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Summary

In this project, a steady-state HYSYS model of the precooling section of a C3-MR process has been rebuilt from an earlier model and adapted to dynamic simulation. Dynamic simulations have been run using two different control setups. Both control setups were tested on two different simulation scenarios – a disturbance in natural gas feed flow rate, and a disturbance in feed temperature. Different aspects of the simulation software and control setup have been discussed, and some problems with the dynamic simulation have been pointed out. Among the conclusions made are:

HYSYS's basic heat exchanger model does not give very realistic results in dynamic simulation.

The controller settings suggested by the HYSYS Dynamic Guide are somewhat aggressive for this model and tend to give oscillation. Cascade control on heat exchangers reduces oscillation.

Contents

1. Introduction

This project deals with the development of a dynamic process model in Aspen HYSYS. The basis is a steady state model of the APCI C3-MR process for production of liquefied natural gas (LNG). The model was developed during a summer internship in 2006 at Norsk Hydro's Oil & Energy Research Centre in Porsgrunn.

The process that has been modelled is the pre cooling part of the C3-MR process for liquefaction of natural gas. In this part of the process, natural gas and the mixed refrigerant (MR) are both cooled to approximately - 40°C in a propane cycle.

The main aims for the project are the following:

- To get a better understanding of dynamic process simulation software and of the task of building a dynamic model from a stationary model
- To study how well Aspen HYSYS is suited for simulation of this type of process and to explore the pros and cons of the program as a dynamic process simulator
- To explore different process control setups and evaluate how they work with respect to stabilizing of the process, thereby getting a better understanding of process control

This is to be accomplished through building a dynamic model of the pre cooling part of the C3-MR process including necessary control loops, and running simulations in order to see how the model handles process upsets.

2. Background

2.1 Steady-state and dynamic process simulation

Modern computer technology has allowed engineers and researchers to study the behaviour of process plants without having to manipulate actual plants or building expensive pilot plants. Different computer software has been developed for this purpose, like Aspen, HYSYS, gPROMS, ChemCAD and more. Mathematics software like MATLAB can also be used for simulation purposes, but are typically not used for simulation of large processes. The programs are different in several ways – user interface, what kind of user input is required, and what kinds of equation systems the program can solve. Some programs have predefined blocks resembling process units; other programs let the user define the blocks by giving equations and parameters. Combinations also exist, where one can choose between using the program's prebuilt units and defining the equations by oneself.

The backbone of any simulation software is the solver algorithm it uses. To the computer, a simulation case is nothing more than a set of equations that have to be solved. For a steady-state simulation, the set consist of only algebraic equations. These can be solved sequentially, simultaneously or as a combination of these. For a dynamic simulation, the equations include both differential and algebraic equations. The differential equations can be ordinary (for lumped units) or partial (for distributed units like tubular reactors). The methods for solving ODEs and PDEs are different from program to program, some simulation programs can only handle ODEs (lumped systems) while others are constructed specially for good performance on PDE systems.

The simulation program used for this project, Aspen HYSYS, is a block-oriented program where the user builds the process model from predefined blocks and supplies necessary parameters. HYSYS runs both steady-state and dynamic simulations and has built-in tools for dynamic initialization and equipment sizing. HYSYS does not handle distributed systems, but instead divides units into zones where each zone is calculated as a lumped unit. This applies to heat exchangers and tubular reactors (separators and tank reactors are treated as a single hold-up). HYSYS uses the implicit Euler method for solving the differential equations; the step size used by the solver is set by the user.

Dynamic simulation of process plants is a complex task. A dynamic model must include all necessary unit operations (also some that may be neglected in steady-state simulation), all physical units have to be sized in a realistic way, consistent initial conditions need to be provided and the correct specifications have to be set on boundary streams. In addition, one needs a basic control structure to stabilize the model. Otherwise, small errors in the solver algorithm could easily build up and make the simulation drift away from the desired process conditions.

If a steady-state model is available, as in this project, one will typically have a consistent set of initial conditions. Adding control valves or buffer tanks to the system will usually not lead to large changes in process conditions such as flows and temperatures. This means that adapting a stationary model to dynamic simulation does not necessarily change the process conditions much.

2.2 Brief description of the C3-MR process for LNG production

General about LNG

When natural gas needs to be transported over long distances (for example, trans-Atlantic) it is typically freighted on ships. In order to transport large quantities of gas, it is cooled to a temperature at which it is completely liquid. This typically means a temperature of approximately -162°C at slightly above atmospheric pressure. Typically the liquefaction is carried out using a cascade of cooling loops or in the simplest case, like in the PRICO process (for a paper where this process is used as an example process, see [1]) only one cooling loop where the refrigerant is compressed to high pressure, cooled by water, expanded through a valve and sent to the main exchanger where it cools the natural gas to the desired temperature. Other processes include the Mixed Fluid Cascade Process (MFCP, Linde), Dual Mixed Process (DMP, Shell/Sakhalin), and the C3-MR process (APCI).

The C3-MR process

In this process, the natural gas is pre-cooled with propane before entering the main exchanger. Propane is also used to cool the mixed refrigerant (MR) for the main exchanger. This is carried out in the following way: Propane at high pressure is condensed with water, and then sent to the vaporizers. Before each vaporizer, there is a choke valve where propane pressure is reduced in order to obtain the desired boiling temperature. There are typically three pressure levels (this may vary) and six propane vaporizers (for each pressure level, there is one vaporizer to cool natural gas and one to cool the MR). The propane vapour from each vaporizer goes to the corresponding stage of the propane compressor. The compressor has one stage for each pressure level in the process.

The natural gas is now fed to a distillation column where heavy components are removed, to avoid out-freezing of heavy hydrocarbons in the main liquefaction exchanger (from now on referred to as the MCHE, main cryogenic heat exchanger). The MCHE is a spiral-wound exchanger where the “hot” fluids flow upwards through bundles of thin tubes, and the cold fluid is sprayed over the tubes from the top of the exchanger.

The stream from the top stage in the column is condensed in the bottom bundle of the MCHE, and led to a reflux drum. The vapour is then fed to the middle bundle and is cooled as it flows to the top of the exchanger. It is flashed across a valve to a pressure slightly above atmospheric, and led to a flash tank. The liquid from the flash tank is the LNG product.

The mixed refrigerant (MR from now on) consists of methane, ethane, propane and nitrogen, the exact composition is adjusted after the composition of the natural gas. After the last propane vaporizer, it enters a separator and is split in a liquid and a vapour phase. Some of the vapour phase is used to heat the gas from the LNG flash drum, and it is then expanded and injected at the top of the MCHE. The rest of the vapour and the liquid phase are fed to the bottom of the MCHE and cooled. The liquid is taken out on the top of the middle bundle and expanded before injected on the shell side of the same bundle. The vapour goes all the way to the top where it is flashed across a valve to obtain the desired temperature, and then injected back to the top of the MCHE.

The fluid (the MR) on the shell side of the MCHE should be completely vaporized when it leaves the exchanger. Then it is compressed over three compressor stages – between the stages it is cooled with water. After the last compressor and cooler it is fed to the first propane vaporizer again. The C3-MR process is described and discussed in more detail in [2] and [7]; the illustrations below also come from these papers.

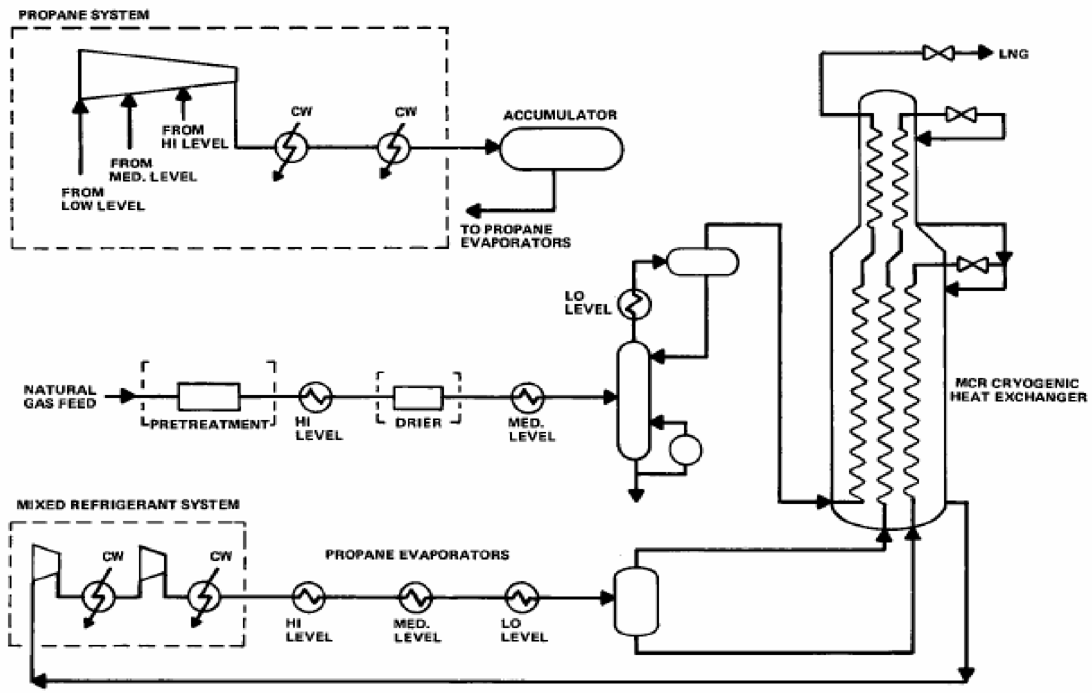

Figure 2.1: Schematic representation of a C3-MR process (from [7])

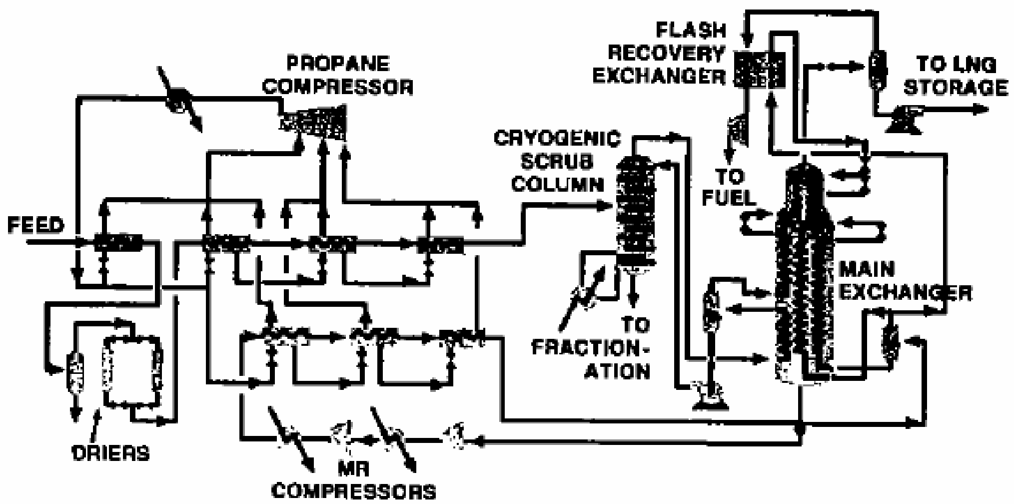

Figure 2.2 Schematic representation of a C3-MR process (from [2])

3. The HYSYS model and how it was built

3.1 Brief description of the HYSYS model

The model built in this project is built in Aspen HYSYS 2004.2. Standard components are used in the component list (no pseudo-components) and the Peng-Robinson equation of state is used as thermodynamic package. The compositions of natural gas and mixed refrigerant are given in table 3.1.

Table 3.1 Mole fractions of components in the process streams

Component	Natural gas	Mixed refrigerant	Propane
Methane	0.70	0.45	0.0
Ethane	0.18	0.45	0.0
Propane	0.08	0.08	1.0
Nitrogen	0.04	0.02	0.0

The model is limited to the propane loop – this means the main cryogenic exchanger, the column for removal of heavy hydrocarbons and the units further downstream are left out. The MR “feed” is the point where it leaves the MCHE, and the natural gas only goes through the three propane/NG heat exchangers (named E101, E102 and E103).

When the natural gas enters the flow sheet, it goes through a flow control valve (VLV-100) before passing through E101 (vaporizes HP propane), E102 (MP propane) and E103 (LP propane).

The mixed refrigerant enters as a cold gas at low pressure and is compressed through three compressors – after each compressor there is a cooler where the MR is cooled with water. The units are, in sequence, named LP MR Compressor 1, LP MR Intercooler, and LP MR Compressor 2, LP MR After cooler, HP MR Compressor and HP MR After cooler. After the HP MR After cooler, the MR is cooled further with propane in the propane vaporizers E104, E105 and E106. The flow to the first compressor is controlled by a valve (VLV-112) on the feed stream.

For a description of the propane loop, start at the outlet of the HP compressor (K-101). The propane is condensed with water in a heat exchanger (the propane cooler/condenser), and leaves as saturated liquid (the pressure at the compressor outlet is set to a value where this boiling point is 10°C, assumed to be a realistic temperature for cooling with water). The water is supplied through VLV-101.The liquid propane enters a buffer tank (V-100), and is then fed to the propane vaporizers. After passing through the level control valve the propane is split to the six propane vaporizers E101-106. Before every propane vaporizer, there is a valve where propane is expanded to reach desired temperature and pressure. The valves are labelled VLV-103 to VLV-108.

The overhead streams from the propane vaporizers go to the compressors; from E103 and E106 to the LP compressor K-103, from E102 and E105 to the MP compressor K-102 (together with the outlet stream from the LP compressor), and from E101 and E104 to the HP compressor K-101 together with the outlet stream from the MP compressor.

In the real process (again, see [2]), the propane compressor is one single unit with three inlets, but as HYSYS does not allow multiple stage compressors, it has to be split into three separate compressors with stream mixers in between. It is possible to “link” compressors in HYSYS; that is, defining them to run at the same number of revolutions per minute (rpm) or with a constant rpm ratio. In real processes, one will often see that several compressors are mounted on the same shaft, for instance if they are run by the same engine. This scenario is allowed for in simulation by linking the compressors. However, this requires that one supplies information on how the efficiency of each compressor depends on speed. Therefore, this option has not been used in this model.

When moving from steady state to dynamic simulation, one needs a basic control system to stabilize the process. The control structures used are described in a later chapter (chapter 4), together with a more general discussion of control of the loop. Figure 3.1 shows the HYSYS flow sheet. Blue lines indicate streams, green lines indicate control and measurement signals. The controllers shown here are the ones used in “Control scheme 1” (see chapter 4.3). Natural gas enters through VLV-100 and mixed refrigerant through VLV-112.

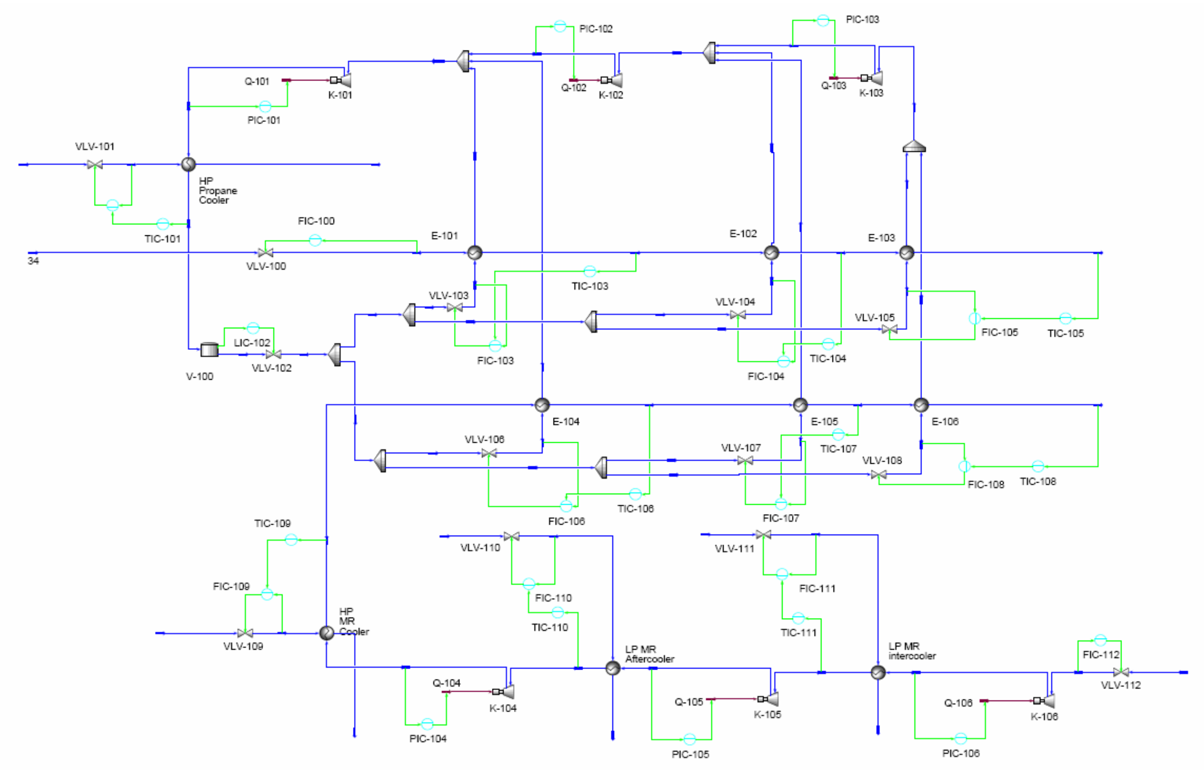

Figure 3.1: Flow sheet of the process as modelled in HYSYS, with controllers included

3.2 Procedure for building the model

General

The starting point for the project was a stationary model built earlier. This model was built in HYSYS 2004, not the 2004.2 version. At the time when the work with the dynamic modelling started, only the HYSYS 3.2 version was available at campus. Therefore, the stationary model was rebuilt in this version as old versions of HYSYS can not run cases created in newer versions. When rebuilding the steady state model, some simplifications were done – to make the conversion to dynamic simulation easier. For example, in the original model, the propane vaporizers were only partial vaporizers, where liquid and vapour were separated in a tank after each heat exchanger, and were all the propane passed through the high-pressure vaporizers. In the new model, all vaporizers were total and the separator tanks could be omitted.

The main difference between dynamic and steady-state simulation is the need for dimensions of vessels, as well as the need for control operations (valves and controllers). In addition, all the pressures and flows are calculated simultaneously. Typically, pressure drops over vessels are either given or calculated from an equation containing parameters calculated by HYSYS or supplied by the user. Flows are determined from pressure differences, the pressures at feed and exit streams are typically specified by the user. This is the most realistic approach to flow in a real process, as flows are in fact caused by pressure differences. If there is no pressure gradient in a pipeline, there will not be any flow.

The valves that needed to be added for dynamic simulation were the inlet valves on natural gas (NG) and mixed refrigerant (MR), as well as a level control valve on the outlet of the buffer tank. Heat exchangers do not require valves between them, as each exchanger supplies a pressure-drop calculation itself. The choke valves on the propane streams to the vaporizers were left unchanged from the stationary model.

When switching a simulation case from steady-state to dynamic, HYSYS automatically suggests possible specification changes that are required or recommended for dynamic simulation. These are typically removing internal flow and pressure specifications, removing fixed specifications of pressure drops and replacing these with flow vs. pressure drop relations, as for example

$$f = \sqrt{\rho} \cdot k\sqrt{\Delta P} \qquad (2.1)$$

If some units are not sized, the dynamic assistant tells the user to do so, and if some unit in the process is over specified (t. e. a heat exchanger with all four temperatures and both flows given) this is pointed out for the user. If specifications are in conflict with each other, such as a pressure drop and a flow in opposite directions, this is suggested changed (for some reason, the steady-state solver accepts this. The dynamic does not.)

Sizing of units

HYSYS has the capability of sizing process equipment automatically when switching from steady-state to dynamic simulation. When using the Dynamic Assistant straight ahead, this is performed (if the units are not sized already). For valves, the automatic sizing requires that pressure drop, flow and valve opening are known. Typically one has defined a pressure drop and a flow from the steady-state model. Then one must decide what valve opening should correspond to these values. In this model, this value was set to 50% for all valves. When flow and pressure drop are known for the given valve opening, HYSYS can calculate the necessary constants for the valve, C_V and C_g. The default method used by HYSYS is the Universal Gas Sizing method.

For compressors, one does not need to supply geometrical sizing data, because the hold-up in the compressor is neglected in the calculations. A maximum compressor power is needed, though. In this model, all compressors have a maximum energy input such that the nominal value equals 90 % of the maximum available. This maximum input is defined through the "control valve" option belonging to the energy stream connected to the compressor (see figure 3.2). However, if a compressor is defined as variable-speed, one must supply performance curves for the compressor. If a surge controller is to be modelled, one must also supply a surge curve.

