levels in tanks.**

As you can see in Fig. 62 the APC try to fluctuate level in T1 in reverse with T2 level. This control causes less fluctuation in T3 level. In Fig. 63, Fig. 64, Fig. 65 there are all tanks – levels and output flows. T1 level oscillates between 45 and 65%. In order to set speed and capacity of T1 compensation CV1 quadratic coefficient and limits can be changed. The higher quadratic coefficient for CV1 means faster compensation. Wider limits for CV1 mean higher capacity for compensation. For the purpose of analysis, it would be good to show results for different configurations. However this would mean upset for normal operation, so we minimized the time for configuration and there are no relevant data to show. For this reason, we did not verify our assumptions about effects of changes CV1 parameters.

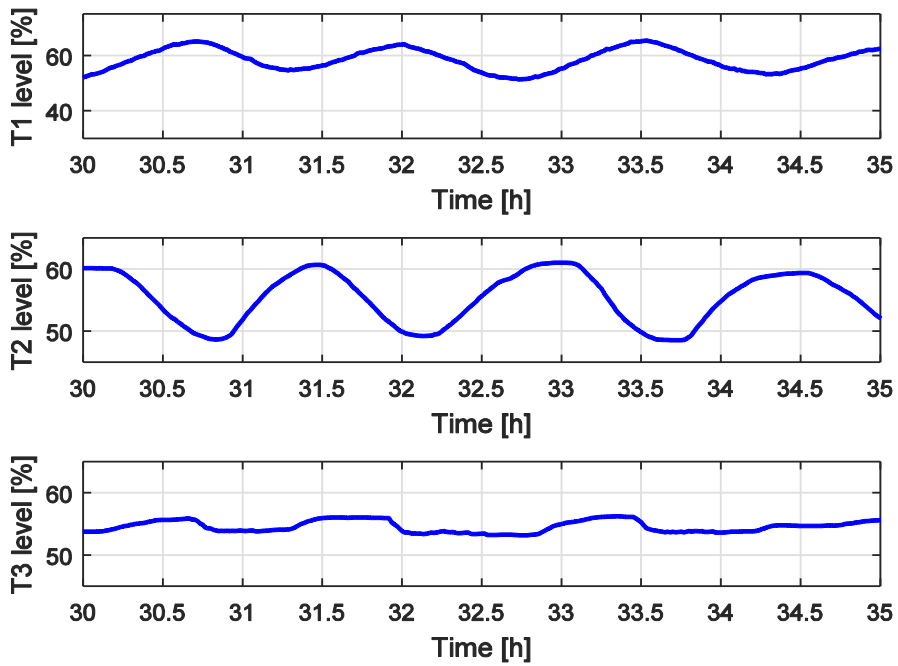


Fig. 62. Detail of control performance.

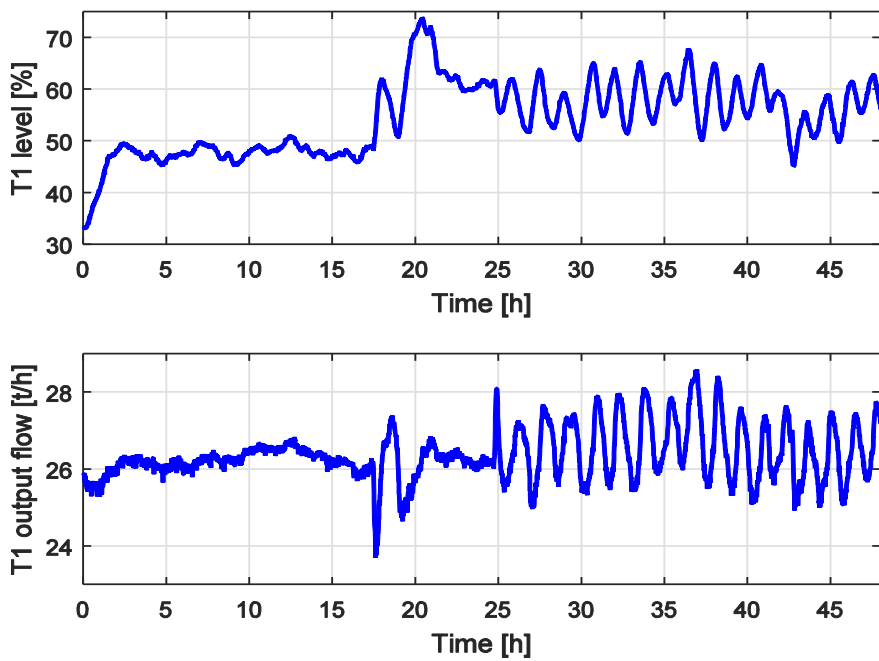


Fig. 63. T1 control.

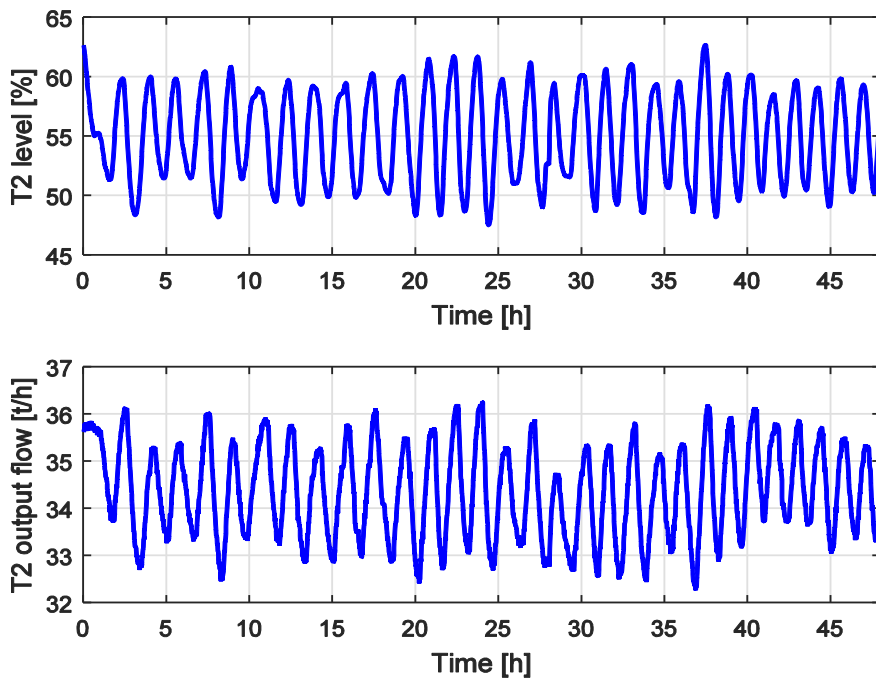


Fig. 64. T2 control.

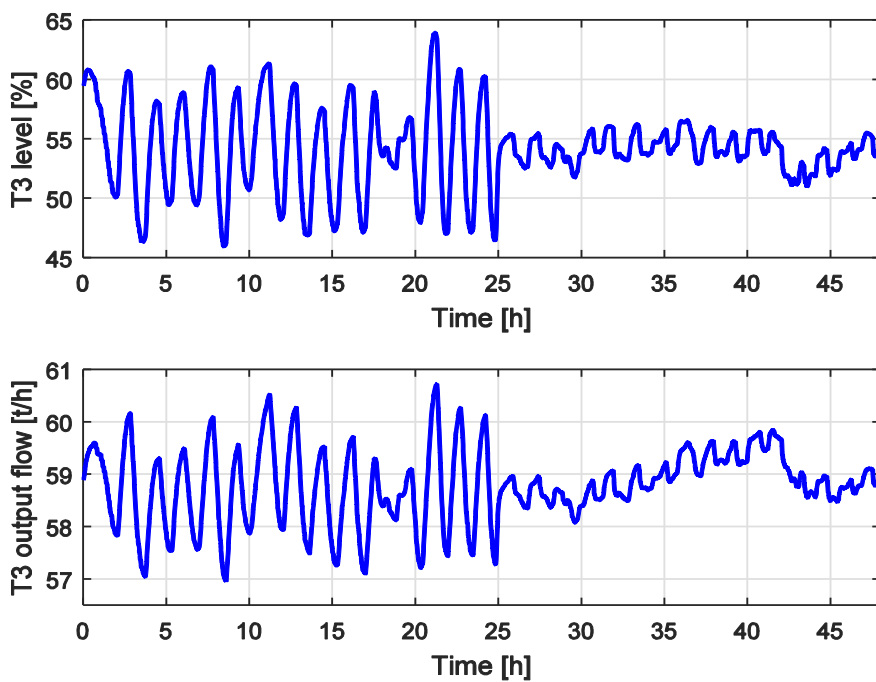
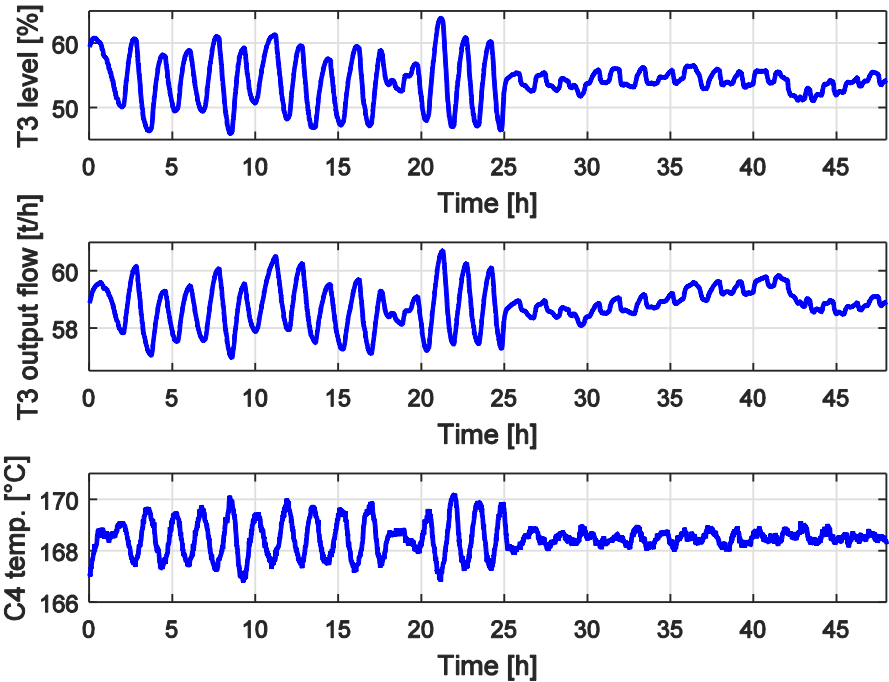