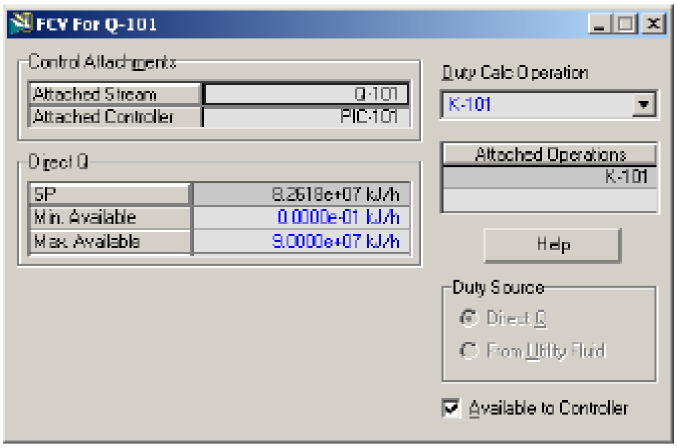

Figure 3.2: Specifying a maximum energy flow to a compressor

As pipe sections are not included in this model (this is mainly for simplicity and because adding pipe sections would not add realism unless realistic pipe dimensions were known) the only remaining units for sizing were the heat exchangers.

The sizing of heat exchangers was initially done by assuming an average heat transfer coefficient $U \approx 1{,}5$ kW/m^{2}K based on heat transfer coefficients from the model made at Norsk Hydro in the summer of 2006 using TASC. The heat transfer area necessary would then be found from the temperatures and duty from the stationary model after the standard heat transfer equation (2.2):

$$Q = U \cdot A \cdot \Delta T_{lm} \cdot f_t \tag{2.2}$$

By assuming that tube diameter (inner and outer) and length were the same as in the original model, the necessary number N of tubes could be calculated. The total inner volume of the tubes then was easily calculated from $V = \pi \cdot r \cdot l \cdot N$ with inner tube radius *r* and tube length *l*. As suggested on page 16 of [3], the free shell volume was assumed to be equal to the total inner volume of the tubes, and the necessary shell diameter was calculated. The thermal conductivity was set to be equal to that used in the steady-state model (51.92 W/m K) and standard values were used for baffle spacing and baffle cut (the fraction of the cross-sectional area covered by baffles). The exchanger layout was set as A-type front end heads, E-type shells and U-type rear end heads.

This sizing procedure used with the detailed rating model gave far too short residence time for both shell and tubes – in an order of magnitude of fractions of seconds, thus giving a model that gave very quick failures in the pressure-flow solver. Therefore, it was decided to use the basic rating model instead, and adjusting all volumes (shell and tubes) for each exchanger to give a residence time of 1 minute (for pure gas flow) or 5 minutes (for liquid or two-phase flow). To obtain these volumes while keeping the heat transfer area and sticking to the detailed model, the tube diameters would have to be increased significantly. This approach would lead to a loss of realism, just like the switch to the basic model does.

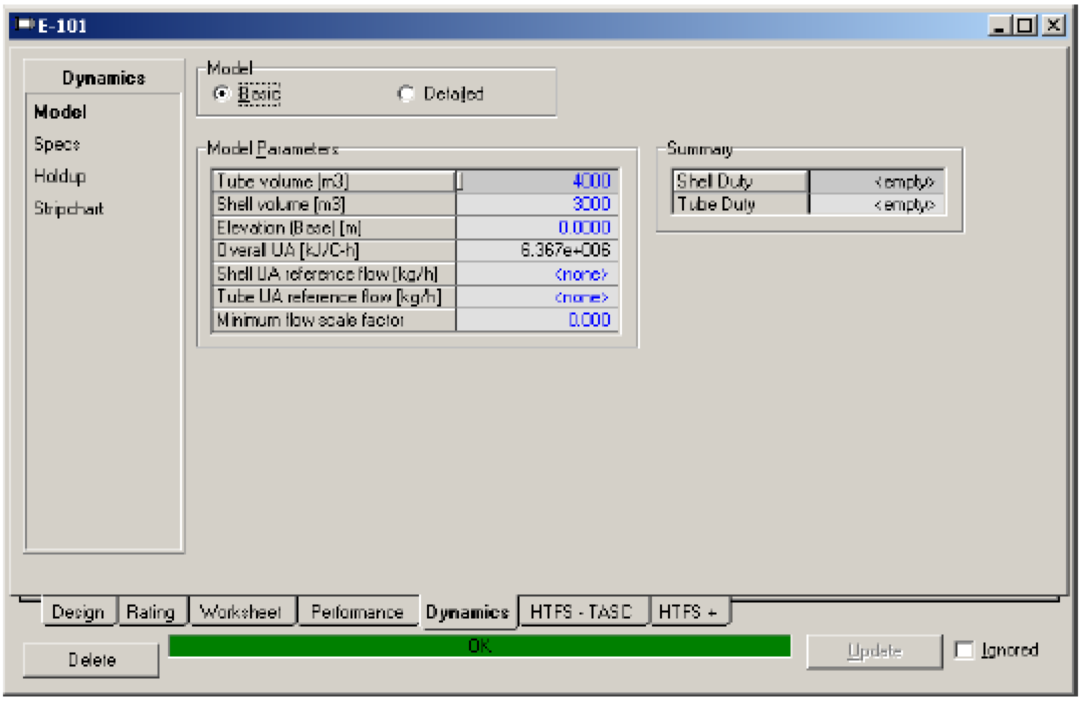

Figure 3.3: Specifying volumes in the basic heat exchanger model

Whenever the volumetric flow rate changed throughout a unit, the *larger* value was used. This was supposed sufficient to make the model stable. The problem with simulation of large heat exchangers in HYSYS is discussed in section 6.1. Dimensions of the buffer tank and heat exchangers are shown in table 3.2a; pressure drop and CV for valves in table 3.2b. The rest of the steady-state stream data (these are also used as initial values for the dynamic simulations) are given in Appendix A. Notice that for the buffer tank V-100 a volume of 1000 m^3 was assumed initially – it should probably have been larger, see section 6.3.

Table 3.2a: Volumes of heat exchangers (tubes and shells) and for V-100

Vessel (exchanger/tank)	Shell volume (m3)	Total tube volume (m3)
E-101	1333	408
E-102	4521	367
E-103	11550	302
E-104	4525	766
E-105	23030	631
E-106	30480	395
V-100	1000	

Table 3.2b: Pressure drops across valves at 50% open, values for C_V

Valve	Pressure drop at 50% open (kPa)	Valve constant $(C_V)/10^4$ USGPM
VLV-100	10.00	4.148
VLV-101	10.00	44.32
VLV-102	10.00	3.972
VLV-103	147.8	0.04649
VLV-104	372.6	0.05391
VLV-105	499.2	0.06004
VLV-106	147.8	0.1575
VLV-107	372.6	0.2747
VLV-108	499.2	0.1584
VLV-109	25.00	3.665
VLV-110	25.00	1.496
VLV-111	25.00	1.705
VLV-112	9.990	28.37

In addition to vessel volumes, pressure drops had to be specified for the steady state model. For the exchangers using water as coolant, pressure drops were specified as 5.00 kPa on the shell side and 10.00 kPa on the tube side. For the propane vaporizers, these values were set to 50.00 kPa (tube side) and 10.00 kPa (shell side). These values are in the same order of magnitude as those obtained using TASC. When changing the mode from steady state to dynamic, these specified values were replaced by k-values calculated from a relation like equation(2.1). When such a value is known, the pressure-flow solver can establish a relation between pressure drop and flow through the exchanger.

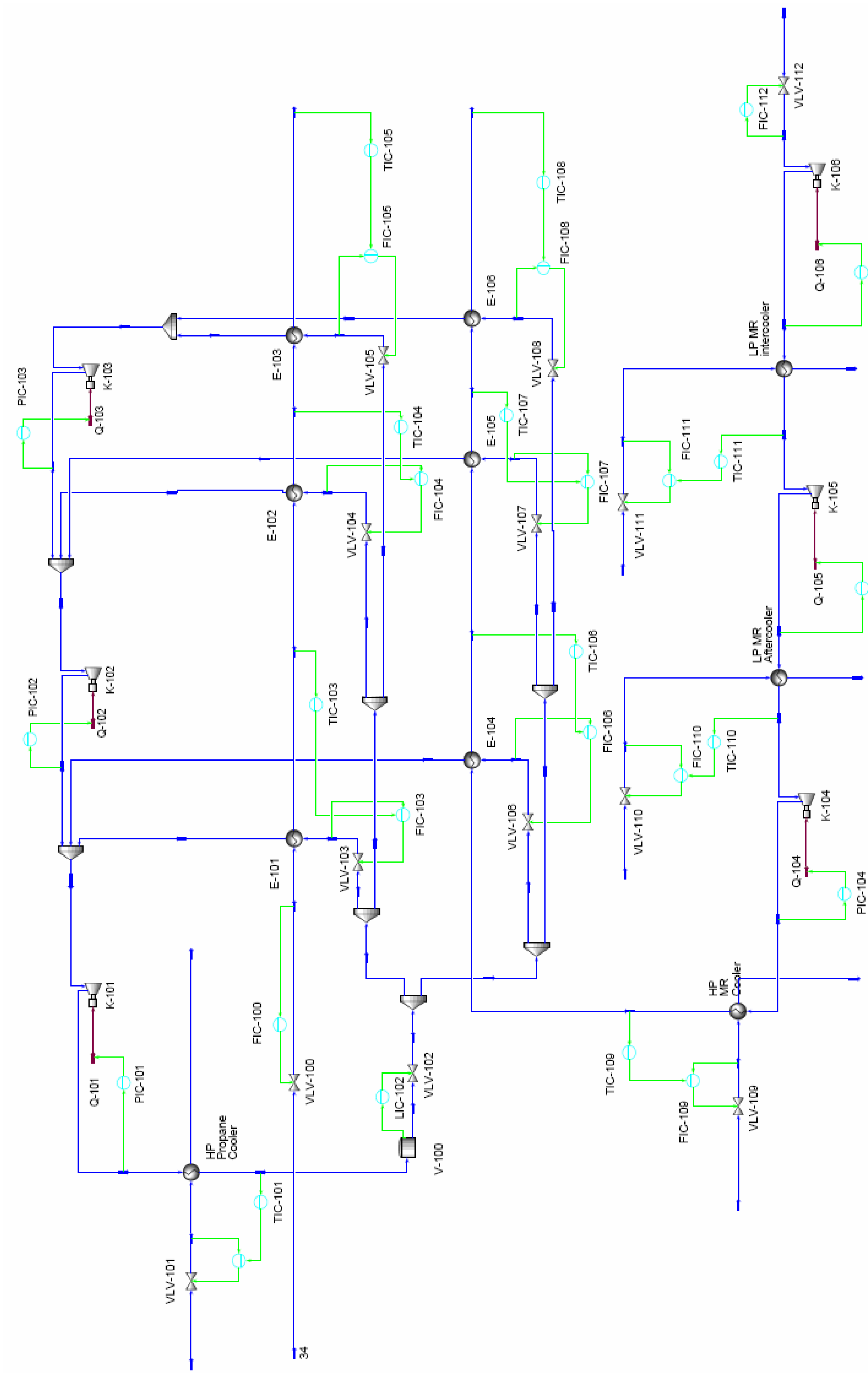

Figure 3.2: Larger view of HYSYS flow sheet with controllers included

4. Setting up the control structure of the model

4.1 Control – general

For any process plant, one needs to have at least a basic control system to keep the plant stable. The same goes for a model of the plant. A real process system will always be affected by the surroundings and by other systems (utility systems, upstream processes and so on) and would drift away from the desired operating conditions without control. For a process simulator, the control system is necessary to assure stability in simulation, because small deviations (calculation errors) otherwise could make the model drift away from the desired conditions. Dynamic simulations can also be used to test the behaviour of different control structures and settings. This allows exploring different aspects of process control. Stability, robustness and precision of different control schemes and controller settings can be studied.

When setting up a control scheme, one must consider which outputs are most important to maintain stable, which inputs are available for control and which inputs that are best suited for control. There may be different criteria that contribute to the decision; product quality requirements, limitations due to safety, power consumption, equipment capacity. Typically one will have some restrictions on inputs. These could be maximum available flows, pressures or compressor outputs. For the outputs one will typically have specifications required for downstream operations to function properly, and requirements to product quality. From an optimization study, one can extract the input and output values that give the best overall result. The goal for process control then must be to keep the process variables at these optimal values. To do this, both the optimal values and input constraints must be clearly defined. Of particular interest are the variables that can keep the process at optimum (or, at least, within an acceptable range) when controlled at a constant set point. The point is that when choosing these variables correctly, one should not need to do a new optimization when a disturbance occurs. This kind of control configuration is called *self-optimizing control*. For more on self-optimizing control, see Skogestad and Postlethwaite [4].

When simulating a process for learning purposes with focus on simulation, the desired outputs and inputs do not need to be optimal, the principles for control will still be the same. Find inputs that, when manipulated, are well suited to the purpose of keeping the outputs at the desired values. Optimization of the process and selection of self-optimizing variables is beyond the scope of this project.

4.2 Different types of process control – a brief summary

Feedback control

This is the simplest type of control, and the most widely used in the lowest control layer in a process plant. In feedback control, one measures the output, compares it to the desired value (the set point), and calculates the input needed to correct any deviation. This is usually a simple task, the controller settings can be found through step testing of the process, and it automatically corrects for any deviation in the output. The problem is that it is always acting *after* a disturbance has forced the output away from the desired value. This means that a feedback loop will always have a time lag.

Feed forward control

In feed forward control, one measures the disturbances rather than the outputs. This means that the controller takes pre-emptive action, to *prevent* that the output drifts away from the set point. In order to use feed forward control, one will need to have a precise process model, so it is more expensive in terms of work needed. In addition, a feed forward loop does not correct a deviation in the output automatically as it has no measurement of the output. Any small error in the feed forward controller will make the output drift off the set point. But, if done properly, feed forward control will be able to keep the outputs very close to the desired values all the time, reducing the chance of producing off-spec products or running the process at non-optimal options.

Cascade control

Because of the problem mentioned above in feed forward control, one will typically combine a feed forward loop with a feed back loop. The feedback controller measures the output and gives a correction to the feed forward controller, typically as a new set point for the variable manipulated by the feed forward controller. This kind of control, where one controller supplies the set point to another controller, is called *cascade control*. The "slave" controller does not have to be feed forward. A widespread use of cascade control is in flow control, where a valve is supplied with a desired flow from another controller, and the valve uses a flow measurement in its outlet stream to manipulate the valve opening. When this is done instead of letting the first controller manipulate the valve opening directly, one removes the uncertainty that is caused by irregular behaviour in the valve. In this scheme, the "slave" is a feedback controller.

The PID Controller

The most widely used control operation in process industry is the PID controller. Here, the input signal is obtained by adding a proportional (P), an integral (I) and a differential (D) term, and multiplying the error by the sum of these. A common way to describe a PID controller is (this is also how it is written in the HYSYS dynamics guide)

$$u(t) = K_C \cdot e(t) + \frac{K_C}{\tau_I}\int_0^t e(t)dt + K_C \cdot \tau_D \cdot \frac{de(t)}{dt} \tag{3.1}$$

In this equation, u is the controller output, e the deviation from set point, K_C is the controller gain, and τ_I and τ_D are two constants called integral time and derivative time, respectively. One can set τ_I to infinity and τ_D to zero and use proportional control only, or assign a value to τ_I while setting τ_D to zero (P control or PI control, respectively). The parameters K_C, τ_I and τ_D can be set to different values according to what is given priority. In some cases, *fast* control is most important (like in compressor anti-surge control), while in other cases, *robust* control is given priority. What are the best values for the tuning parameters depends on whether control should be fast or robust, and there are several methods of deciding these values. These can be based on a process model, practical testing or both. Examples of such methods are Ziegler-Nichols and SIMC. Further reading about PID controller tuning: [9] and [10].

4.3 Control structure of the HYSYS model

As this project deals with dynamic simulation, the focus will be on finding a control structure that keeps the process stable – self-optimizing control, model predictive control is beyond the scope of this work. Therefore the control *structures* considered will all be rather simple – with feedback PI or PID control or, at most, simple cascade structures involving two controllers (PI or PID). Because it took quite much time to get the dynamic model running, there was not time for a detailed degree-of-freedom analysis (discussed in section 6.3).

When a simple control structure is desired, it is wise to pair variables that are close to each other physically. This means that when you want to control some variable in the outlet stream of a compressor, the manipulated variable should probably be one directly related to the compressor (for example feed flow rate, feed temperature or compressor speed).

For this model, there are two different kinds of process units that are the main focus; heat exchangers (especially the propane vaporizers and condenser) and compressors. The main target is to deliver the desired amount of natural gas and mixed refrigerant to the MCHE, at the desired temperature (here: -34°C). In a real LNG plant the goal will usually be to deliver as much LNG as possible at the correct temperature and with nitrogen content below some limit (typically a few %, see [2]). The maximum compressor power is the limiting variable in this situation. As the simulation here deals only with the pre-cooling part, it is natural to try keeping the exit streams from the process at desired temperatures, varying the amount of cooling water, the power input in the compressors and the different propane flows in the cycle.

The simplest way of controlling the process is to maintain temperatures constant at heat exchanger outlets and pressures constant at compressor outlets. The throughput of natural gas and MR are simply controlled by flow valves.

Compressors

In HYSYS there are two main ways to control a compressor that has a feed at given conditions. The compressor can be specified as constant-speed, or as variable-speed. The first does not require any compressor data like efficiency curves, the latter does. Typically the curves will be on the form efficiency vs. head (or flow) for different speeds. When curves are supplied, one can use the compressor speed as a manipulated variable to obtain the desired outlet pressure.

If one desires to use a constant-speed compressor, the simplest way to control the compressor is to control the outlet pressure by varying the energy input to the compressor. This is easiest done by using the compressor's energy stream as the manipulated variable or OP for the controller, and using the "control valve" option for this energy stream (fig. 3.2). By doing this, the compressor power is adjusted to give the desired discharge pressure. Fig. 4.1 illustrates how the controller is connected to the compressor and the discharge stream.

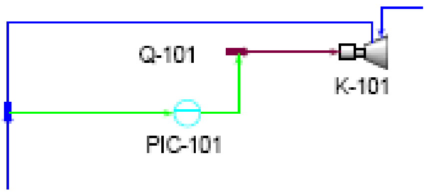

Figure 4.1: Control of compressor K-101

Heat exchangers

The variable to be kept constant for a heat exchanger is, as mentioned, the outlet temperature on the tube side (the natural gas/MR stream). There is one variable that could be manipulated for each exchanger, the opening of the choke valve one the propane feed to the exchanger.

If using cascade control, the temperature controller can give a set point to a secondary controller that manipulates the valve (fig. 4.2, left). The valve can be used to control either liquid level in the shell, the propane pressure, or the propane flow. In this model, though, controlling the liquid level in heat exchangers will not be of any use, because the basic heat exchanger model does not take this level into account when calculating the heat transfer.

If not using a cascade, the temperature controller will manipulate the valve directly (fig. 4.2, right).

For this project, a simple approach was taken – using the propane flow to the exchanger to control the exit temperature of natural gas /MR. Figure 4.2 shows this for the LP propane/MR heat exchanger, E-106. Notice that the TIC does not manipulate the valve directly, but instead gives a set point to the FIC. In

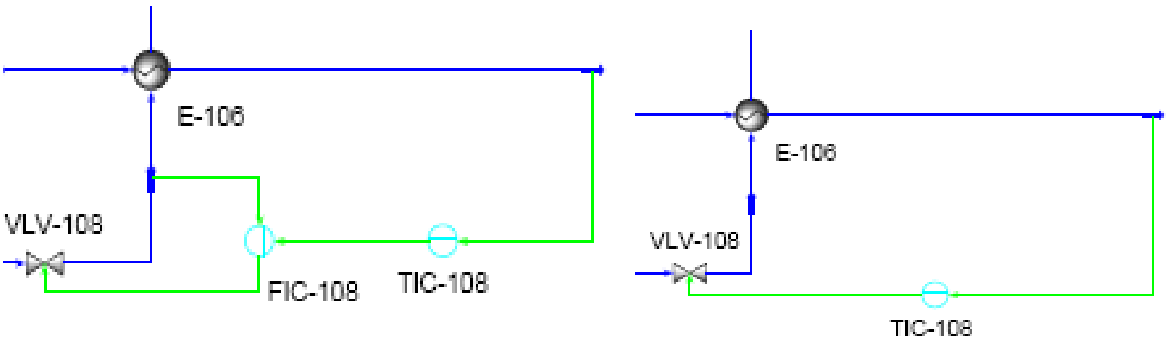

Figure 4.2: Control setup for heat exchanger (here: E-106). With cascade on the left, without on the right.