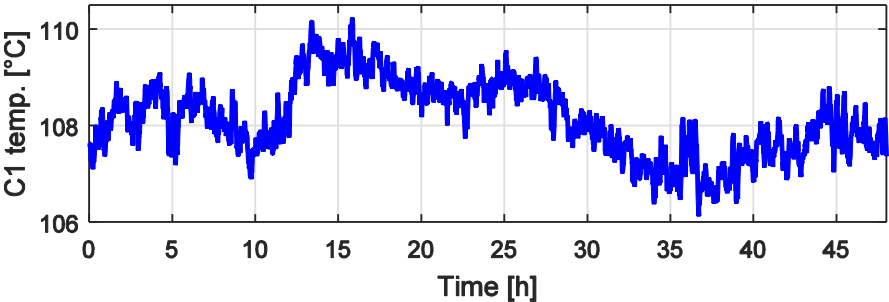
Fig. 65. T3 control.

Stabilisation of level in T3 implied stabilisation of output flow from T3, which is a feed flow for a C4 distillation column. The bottom temperature in the column also stabilised - Fig. 66. According to the conservative assumption, at least 60% temperature oscillation was removed.



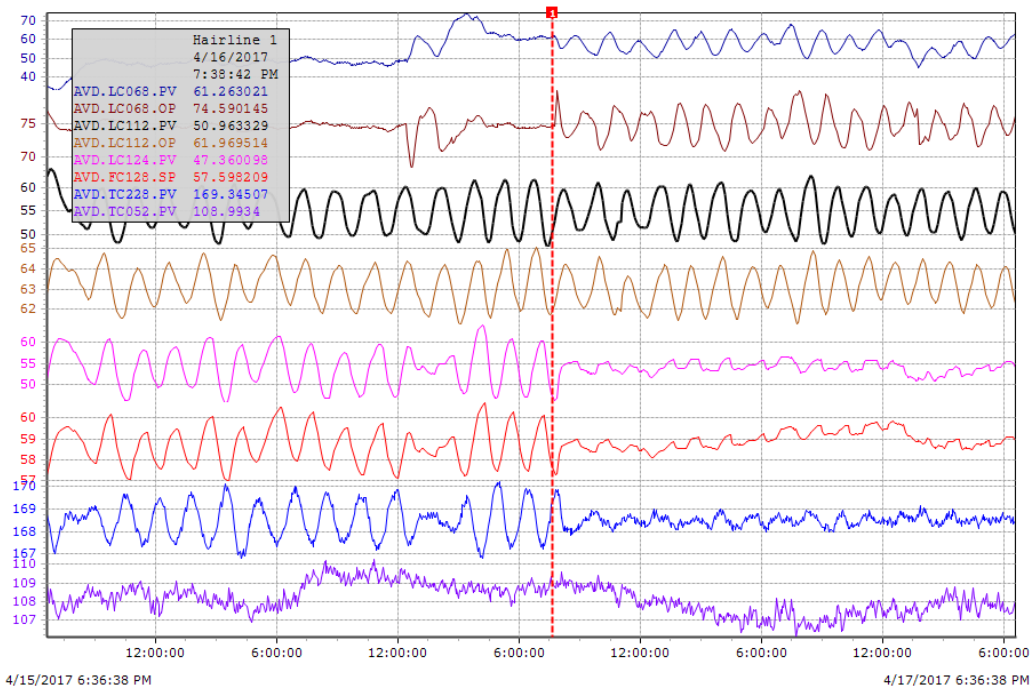
**Fig. 66. Final results.**

There was no significant change in processes upstream after turned on APC. This is presented in Fig. 67. As you can see there is no change caused by APC after 25th hour.



**Fig. 67. The temperature in C1 column.**

For illustration there is graph from Uniformance Process History Database (Honeywell) used in Slovnafit:



**Fig. 68. Uniformance PHD trend.**

**Table 9. The meaning of signals.**

Signal	Meaning	Unit
AVD.LC068.PV	Level in T1	%
AVD.LC068.OP	Output from controller to valve	%
AVD.LC112.PV	Level in T2	%
AVD.LC112.OP	Output from controller to valve	%
AVD.LC124.PV	Level in T3	%
AVD.FC128.SP	Output from controller to valve	t/h
AVD.TC228.PV	Temperature in C4 column	°C
AVD.TC052.PV	Temperature in C1 column	°C



# Conclusion

The aim of this work was to reduce the fluctuation of output flow from a system of three horizontal cylindrical tanks. This system of tanks is a part of distillation unit in Slovnaft called BCDU6. Fluctuation of output flow disturbs downstream operations.

Our first task was to derive a mathematical model of horizontal cylindrical tank. There was an interesting fact that inside each tank is a barrier used for dividing gasoline from water. In this first part, we used mainly MATLAB with Simulink. After deriving model, we validated it and continued with control design. We focused only on the last tank T3. At first, we tried to reproduce Slovnaft control system of level – cascade control. Slovnaft engineers tried to reduce non-linearity of the system using gain scheduling as we did. Next work was about developing our own control strategy to minimize fluctuation of output flow. We used a strategy called averaging level control. This strategy is based on an idea that we can use all possible volume of the tank to maintain output flow as steady as possible. Using this strategy we reduced fluctuation by half. The stronger improvement we achieved using tuned PI controllers with gain scheduling. The last strategy that we tried by simulations was a percentage of working volume control instead of level. Using this strategy makes our system linear which is better for control. This strategy was also the basis for APC control in the second part.

The second part was about APC controller design. We used Profit Suite from Honeywell which contains several types of software for design and application of APC controller. The first step was the identification of system. We had to classify variables and did some step tests. Step tests we did in MATLAB – Simulink and data were exported into Profit Design Studio. Here, we investigated how depended variables (MVs, DVs) affect independent variables (CVs). We constructed a model matrix which contains all sub – processes. In the next step, we built URT Platform which serves to implement APC applications. In URT Explorer we defined necessary calculations to convert level into a percentage of working volume. The last thing that we had to do was running a controller and define limits and some another parameter for variables. After running designed controller for few hours, significant progress has been made in output flow fluctuation. Reducing of output flow fluctuation reduced also fluctuation of temperature (at least 60%) in the next distillation column, which has a positive impact on operations conducted downstream. The controller was accepted by unit staff and they use it. Since the model is linear and it can be theoretically calculated, our solution is easily transferable to other tank systems.

# Resumé

Táto diplomová práca sa zaoberá návrhom a implementáciou riadenia sústavy troch zásobníkov kvapaliny v rafinérii Slovnaft a.s. Jedná sa o prevádzku BCDU6 na ktorej prebieha atmosférická a vákuová destilácia ropy za vzniku produktov. Spomínanú sústavu zásobníkov kvapaliny tvoria valcové zásobníky umiestnené horizontálne. Prvé dva (T1, T2) zachytávajú produkt z rektifikácie, ich výstupné prúdy sa spájajú a tvoria vstup do tretieho zásobníka (T3), ktorý zadržiava nástrek pre vákuovú kolónu C4. Výstupný prietok z T2 kolíše, a tento fakt spôsobuje kolísanie výšky hladiny v T3, výstupného prietoku z T3 a následne aj teploty v kolóne C4. To má za následok nepriaznivé podmienky pre riadenie a samotný chod kolóny. Cieľom tejto práce bolo stabilizovať kolísanie hladiny v T3 a tým aj kolísanie výstupného prietoku z T3.