In addition, to keep the buffer tank from running out of liquid propane, a level control valve was placed on the outlet side of the liquid tank. This control valve was tuned automatically by HYSYS when installed.

The setup with cascades was called Control Scheme 1; the setup without cascades was called Control Scheme 2.

5. Simulation runs – testing of model and control schemes

5.1 Simulation scenarios

To test the model for possible problems and to check the different control schemes, two different simulation scenarios were run for the control schemes in question and some different variables tracked (the controlled/measured and manipulated variables and the disturbances). The scenarios were the following:

- Reducing the inflow of natural gas by 50 % (considered a large process upset) after 40 minutes and increasing it back to nominal value after 80 minutes. The changes were done indirectly – instead of defining the flow, the set point for the flow controller FIC-100 was defined.

- Increasing the natural gas feed temperature by 10°C after 20 minutes, and then decreasing it back to the nominal value of 21°C after 80 minutes. As opposed to the previous case, the temperature in the feed was defined directly.

Both scenarios were run for 120 minutes.

HYSYS's built-in Event Scheduler was used in order to introduce the disturbance after a given simulation time. For both cases, the event scheduler was used as following: A Sequence was defined, each Sequence with two Events: A change in the desired variable after 40 (20) minutes of simulation time, and a change back to the nominal value after 80 minutes simulation time. When defining an Event in the scheduler, one defines the following:

- The condition that should be met for the Event to happen – this could be a specific variable reaching a desired value, a variable stabilizing, or the simulation time reaching a certain value (the latter is used here, as mentioned above).

- The actions that should be carried out when the condition is met. This could be any manipulation of the process; typical manipulations will be set point changes for controllers, steps in disturbances, or defining new values for process variables or parameters.

Summarized, each of the scheduled sequences consists of two Events where each Event consists of one Action. When executing the schedules, the desired Sequence was started first, and then the Integrator which would run until the simulation time reached 120 minutes.

5.2 Cascade flow control on vaporizers, constant-speed compressors (control scheme 1)

In this scheme, outlet temperatures on the tube side of all heat exchangers were controlled by manipulating the flow of the cooling fluid (propane or water) by using a cascade as shown in fig. 4.2: The master controller (TIC) measuring the temperature of the stream and calculating a set point for the coolant flow, then giving the set point to a flow controller (FIC). The compressors were controlled by measuring discharge pressure and manipulating compressor power. In addition, there was a level controller at the outlet of V-100 (LIC-102). It can be argued that this controller should be dropped – see section 6.3.

For controller tuning, the parameters suggested in the HYSYS Dynamic Modelling Guide were used. In the guide, ranges were suggested for controller gain (K_C), integral time τ_I and derivative time τ_D. The least aggressive parameters were chosen (meaning the lowest value for K_C and τ_D and the highest for τ_I). The exception was the level controller (LIC-102) on the liquid tank; this controller was tuned automatically by HYSYS when it was introduced. Controller parameters for all the types of controllers are shown in table 5.1.

Table 5.1: Controller parameters used in Control Scheme 1

Type of controller	K_C	τ_I /min	τ_D /min
Flow (FIC) *)	0,40	0,250	
Temperature (TIC)	2,00	10,0	1,0
Pressure (PIC)	2,00	10,0	
Liquid level (LIC)	1,80	17,6	

*) For FIC-100, K_C = 0.25

Results from the simulations are shown as plots of set points, process variables and manipulated variables for different controllers in the process. In all the figures, the red curve represents the set point, green the process variable and blue the valve opening in % (or more generally, the value of the manipulated output in % of the maximum value). Also notice, for the cases where the numerical value of the OP is not shown in the figure, that when the blue curve levels out, this indicates that the manipulated output has reached saturation. This means the corresponding valve is completely open, completely shut or the corresponding compressor is running at full power.

Because of space considerations, the figures do not include scales for all variables – typically only for the process variable. The auto-scaling function was used to give good and readable figures. For the plots not included here, see the HYSYS files.

Step in natural gas feed flow rate

First, a simulation was run just to see whether the controller tuning was anywhere near reasonable. This showed that the natural gas feed controller, FIC-100, could not handle a 50 % reduction in its set point with the given tuning parameters. The PV and OP would both oscillate violently. However, the return to the initial set point was smooth. The measure taken was to reduce the controller gain K_C from 0.40 to 0.25. This gave smooth controller behaviour, almost without noise. For a comparison of the two tunings, see figure 5.1a-b. The rest of the process seemed to behave OK, so the settings listed in table 5.1 were kept.

The most important results were the temperatures at the outlets from the low-pressure propane vaporizers (that is, the temperatures of NG and MR to the MCHE). Figures 5.2a and 5.2b show how the outlet temperatures changed with time. Notice the large fluctuations in the manipulated variable – reaching saturation several times during the simulation. However, this seems to give good set point tracking.

Figures 5.3a and 5.3b show how pressure varies with time in two of the propane compressors, K-101 and K-102. The low-pressure compressor showed the same behaviour. The level controller LIC-102 kept the liquid level in the tank close to set point, see figure 5.4. For the MR compressors and intercoolers, the process conditions varied little – as expected since no disturbance was introduced in the MR stream. Figures 5.5 and 5.6 illustrate this point – that the MR compressors and MR inter coolers were little influenced by the downstream disturbances.

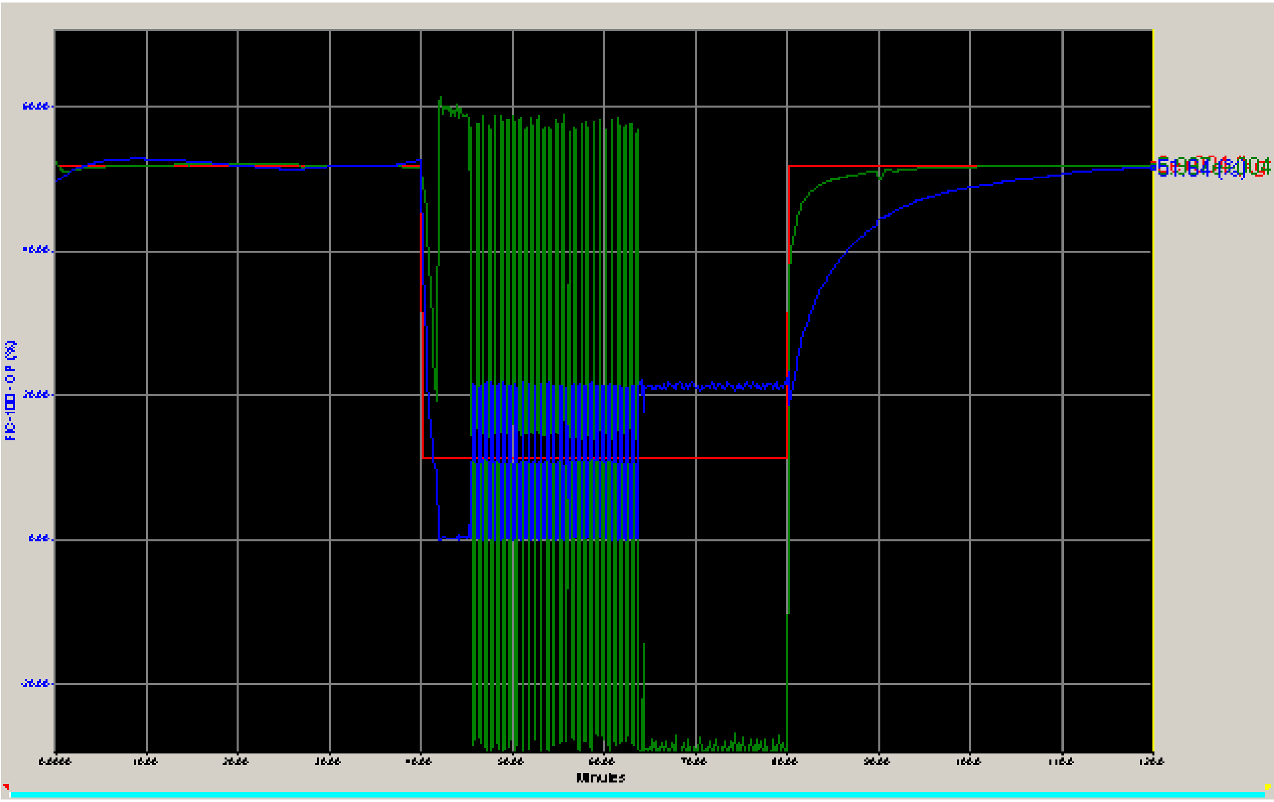

Fig. 5.1a: Violent oscillation with K_C = 0.4

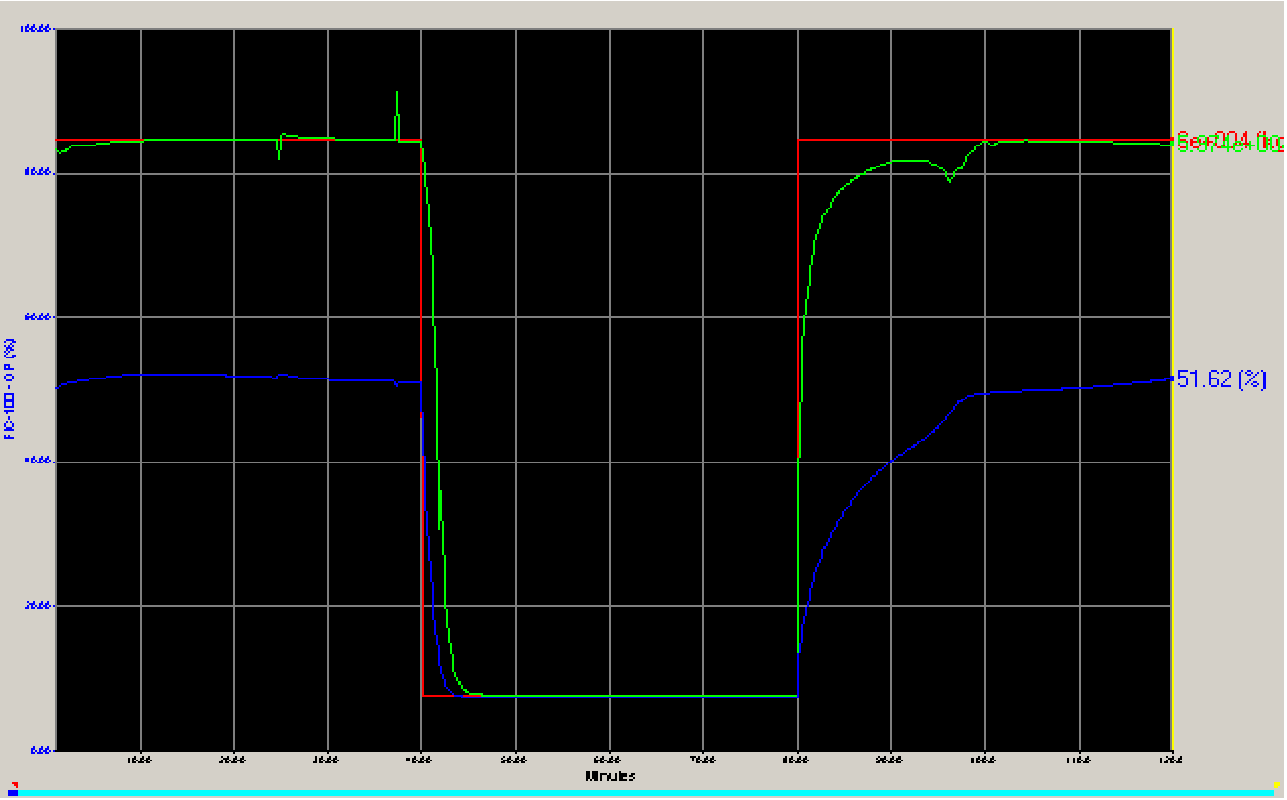

Fig. 5.1b: Smooth behaviour with $K_C = 0.25$

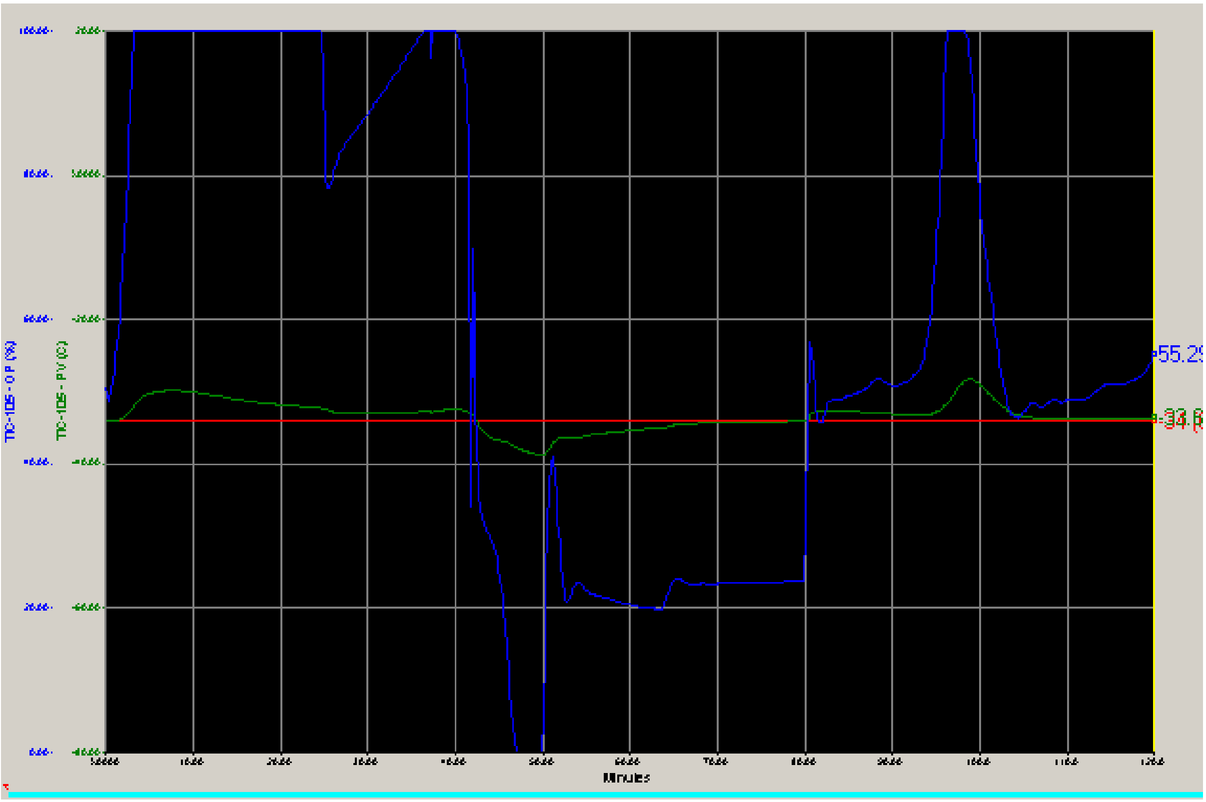

Fig. 5.2a: Temperature at tube outlet, E-103 (natural gas to MCHE)

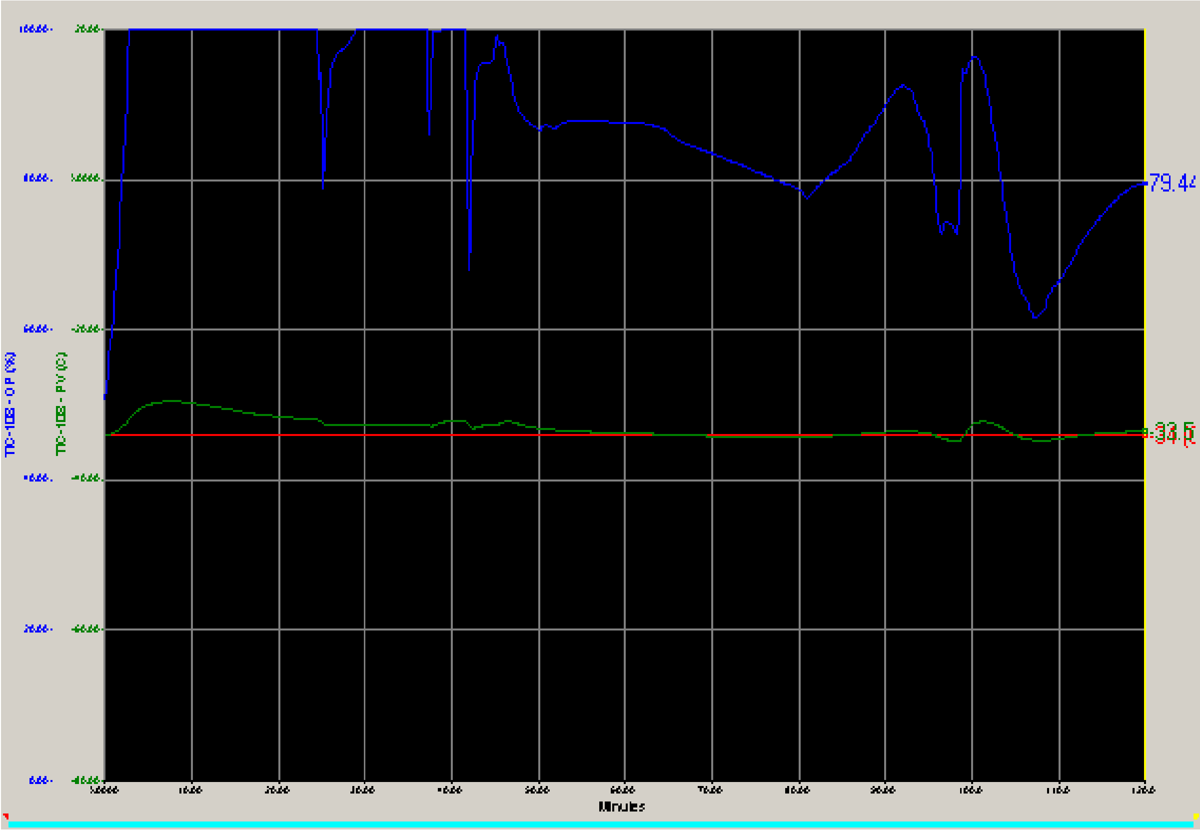

Fig. 5.2b: Temperature at tube outlet, E-106 (MR to MCHE)

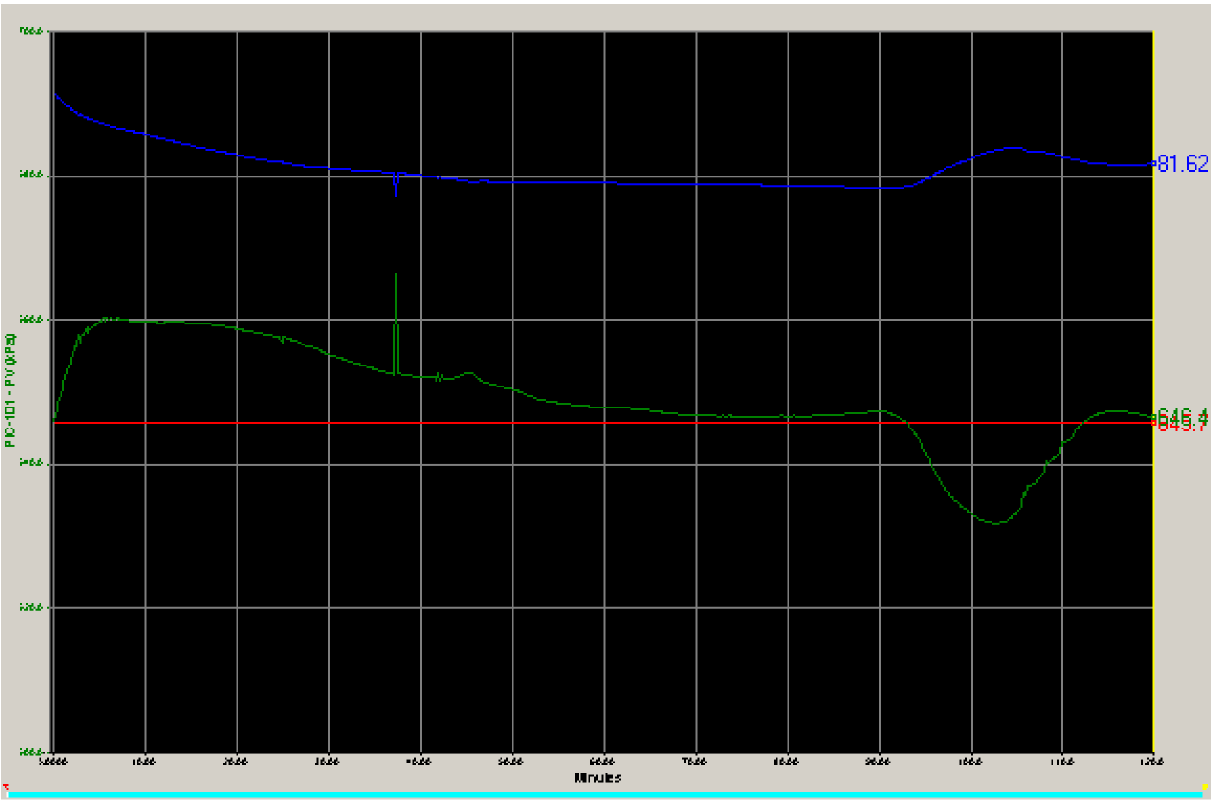

Fig. 5.3a: Pressure at outlet of K-101 (green), set point (red) and compressor power in % of max (blue)

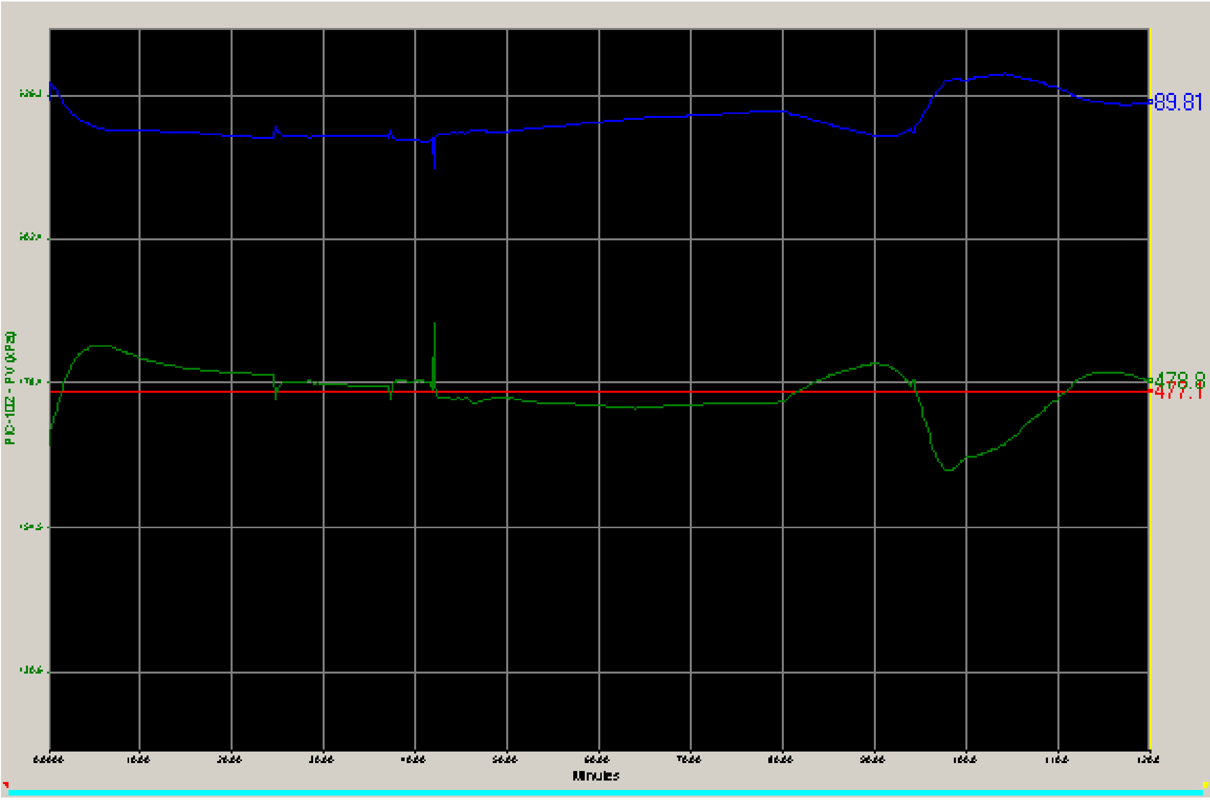

Fig. 5.3b: The same as in 5.3a, but for compressor K-102

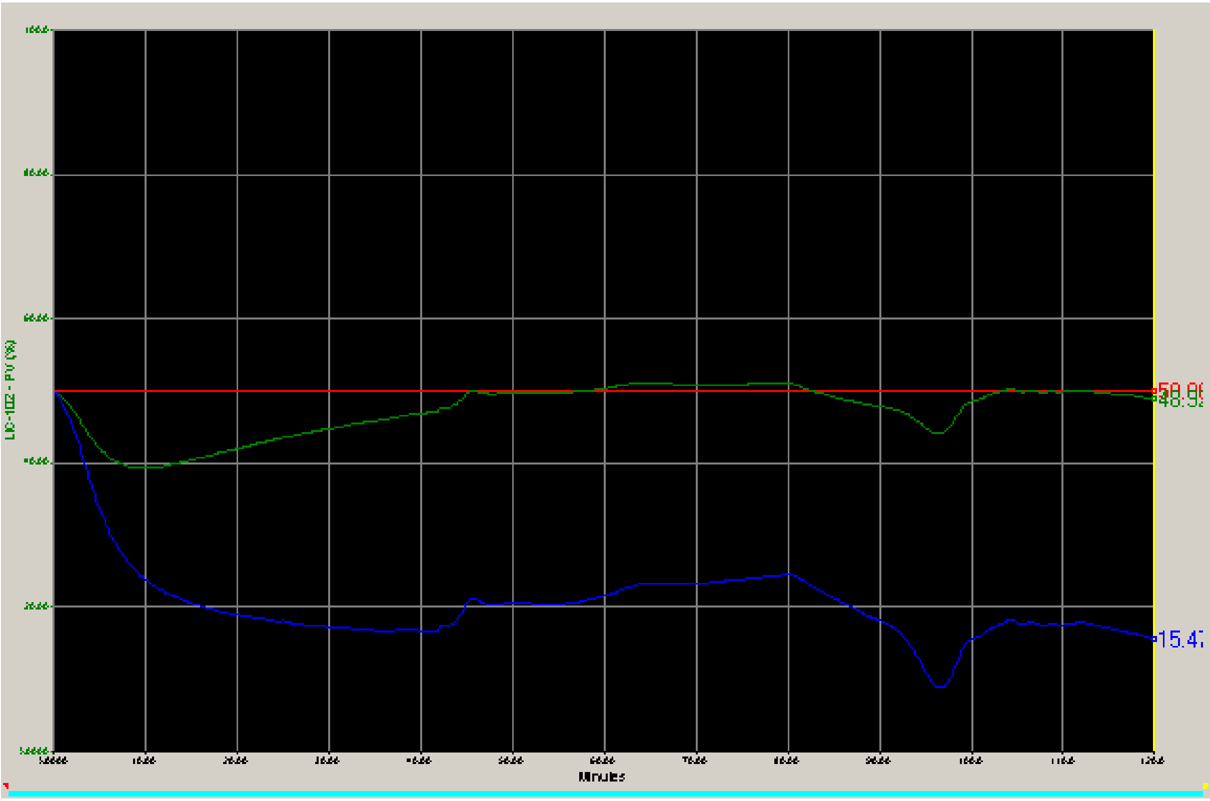

Fig. 5.4: PV, SP (% liquid level) and OP (% valve opening) for LIC-102

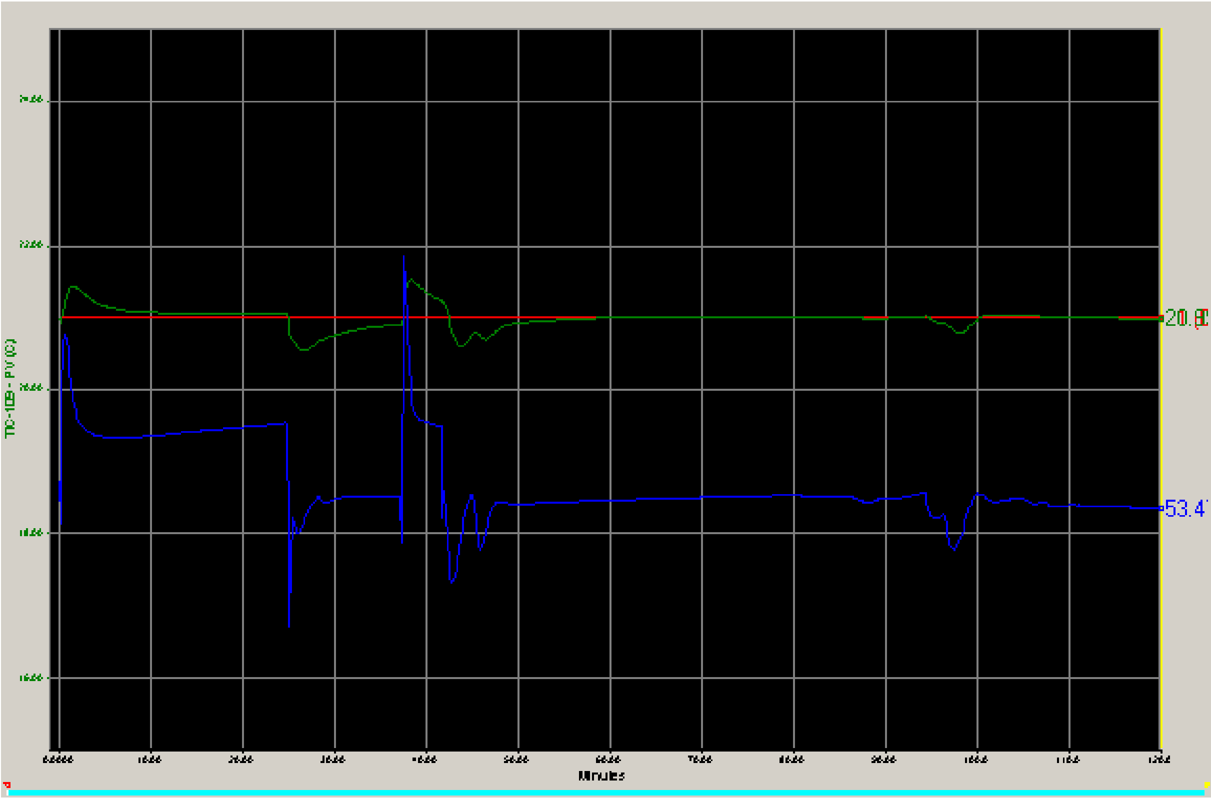

Figure 5.5: PV, SP and OP for TIC-109 (MR after cooler). OP varies between 45% and 75%.

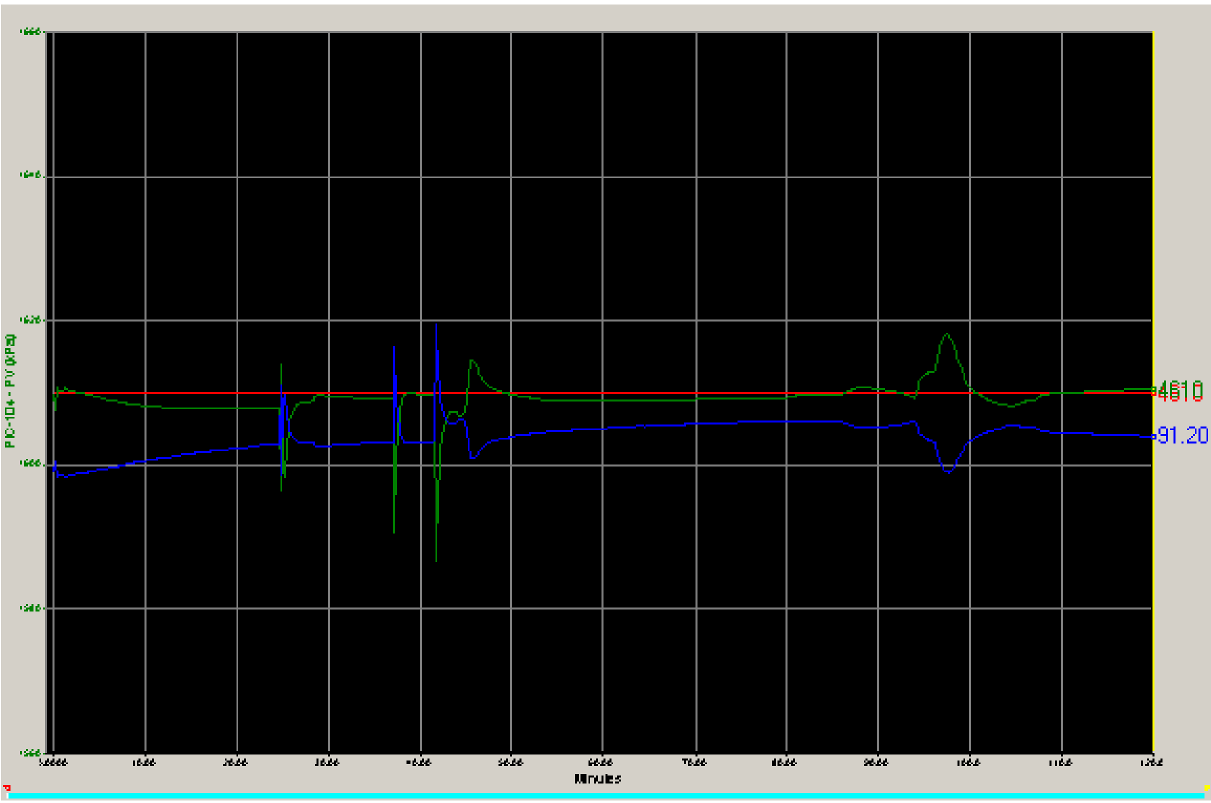

Figure 5.6: PV, SP and OP for PIC-104 (HP MR Compressor discharge pressure)

Step in natural gas feed temperature

In this case, the temperature of the natural gas feed (34) was increased from the initial 21°C to 31°C after 20 minutes of simulation time, and decreased back to 21°C after 80 minutes. All controller settings and initial conditions were the same as in the first simulation.

Figure 5.7 shows how the natural gas inflow varied with time. The temperature change gives a disturbance in the molar flow because the molar volume of a gas depends on temperature. Figures 5.8a and 5.8b show how TIC-103 and FIC-103 responded to the disturbance. TIC-103 is the temperature controller at the outlet of E-101 and FIC-103 is the flow controller that receives the signal from TIC-103. Figures 5.8c and 5.8d show the same for TIC-104 and TIC-105, the temperature controllers after E-102 and E-103. One can see that the temperature after the last propane vaporizer is not down to its set point until at the end of the simulation time. Also notice that the propane flow to this exchanger is the first of the three that *leaves* saturation conditions after the feed temperature has dropped back to 21°C – the OP for TIC-104 is nearly on saturation at this point. Figures 5.9a and 5.9b illustrate that the compressors behaved well throughout the simulation.

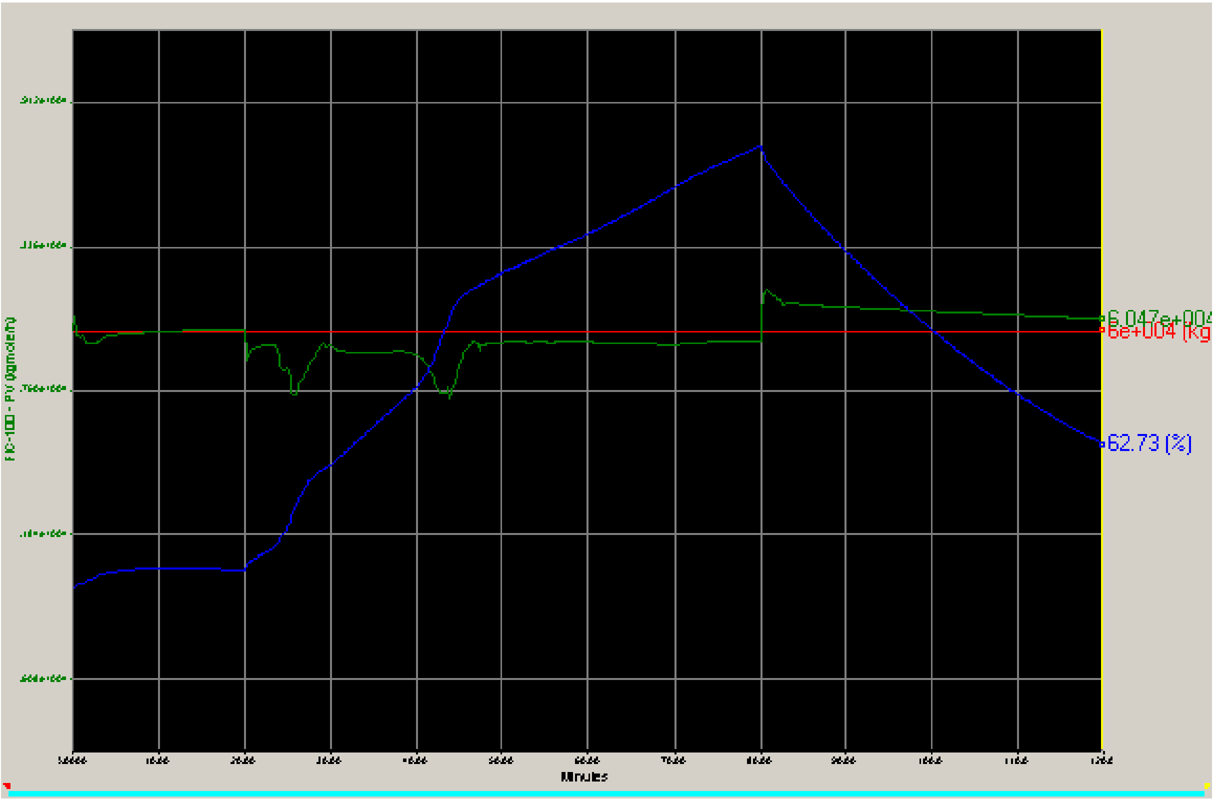

Figure 5.7: PV, SP and OP for FIC-100. Notice the slow response.

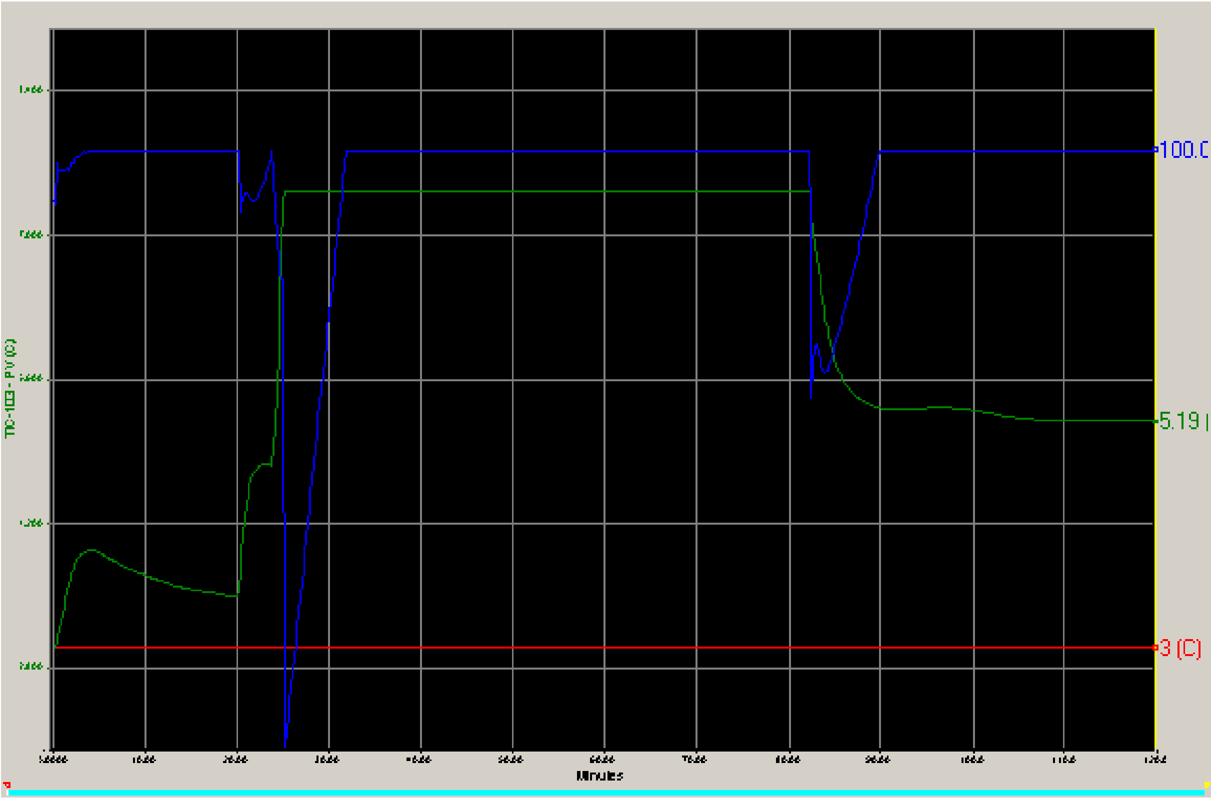

Figure 5.8a: PV, SP and OP for TIC-103. The output goes to saturation.