Prvá časť práce bola zameraná na modelovanie systému, teoreticky návrh riadenia a overenie pomocou simulácii v programe MATLAB – Simulink. Prvým krokom bolo získanie matematického modelu horizontálneho valcového zásobníka kvapaliny. Všetky tri zásobníky mali rovnakú geometriu, líšili sa len veľkosťou. Zaujímavosťou pri týchto zásobníkoch bola zabudovaná prepážka vo vnútri zásobníka, ktorá slúžila na oddeľovanie zvyškovej vody v produkte rektifikácie. Odvodený matematický model bol nakoniec validovaný pomocou dát poskytnutých spoločnosťou Slovnaft a.s.

Pri teoretickom návrhu riadenia sme sa zamerali na zásobník T3. Slovnaft riadi hladinu v T3 pomocou kaskádovej regulácie PI regulátormi. Výška hladiny v zásobníku je nelineárny systém, čomu mal v riadení dopomôcť gain scheduling.

Po úspešnom odsimulovaní riadiaceho systému používaného rafinériou Slovnaft sme sa zamerali na vytvorenie vlastnej stratégie riadenia, ktorá by mala znížiť kolísanie výstupného prietoku z T3. Stratégia sa nazýva „Averaging level control“. Hlavnou myšlienkou tejto stratégie je využiť celý možný objem zásobníka bez dosiahnutia alarmov, kedy by sa mal výstupný prietok v značnej miere ustáliť. Touto metódou sa podarilo dosiahnuť menšie kolísanie výstupného prietoku približne o polovicu. Ďalšie zlepšenie bolo pozorovateľné ladením regulátorov a pridaním gain scheduling-u. Poslednou alternatívou, ktorú sme simulovali bola metóda, pri ktorej namiesto výšky hladiny v zásobníku riadime percento objemu zaplnenia zásobníka. Ako už bolo spomenuté, výška hladiny v našom zásobníku je nelineárny systém, čo neplatí o percente objemu. V tomto prípade bolo dôležité zásobník rozdeliť na niekoľko častí a vypočítať ich objem. Následne sme podľa rovníc v kapitole 2.1 z výšky hladiny vedeli vypočítať percento

zaplnenia objemu zásobníka. Aj táto metóda priniesla úspech v podobe zmenšeného kolísania výstupného prietoku zo zásobníka T3.

Druhá časť bola venovaná návrhu a implementácii APC regulátora. Používali sme softwarový balík Profit Suite od Honeywellu, ktorý mal mnoho súčastí. Najprv bolo potrebné rozdeliť premenné na riadené (CV): výšky hladín v zásobníkoch T1 a T3, riadiace (MV): výstupné prietoky zo zásobníkov T1 a T3 a poruchové (DV): výstupný prietok z T2. Výstupný prietok z T2 bol zaradený medzi poruchové veličiny kvôli zadrhávaniu ventilu, kedy nie je možné ovládať ho pre potreby riadenia. V ďalšom kroku sme pristúpili k identifikácii systému. Na začiatku bolo potrebné vykonať skokové zmeny alebo tzv. steptesty, kde sme sa snažili zistiť vplyv riadiacich a poruchových premenných (MV, DV) na riadené premenné (CV). Tieto steptesty sme robili simulačne v programe MATLAB - Simulink a vygenerované dáta sme spracovali v programe Profit Design Studio. Výsledkom bola modelová matica zložená z jednotlivých čiastkových procesov. V ďalšom kroku sme vytvorili URT Platformu, ktorá slúži na implementáciu APC aplikácii. Pri vytváraní APC regulátora sme sa rozhodli použiť metódu riadenia percenta objemu zaplnenia spomínanú v prvej časti tejto práce. Výpočty potrebné na prepočet výšky hladiny na percento zaplnenia objemu zásobníka sme definovali v programe URT Explorer.

Posledným krokom bola implementácia vytvoreného APC regulátora. APC regulátor bol nahraný do systému a boli vykonané určité nastavenia v rozhraní pre operátorov – Profit Suite Operator Station. Po spustení navrhnutého regulátora a následnej niekoľko hodinovej prevádzke bolo vidieť, že výška hladiny v zásobníku T1 ako aj jeho výstupný prietok začali výrazne kolísať. Pomocou tohto rozkolísania hladiny v T1 sa znížilo kolísanie hladiny v T3, jeho výstupný prietok a následkom toho aj kolísanie teploty v kolóne C4 o najmenej 60%. Regulátor bol akceptovaný prevádzkou a používa sa. Pretože model je lineárny a je ho možné vypočítať teoretický, je naše riešenie ľahko prenosné na iné sústavy zásobníkov.

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