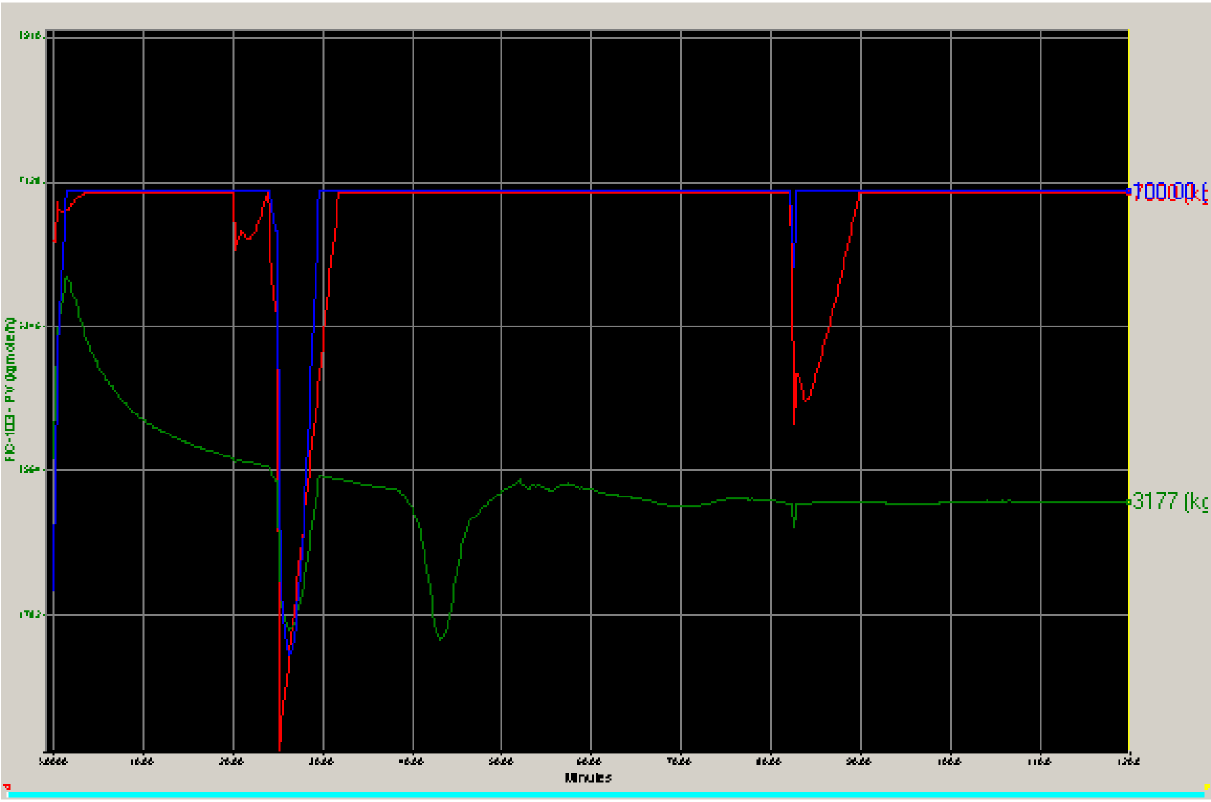

Figure 5.8b: PV, SP and OP for FIC-103 (the “slave” of TIC-103)

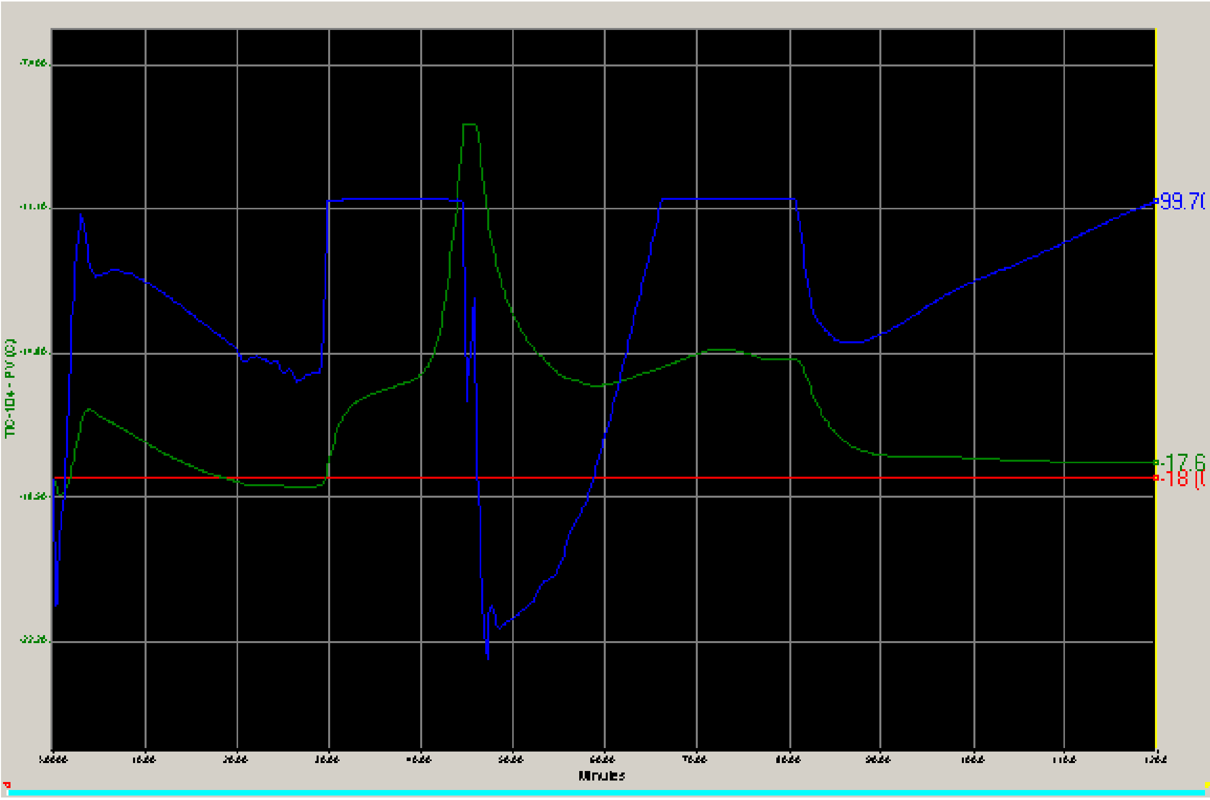

Figure 5.8c: PV, SP and OP for TIC-104

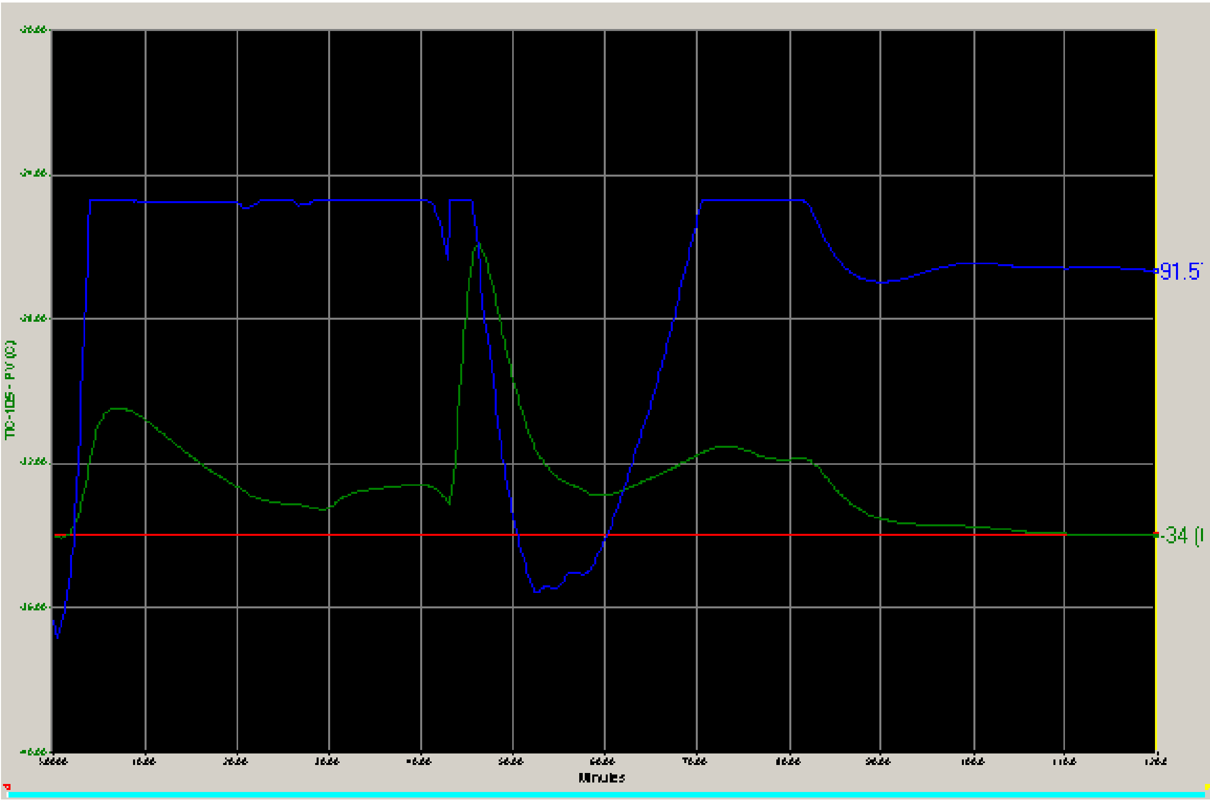

Figure 5.8d: PV, SP and OP for TIC-105

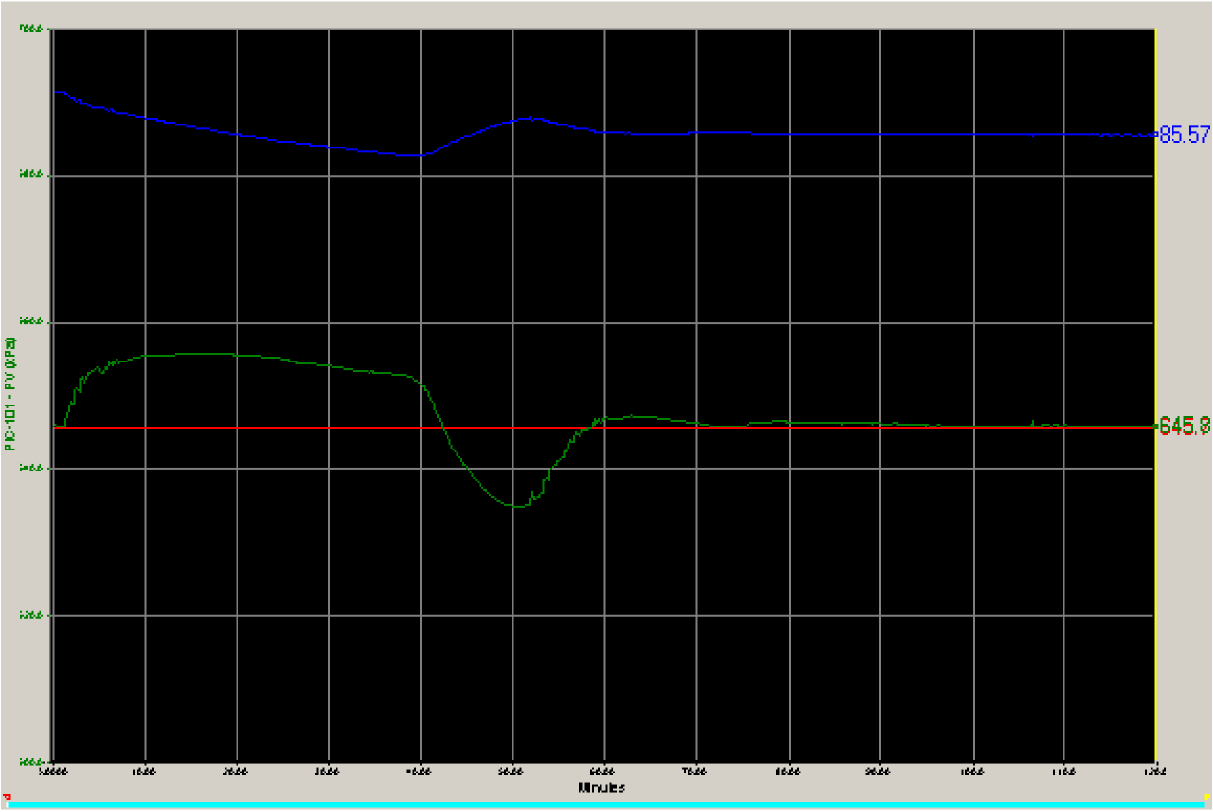

Figure 5.9a: PV, SP and OP for PIC-101

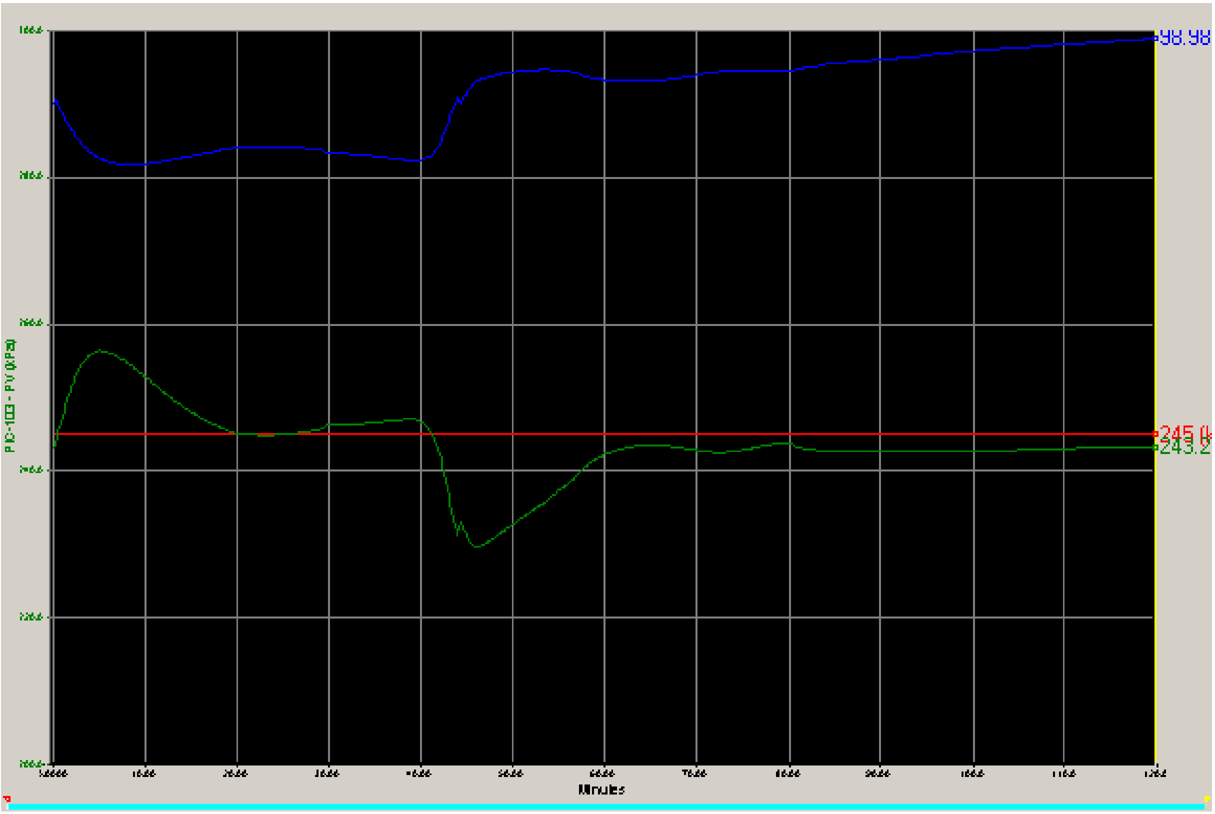

Figure 5.9b: PV, SP and OP for PIC-103

These figures show that the compressor capacity was not a problem in this scenario – the compressor power never went to saturation.

5.3 Effect of omitting cascades - control scheme 2

To see whether cascade control had any real effect in this case, the same simulations were run with a slightly changed controller setup: Instead of letting the temperature controllers on heat exchanger outlets manipulate a set point, they would now manipulate the propane flow directly. For the compressors, the same control setup as in scheme 1 was used. For all controllers, the same tuning parameters as in scheme 1 were used (table 5.1).

Step in natural gas flow rate

The same simulation as for control scheme 1 was carried out, with a step in the set point for the feed flow rate after 40 min and with a step back to the nominal value after 80 min. Figures 5.10-5.15 show PV, SP and OP for different controllers in the model. Figure 5.11a shows these variables for TIC-103, which manipulates propane flow to E-101. This figure shows oscillation after some time – but the set point tracking seems OK.

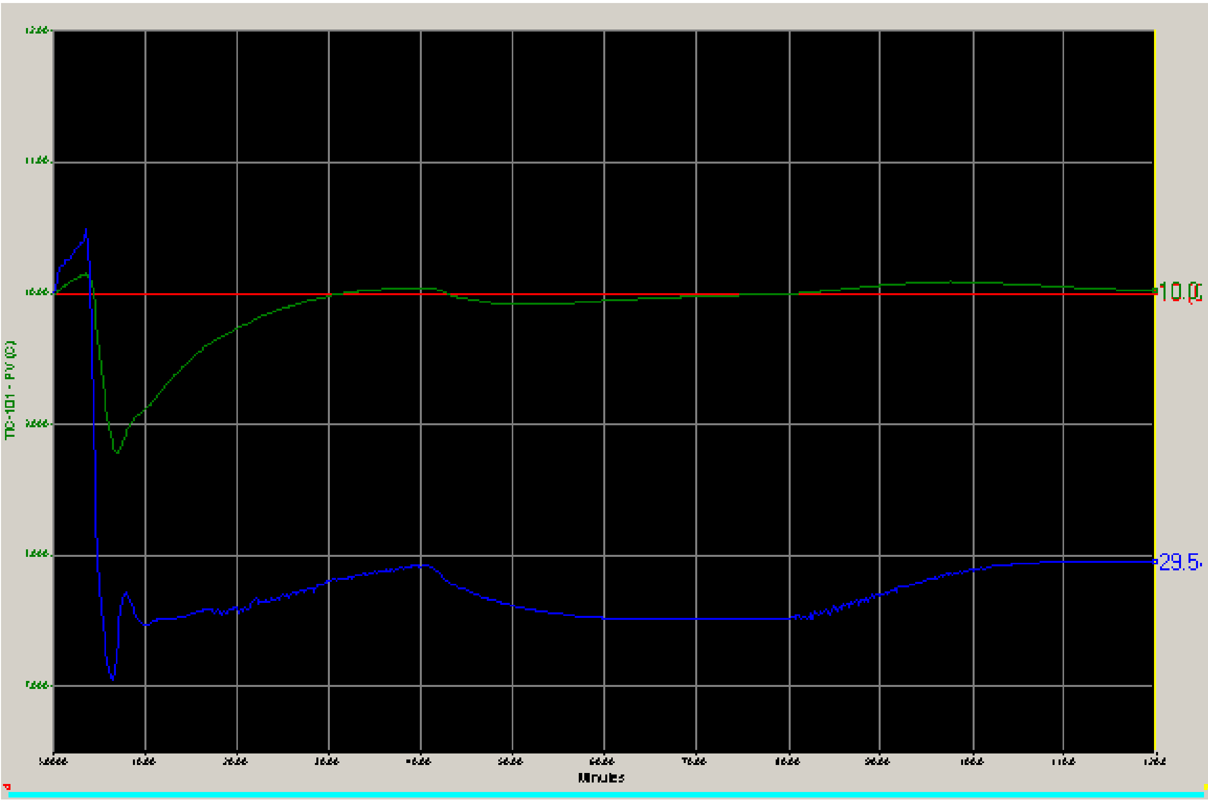

Figure 5.10: PV, OP and SP for TIC-101 (temperature after propane cooler)

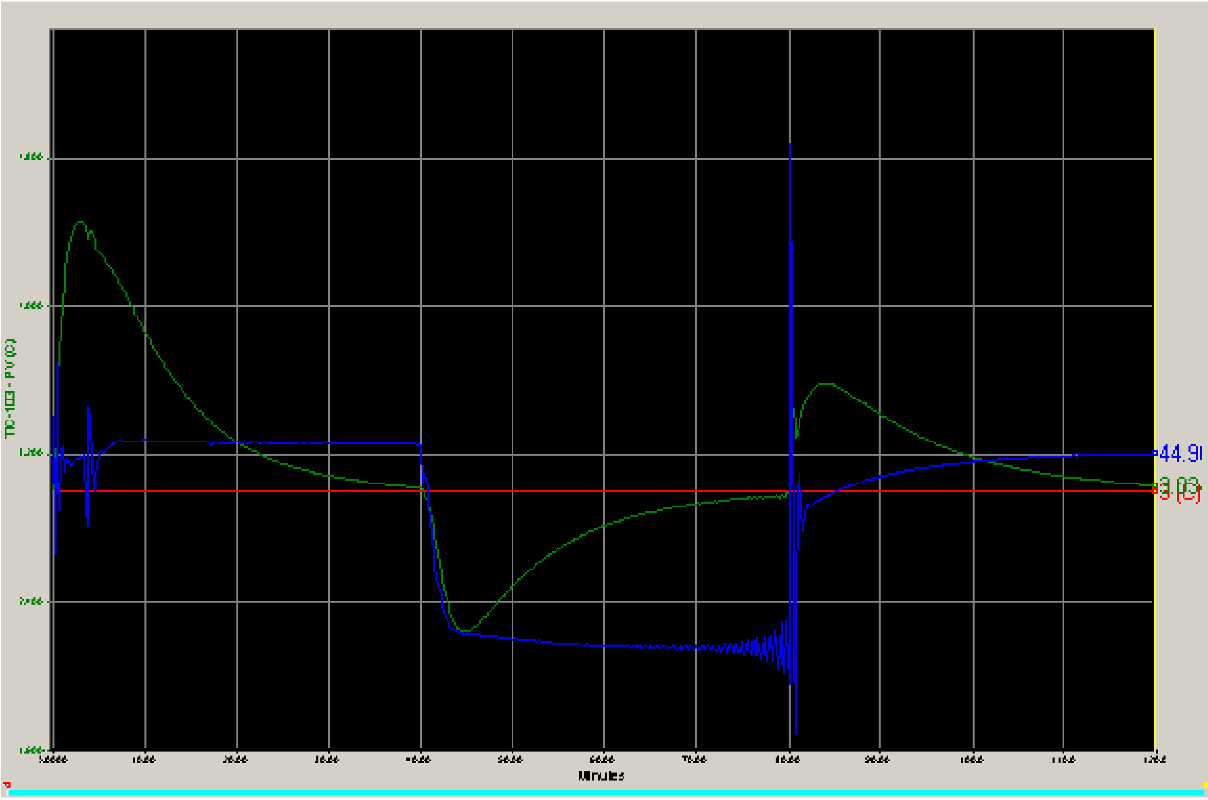

Figure 5.11a: PV, SP and OP for TIC-103 without cascade control

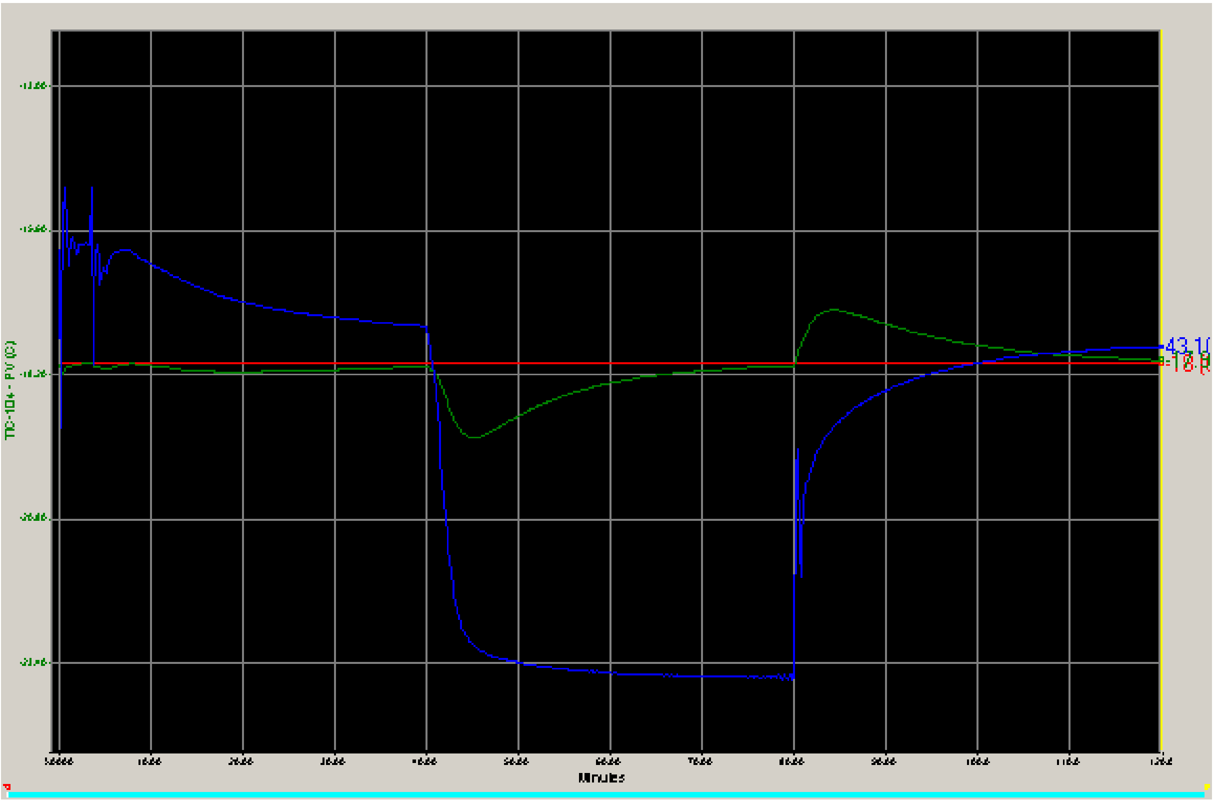

Figure 5.11b: PV, SP and OP for TIC-104

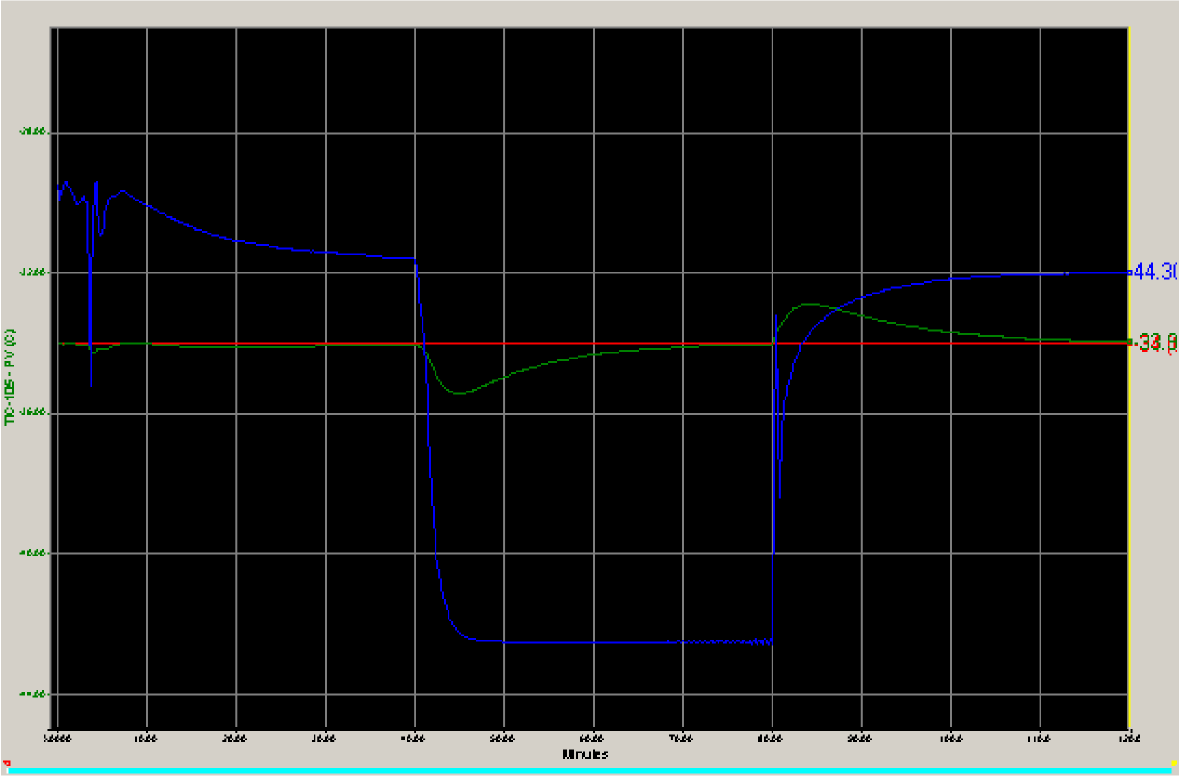

Figure 5.11c: PV, SP and OP for TIC-105

Figures 5.11a-c show that in fact, all the controllers show tendencies to oscillating behaviour, but the oscillations seem to be smaller for the last two. The exchangers E-104, E-105 and E-106 are little affected by the disturbance, and the controllers on the MR streams from these exchangers do not show oscillation. Figure 5.12 illustrates this for E-104 (TIC-106).

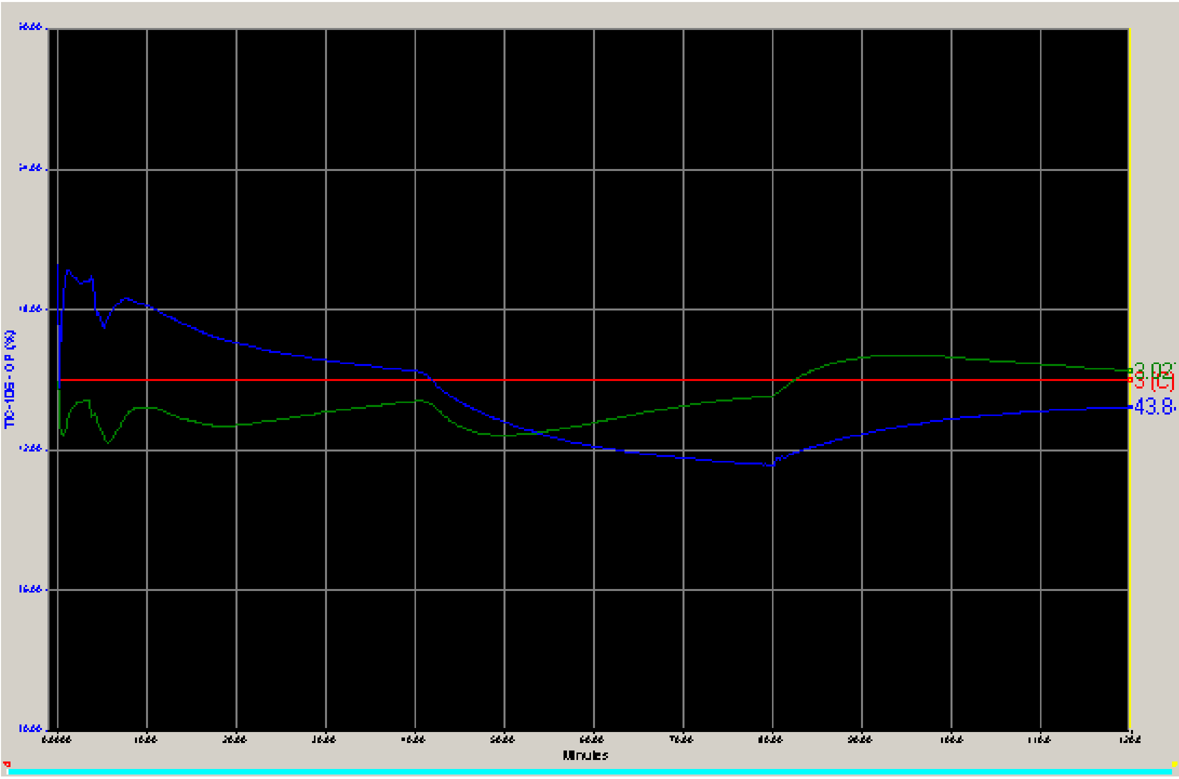

Figure 5.12 PV, SP and OP for TIC-106

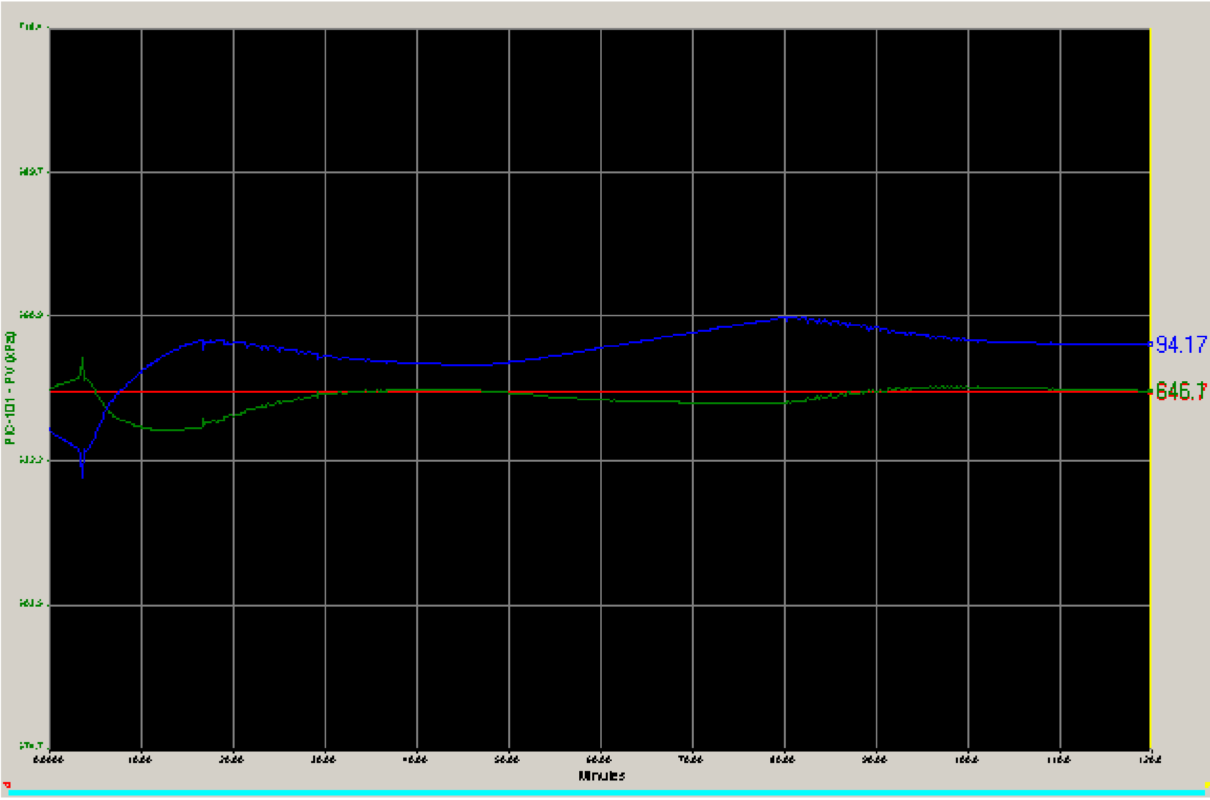

Figure 5.13: PV, SP and OP for PIC-101

Figure 5.13 shows the discharge pressure in K-101; one can see that this compressor never needs to go to full power in order to maintain the correct pressure. However, the other compressors go to full power in some periods, as indicated by figure 5.14.

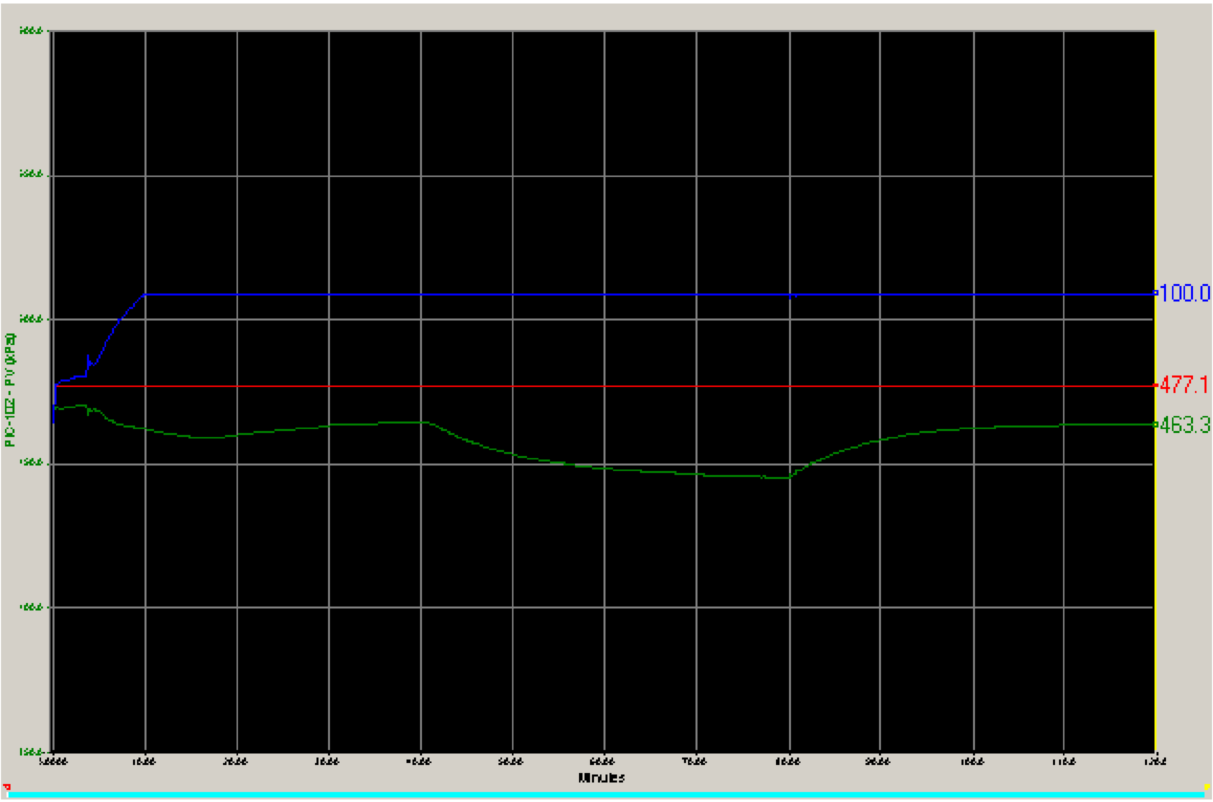

Figure 5.14: PV, SP and OP for PIC-102

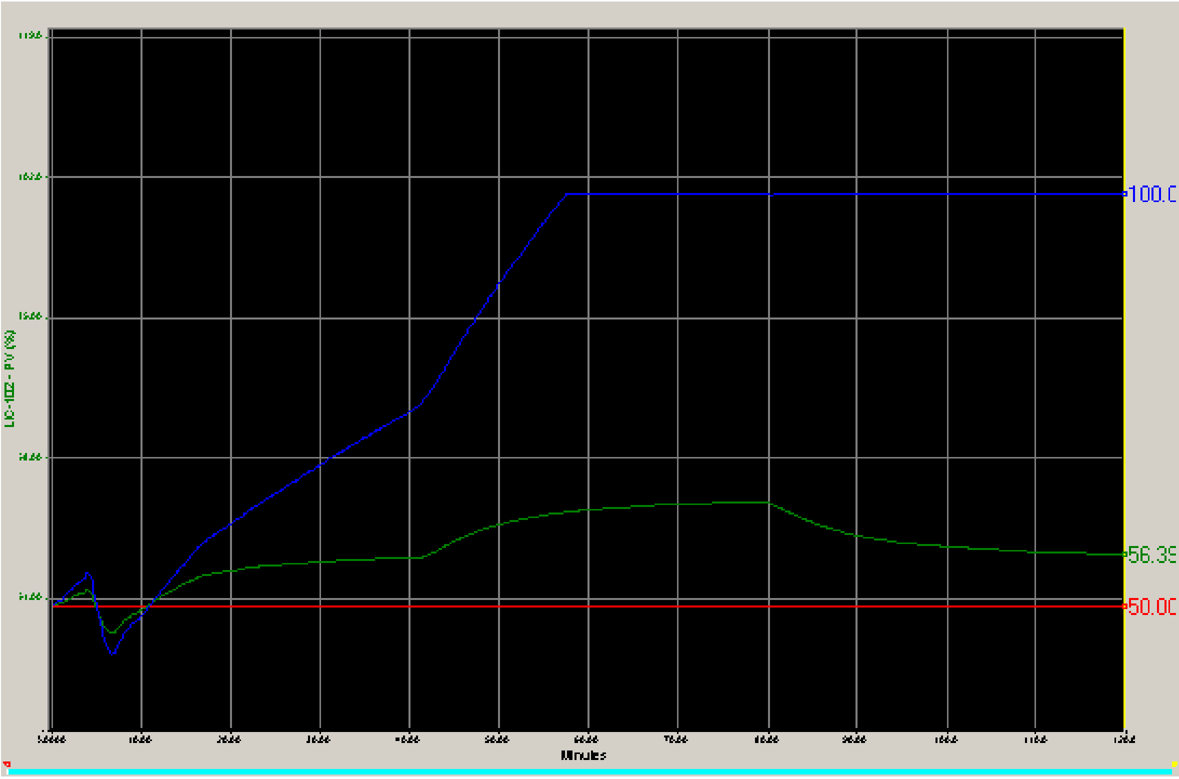

Figure 5.15: PV, SP and OP for LIC-102

In this simulation, the level control valve actually goes to 100% open and stays there, as seen on figure 5.15.

Step in feed temperature

The temperature was again increased by 10°C after 20 minutes and decreased by 10°C after 80 minutes. This simulation showed results that were remarkably different from the previous; the temperature controllers TIC-103, TIC-104, TIC-105 all show oscillation (see figures 5.17a-c). Figure 5.16 shows that the level was kept close to set point throughout most of the simulation.

Figure 5.18 illustrates that the oscillations caused by the temperature controllers also influences on the compressor power – because valve position affects downstream pressure. The cooling requirement in the propane condenser also was affected, as seen from figure 5.19 where the OP shows oscillation.

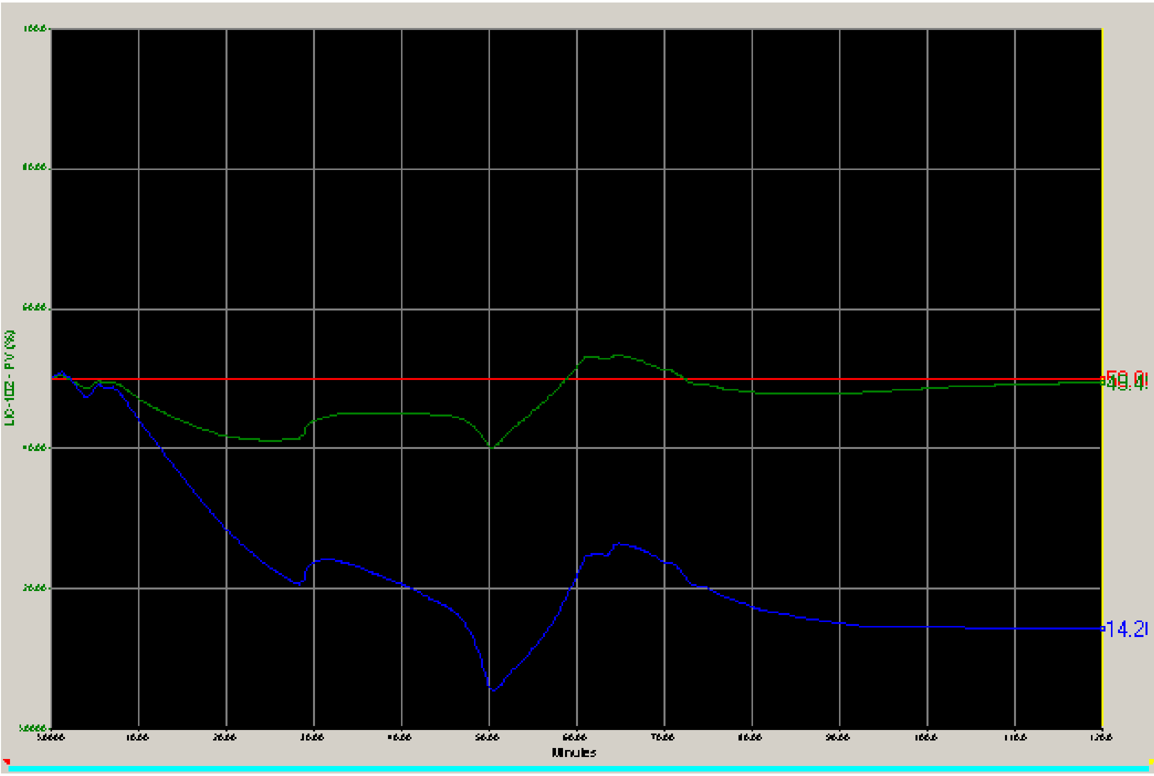

Figure 5.16 SP, PV and OP for LIC-102 (V-100 level control)

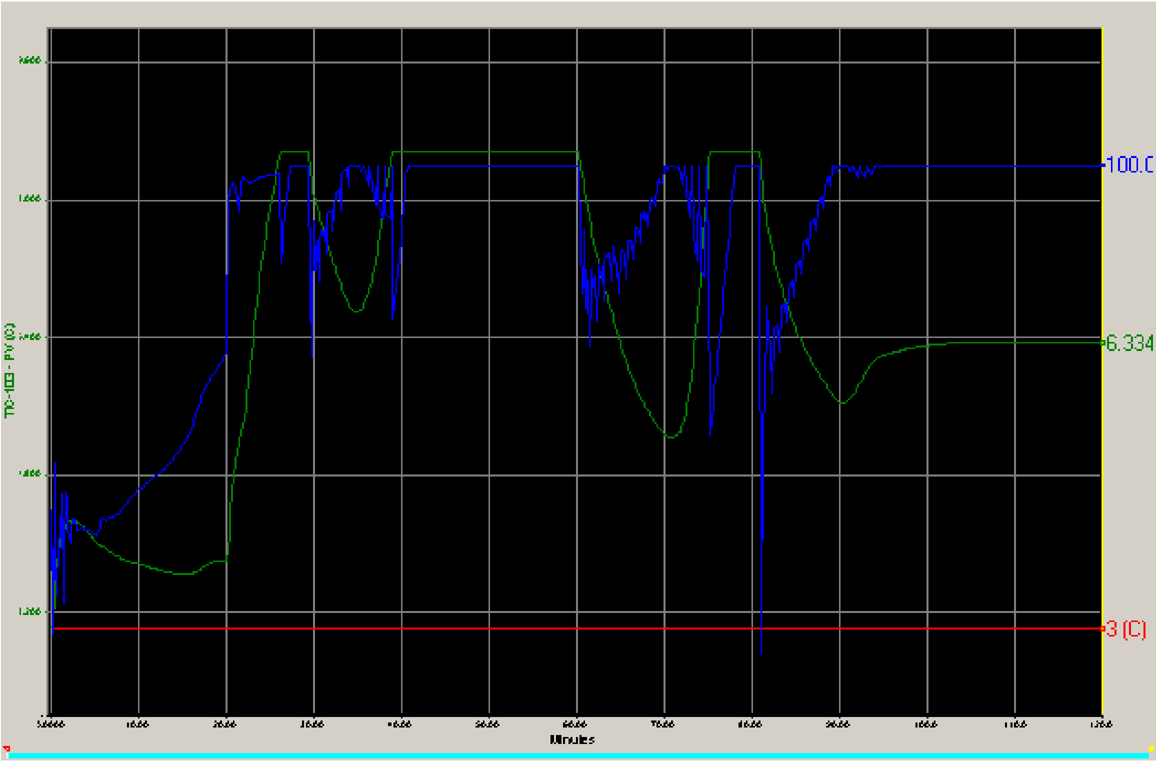

Fig. 5.17a: SP, PV and OP for TIC-103

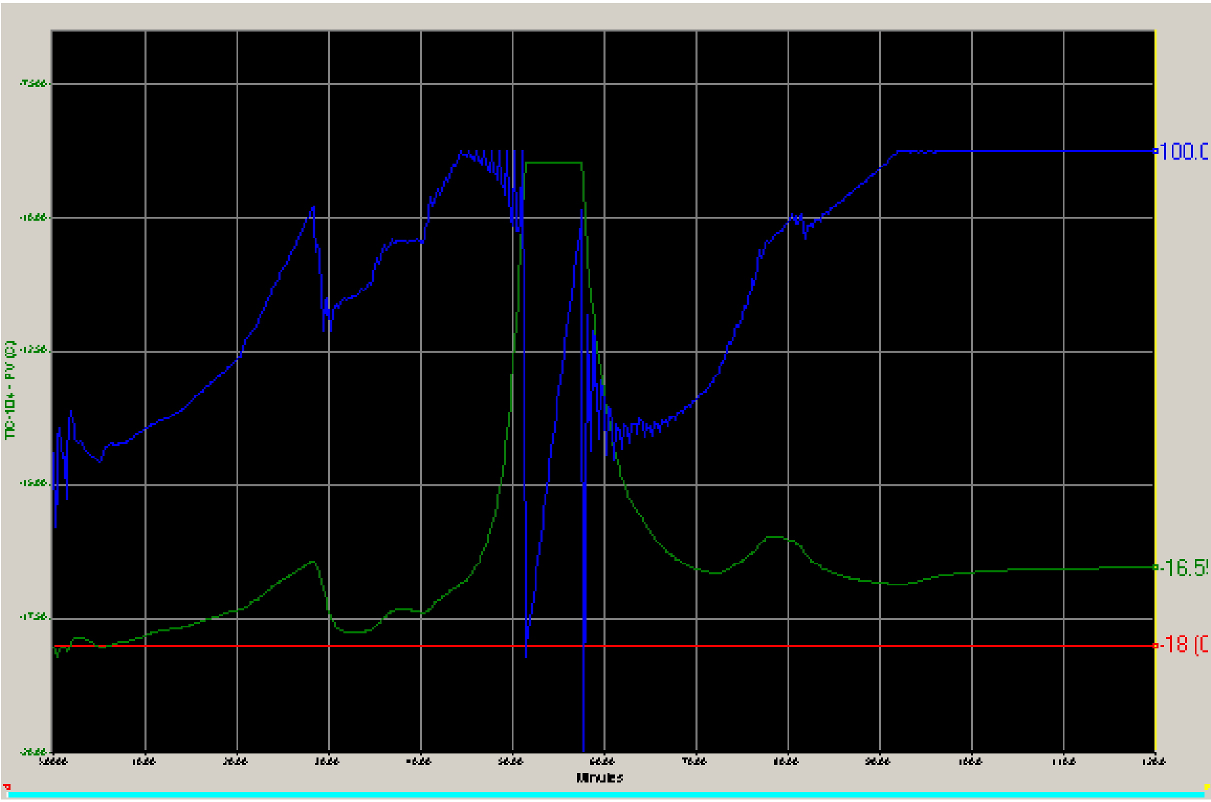

Fig. 5.17b: SP, PV and OP for TIC-104

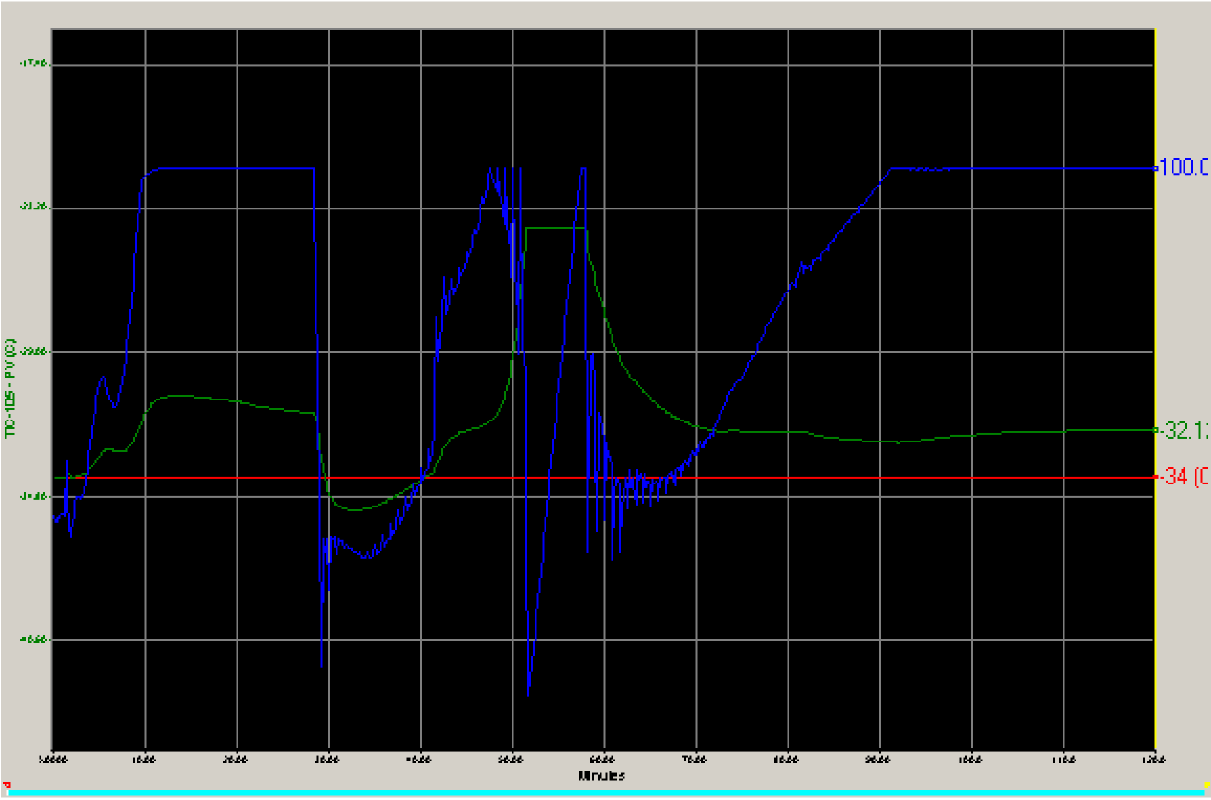

Fig. 5.17c: SP, PV and OP for TIC-105

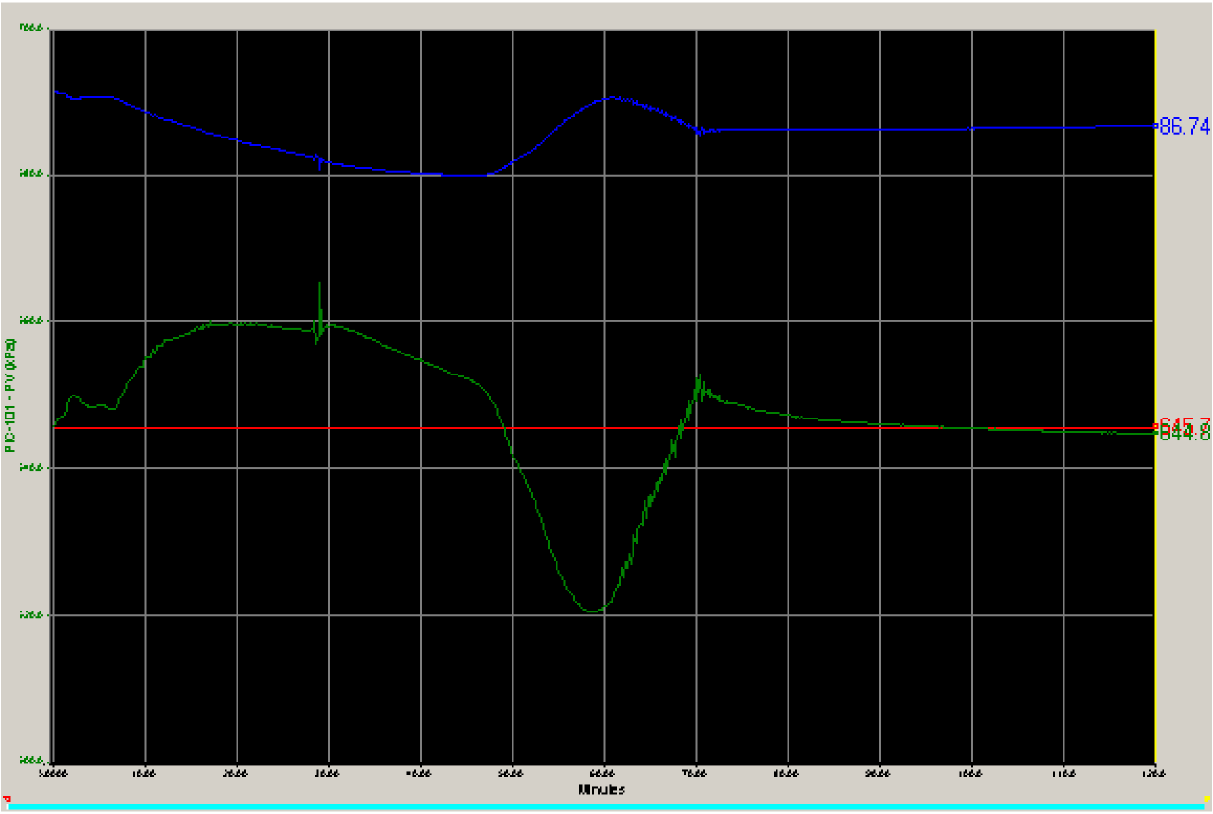

Figure 5.18: SP, PV and OP for PIC-101

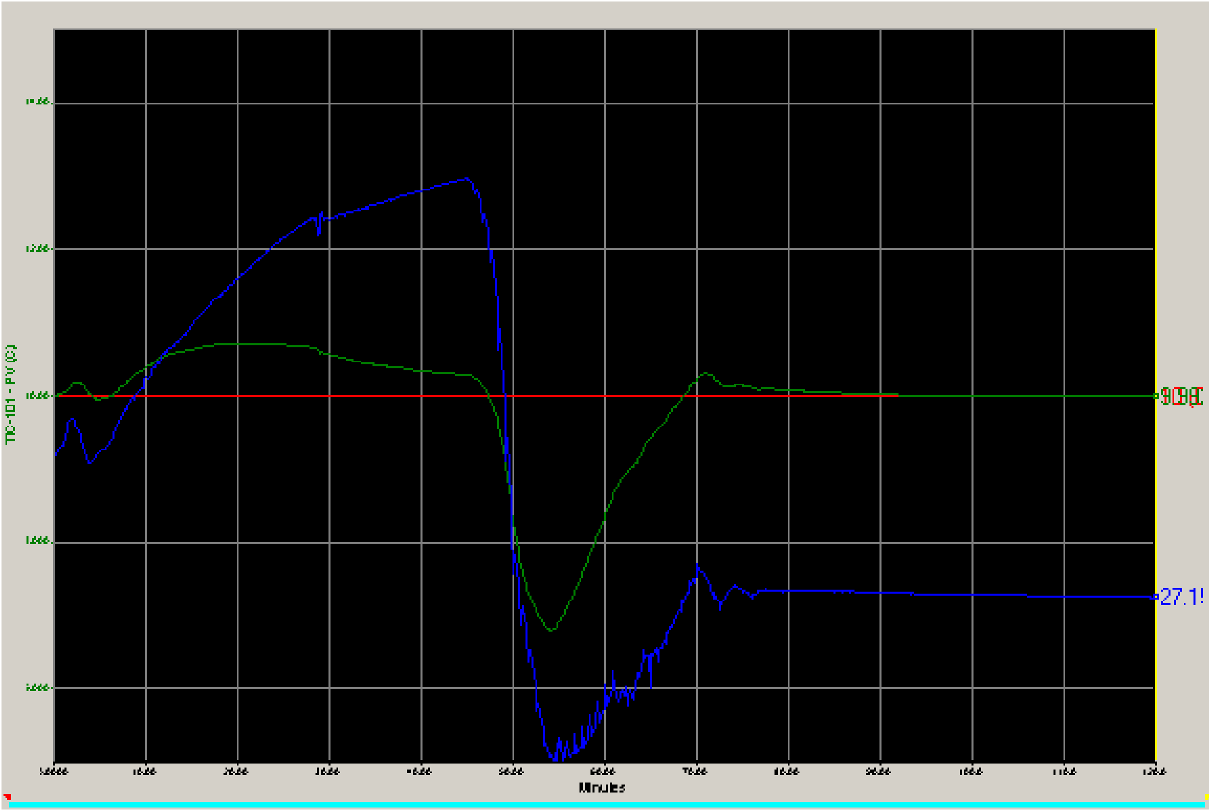

Figure 5.19: SP, PV and OP for TIC-101 (the propane condenser)

6. Discussion

6.1: About the model development, steady state and sizing

Building a steady state model of the process was fairly simple as it was a simplification of an already existing model. However, building the original model required quite much time. Main difficulties were to avoid over specifications in the propane loop. For natural gas and MR streams the modelling was fairly straight-forward – by specifying the inlet conditions one could easily work through the process and adding only one new specification after each process unit (a temperature at cooler outlets, a pressure at compressor outlets) and adding other specifications to the unit operations (like pressure drops and compressor efficiencies – for the latter, default values were provided by HYSYS).

The propane loop modelling was somewhat more difficult as there was a risk of over-specifying some process units, especially when a tank and a new valve (V-102) were introduced after the cycle had been closed.

All variables that were actually known were the natural gas/MR temperatures in and out of the propane vaporizers and the temperature levels of propane. All flows and pressures were unknown – so an assumption had to be made to establish a pressure at some point. The choice fell on the outlet stream of the propane condenser, which was assumed to be saturated liquid at 10°C – this determined the pressure as well. The problem with this assumption was that the choking valves would make the propane stream flash – so the feed to the vaporizers would not be 100 % liquid.

To obtain 100% liquid after each valve, this would have to be specified explicitly for each stream in question – this would lead to different pressures at stream splits with the current model layout. This would not be acceptable in dynamics where one will typically specify that all streams entering or leaving a tee or a mixer have the same pressure. (HYSYS recommends using the “equalize all”-setting for pressures in the mixer unit operation). Therefore it was accepted that the feed to the vaporizers contained vapour.

To find the propane flow needed to obtain the desired temperatures in each NG/MR stream, one further assumption had to be made; the propane stream that left each exchanger was defined as being saturated vapour, i. e. no superheating. When changing to dynamics, this was assumed to be OK as long as the outlet nozzles on the shell side were placed on top – only vapour should leave.

Sizing was a fairly simple task for valves – by using a standard pressure drop, a standard valve opening (50 %) and the actual flow in the steady state model, HYSYS could calculate the valve constant, C_V, given the valve opening function (in this case, linear). The constants calculated seemed to work well in dynamic simulation. Also for the liquid tank the sizing procedure was simple – the interface accepts that volume and orientation are specified, once these are given, it calculates diameter and length. The tank should have been larger, though – see section 6.3.

The heat exchanger sizing was a major challenge, because HYSYS does not support multiple-shell heat exchangers. In the original stationary model, the heat exchangers were designed using TASC, which suggested exchangers with from 6 to 10 parallel shells. This was due to limitations in shell and tube diameters and lengths combined with limitation in fluid velocity in the exchangers. TASC takes equipment vibration into account and will typically suggest using more parallel shells in order to reduce the flow through each. Using the TASC data combined with the detailed rating model would, if HYSYS could handle them properly, give very precise heat transfer and pressure drop data. However, with standard tube diameters (inner =16 mm and outer = 20 mm) and length (=6 m), and with a heat transfer area in the same order of magnitude as in the TASC design, the heat exchangers would get volumes that resulted in extremely short residence time (below 1 s). This can not be tolerated in dynamic simulations – the simulation can become unstable, and it is totally unrealistic that a liquid stream has a residence time of t. e. 1 s in a large vessel. This was clearly seen when testing the model in dynamic mode – balance equations for the heat exchangers would crash after a few iterations (when inspecting the equation summary view, the biggest errors in unconverged equations would be in the propane vaporizer equations, indicating that these were responsible for the crash). This is easy to understand when considering that the default step length in the integrator is 0.5 s, and the step length used in the simulations was 0.05 s. With the small residence time one could face a situation where an exchanger goes from empty to full in only one integrator step.

The approach taken to this problem, using the basic model instead, reduces the realism in the model significantly. The reason is that the pressure drops and the product UA for the heat exchangers are calculated from simple relations (like equation 2.2 of chapter 3) with constants found from the steady state pressure and flow data, instead of being calculated from viscosity, velocity and tube diameter. Heat exchanger rating for dynamic simulations is clearly an area where HYSYS could be improved – the same goes for the advanced rating methods in steady state.

Another drawback of the heat exchanger model is that kettle-type exchangers (K-shell), with two outlets on the shell side, are not allowed for (with the exception of column reboilers). This means that in steady state mode, one has to model such a heat exchanger as an ordinary shell-and-tube exchanger, followed by a flash tank as mentioned in section 3.2. In dynamic mode, the exchanger can be modelled as a separator tank with a tube bundle, where the propane is in the separator and the natural gas/MR in the tubes. Neither of these is really close to the actual kettle-type exchanger. In the dynamic model used in this project, the problem is avoided because the exchangers are now meant to vaporize all propane that flows through them – they have only a vapour outlet, no liquid outlet.

One point of uncertainty will of course be the heat transfer coefficients that should be used. In this work they were based on an average of the ones obtained in the TASC design of exchangers, this is of course a simplification, but accurate data for heat transfer are difficult to obtain from general literature.

Compressor rating was the least intuitive sizing task, as hardly any guidelines exist for this in the literature used. Specifying the efficiency of a compressor and supplying enough stream data for it to calculate outlet conditions and work is sufficient for simple simulations, but for more realistic rating compressor curves are needed. Such curves are usually available for real compressors, but finding realistic curves for a compressor where nothing is known except flow and pressure rise is difficult. A nice feature would be a kind of “standard” compressor curve selection, at least the HYSYS documentation should contain some guidelines for making approximate compressor curves when these are not known from a real process. Such guidelines exist for other process model parameters like pressure drop across valves and nozzles, PID controller tuning and separator/tank sizing. [5] / [6]

Steady state to dynamics, initialization

Switching from steady state to dynamic was a relatively easy task once all units had been sized. HYSYS also has an initialization option that can be used, but as all variables are known when a converged steady-state model exists, this option was not used here. One feature that was useful was the Dynamic Assistant for finding possible specification problems.

Some points should be noticed when it comes to initialization. When installing a PID controller in the model, the controller would sometimes initialize with a wrong value for the process variable (i. e. different from the actual value). If the controller was in “auto” mode when the integrator was started, this would lead to a kick in the manipulated variable in the moment the simulation started. This kick would cause deviations in the first minutes – this can be seen on most of the figures in section 5.2 – the process does not seem to be in completely steady state from the beginning.

Another point was that hold-ups in heat exchangers were not initialized properly – this may also be cause to some of the deviations occurring before any disturbances were introduced. In fact, these are more likely to be the cause of deviations as the controllers would correct wrongly initiated PV values in the first time step. If there is too little propane in the exchangers from the beginning, this will mean there is less heat transfer than necessary – leading to a raise in the exit temperature on the tube side. This will, of course, lead to increased flow of propane to the exchangers, and the temperatures will reach the set point. It seems, though, as the hold-ups start at zero, meaning the exchangers are empty at start.

6.2 Running the dynamic simulations, influence of control structure

For all the simulations, there was some deviation from steady state even before any disturbances were introduced. This could mainly be blamed on such details as controllers having other operation points defined initially than they should have, or unrealistic initial values for liquid hold-ups in the heat exchangers. In fact, when browsing the hold-up tabs of each exchanger before running the simulation, the hold-ups were all shown as zero, *including the liquid tank!* The liquid tank was initially set to have a level of 50 %, so there is in fact a contradiction between the specified level and the one displayed in the hold-up tab. In addition, it seems as the user is not allowed to change these values.

Some controllers also were initiated with a wrong value for the process variable (PV). An example is the pressure controller PIC-102 in scheme 2. However, in most cases the control system was able to stabilize the process variables at the desired set point within a short amount of time – before any disturbance was introduced.

Control scheme 1

With the recommended PID settings, the flow controller FIC-100 could not handle the 50% reduction of the set point – it gave an oscillating output signal. With a reduction in proportional gain, it tracked the set point nicely without oscillation, see figure 5.1b. For the temperature change, the controller seemed a bit too slow – figure 5.7 shows that the flow did not reach the set point before the temperature returned to the nominal value. However, it did never drift off dramatically – at most a few % off set point.

The MR compressors and intercoolers were hardly influenced at all by the downstream disturbances, so fluctuations here were mainly due to noise coming from inaccuracies in the solver. Small pressure changes would occur, but not large enough to cause significant deviations. See figures 5.5 and 5.6. The biggest deviations from set point in fact occurred *before* the disturbance was introduced.

The main control objective was, of course, to deliver the correct temperature on natural gas and MR to the main cryogenic heat exchanger. Figures 5.2a and 5.2b show that the set point was tracked satisfactory, in the first case, but with rather violent manipulations of the propane flow. In fact, the manipulated variable (the set point of the flow controller) was at both extremes (0 and 100%) during the simulation run.

For the case of disturbance in feed temperature, the process was not able to deliver the desired outlet temperature at any time before the feed temperature returned to its nominal value. As shown in figures 5.8a-d, the flow controllers all were at 100% for long periods. The explanation is simple – when the temperature is raised by 10°C, this means the first exchanger has to increase its duty with more than 50% as the natural gas is to be cooled by 28°C instead of 18°C. This requires an equally large increase in the propane flow, and this can not be delivered even at 100% valve opening. Therefore a temperature offset will result, and the next exchanger also has to increase its duty. When the equipment is not capable of delivering enough propane, the result will be a too high temperature on the natural gas from E-103.

The total cooling duty is determined by the number of moles of propane that is vaporized by natural gas and MR. When the flow out from V-100 is at its maximum, there is no possibility for further cooling. In addition, the compressor capacity will become a problem, because if the flow becomes too big, the compressors can not get the pressure to the desired level. Usually, the available compressor power is limiting how much LNG a plant can deliver.

Control Scheme 2

For most controllers/variables, the tendency was the same as with cascade control – set point tracking was fine for the case with a step in feed rate, but for the case with disturbance in feed temperature, the process did not manage to keep the natural gas to the MCHE at the desired temperature. One difference was very clear – there were oscillations on the temperature controllers, as shown on figures 5.14a-c. This indicates that the cascade controllers had a certain oscillation-damping effect, compare the figures with 5.8a-d.

One reason for the oscillatory behaviour can be the fact that the TIC controller gains were too large. The inner controllers would take the "edge" off the oscillation because of the integral action in these controllers. Without these inner loops, the TIC controller gains would have to be reduced in order to avoid oscillations.

6.3 Discussion of control structure

Keeping outlet pressures from the compressors constant seemed to work OK. As long as the overall flow of propane through the loop did not become too large, the pressure from the HP stage (K-101) was always close to set point. Then it would not be a problem that the rest of the pressures were further away – the pressure delivered to the condenser is most important.

One possible problem was that there were actually more valves than necessary between the liquid propane tank (V-100) and the heat exchangers. In fact, the level control valve (or the liquid level in the tank) could possibly be a bottleneck in the process – it would try to keep the level in the tank constant even when this would mean that not enough propane would go to the vaporizers. However, without any level control at all, the tank could run out of propane if not large enough, this would be a more serious problem than a period with too high temperatures. However, the process would probably have functioned better if the level control valve had been moved to the inlet side of the tank like shown in figure 6.1 (taken from [8]). Here, the valve in question is the upper one.

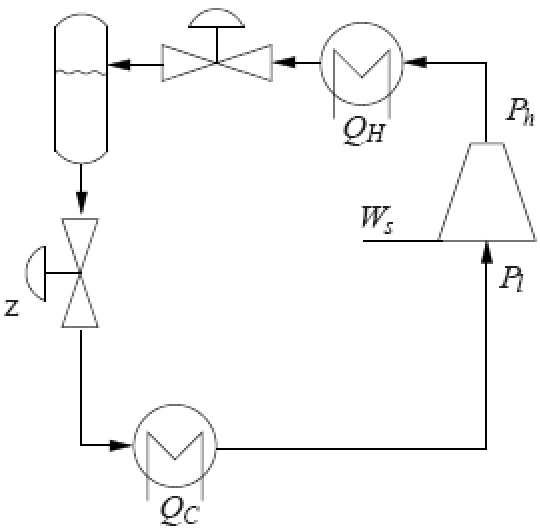

Figure 6.1: Typical positioning of level control valve and buffer tank in cooling cycle

As mentioned in 4.3, a detailed degree-of-freedom analysis was not carried out because of short time. Such an analysis would have shown that level control on the liquid tank would not be necessary if the rest of the levels were controlled. The total molar hold-up in the cycle must remain constant, so if the levels in the vaporizers and the condenser were kept constant, the level in V-100 would also remain constant.

The simulations showed that an *increase* in the necessary cooling duty would become problematic because of saturation of manipulated variables. The simple control structure used here was not capable of handling the 10°C increase in feed temperature – the natural gas temperature did not return to set point before the feed temperature returned to its nominal value. In order to obtain the desired temperatures all the time (if at all possible, given the compressor capacity), a more detailed control structure would be needed. Possible additions or changes could be:

- Varying the amount of natural gas feed in order to not exceed the capacity of the propane loop. For instance, the valves before each heat exchanger could be set to control propane pressure or liquid hold-up in exchangers rather than natural gas temperature, and the natural gas flow could be adjusted in order to maintain the correct temperature from the last exchanger. Of course, there is a problem with this setup as well – the fact that there are three exchangers between the flow valve and the stream that should be controlled, could lead to a large dead time. However, since the stream is gas, the flow velocity is rather fast and the dead time would be shorter than for a liquid stream.

- But: if the propane pressure at each level is kept constant, one must still take into account the propane flow through each exchanger. The compressor power is limited, and both decreased inlet pressure and increased throughput will increase the required power. If one is to maintain the nominal pressure on the high-pressure side, the total propane flow is limited by compressor capacity. If a compressor curve is used, there will typically also be a certain volumetric flow rate over which the compressor's efficiency drops significantly – this is the case for real compressors.

During all simulations, liquid was observed in the outlet streams from vaporizers from time to another. Since the propane outlet nozzles on the exchangers are placed on top (their elevation is set to 100%, meaning the top), this should not occur as long as a vapour phase is present. However, this seems to not be taken into account using the basic model – the result was that liquid was present. To avoid this, it would be necessary to control one more variable for each heat exchanger – for instance, the degree of superheating of the vaporized propane. In a real process, the liquid level in the vaporizers could also be controlled, because with a level below 100 %, only vapour would leave the exchanger. However, as HYSYS's basic model does not seem to take nozzle position into account, it would be of little use – the vapour stream would still have to be superheated.

A possible control layout that could both take care of outlet temperature on the tube side and avoid two-phase flow to compressors, could for instance be using the natural gas flow for temperature control, and controlling vapour fraction of the top stream by manipulating the propane flow to the exchanger.

A more advanced approach could include feed forward control on levels in the propane vaporizers (and V-100 if the controller was not omitted). The required cooling duties for MR and natural gas determine how much propane is vaporizing. Therefore, measuring the temperature of the natural gas feed could be used to determine the amount of propane that should be fed to each exchanger for it in order to maintain the necessary liquid level. This would require use of the detailed heat exchanger model in HYSYS, because the simple model does not take liquid level into account when calculating heat exchanger duty.

7. Conclusions

7.1 About model building in HYSYS

- Building models in Aspen HYSYS, both steady-state and dynamic, is for most cases a rather intuitive process, and the program is flexible when it comes to complexity. The detail level in a model can range from very simple to very detailed.

- A main flaw in the software is that heat exchanger models do not accept multiple shell exchangers, meaning that really large exchangers can not be modelled using the detailed rating model. They must instead be modelled by specifying volumes and UA, as well as k-values (pressure drop-flow relation) – the simplest possible approach.

- Improvements that could be made, include the possibility of simulating heat exchangers with more than one outlet on the shell side (t. e. kettle boilers with both liquid and vapour outlets), and a heat exchanger model that takes into account the arrangement of tubes inside the shell (so that heat transfer in a vaporizer depends on the number of submerged tubes).

All in all, the biggest problem in the project was to model the heat exchangers as realistic as wanted.

7.2 About running dynamic simulations

The tools contained in HYSYS can give the user good opportunities to study the dynamic behaviour of a process model in detail, and the interface is generally user-friendly. The Event Scheduler proved useful for introducing different disturbances in the process.

The basic heat exchanger model in HYSYS did not work well for dynamic simulations with respect to realism. For further studies of this process, one should scale down the process to a scale where the detailed model can be used.

7.3 About control structures used and control guidelines provided in the HYSYS documentation

For pressure controllers, the recommended settings seemed to function well, the same for flow controllers when used as slaves in cascades. The temperature controller settings recommended seemed a bit too aggressive. The same was observed for the flow controller on the feed stream. However, for another process the suggested tuning parameters may work well – a detailed study of the process would be required to find the best controller settings.

The use of cascade control on the propane vaporizers seemed to have a general stabilizing effect, but did not contribute to significantly better set point tracking.

7.4 Possible topics for further studies

Using this project as a starting point, there are several possible topics that may be interesting in the field of process simulation, control and optimization.

- A steady-state optimization of this process (or a similar process), either in HYSYS or in another program, maybe also with focus on finding good variables for self-optimizing control. Such a study would have to take into account the entire process, not only the pre cooling section, in order to be really useful
- A more detailed study of control of the process, using feed forward control or model predictive control (MPC) and finding proper tuning parameters
- Comparing different PID tuning rules (like Ziegler-Nichols or SIMC) for this process to see which rules give the best performance

Trondheim, December 1. 2006

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Literature

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[5] HYSYS 2004.2 Operations Guide. AspenTech, 2005 (www.aspentech.com)
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[8] Jensen, J. B., Skogestad, S.: Optimal operation of simple cooling cycles
[9] Seborg, Edgar, Mellichamp: Process Dynamics and Control, 2. ed., John Wiley & sons, Inc. 2004
[10] Skogestad, S.: Simple analytic rules for model reduction and PID controller tuning, Journal of Process Control (13) 2003

Attached files

On the CD delivered with the original report, the following files are included:

HYSYS files (.hsc):

C3-MR Process (the process model built at Norsk Hydro)
Dyn init – control scheme 1 (initial file for the two first simulations)
Dyn init – control scheme 2 (initial file for the two last simulations)
TEST control scheme 2
RESULTS 1 – control scheme 1 (disturbance in feed flow)
RESULTS 2 – control scheme 1 (disturbance in feed temperature)
RESULTS 1 – control scheme 2 (disturbance in feed flow)
RESULTS 2 – control scheme 2 (disturbance in feed temperature)

APPENDIX A: Steady state stream data

Table A.1: Stream data workbook from HYSYS – initial values for dynamic simulations

Name	1	33	2	8	18
Vapour Fraction	1.0000	1.0000	0.0000	0.0041	0.0041
Temperature (C)	27.09	12.01	10.00	9.442	9.442
Pressure (kPa)	645.7	467.9	635.7	625.7	625.7
Molar Flow (kgmole/h)	8.835e+004	8.835e+004	8.835e+004	3683	6.935e+004
Mass Flow (kg/h)	3.896e+006	3.896e+006	3.896e+006	1.624e+005	3.058e+006
Liquid Volume Flow (m3/h)	7689	7689	7689	320.5	6036
Heat Flow (kJ/h)	-9.233e+009	-9.316e+009	-1.076e+010	-4.487e+008	-8.450e+009
Name	31	14	30	29	34
Vapour Fraction	1.0000	0.9999	1.0000	1.0000	1.0000
Temperature (C)	-14.64	-20.13	-7.591	-38.96	21.00 *
Pressure (kPa)	243.1	243.1	243.1	116.5	4900 *
Molar Flow (kgmole/h)	7.219e+004	6702	3.133e+004	3.133e+004	6.000e+004
Mass Flow (kg/h)	3.183e+006	2.955e+005	1.382e+006	1.382e+006	1.277e+006
Liquid Volume Flow (m3/h)	6282	583.3	2727	2727	3665
Heat Flow (kJ/h)	-7.728e+009	-7.200e+008	-3.339e+009	-3.399e+009	-4.661e+009
Name	36	9	10	11	13
Vapour Fraction	1.0000	0.0667	1.0000	0.0041	0.1806
Temperature (C)	3.000	0.3000	-0.3727	9.442	-19.00
Pressure (kPa)	4840	477.9	467.9	625.7	253.1
Molar Flow (kgmole/h)	6.000e+004	3683	3683	1.532e+004	6702
Mass Flow (kg/h)	1.277e+006	1.624e+005	1.624e+005	6.754e+005	2.955e+005
Liquid Volume Flow (m3/h)	3665	320.5	320.5	1333	583.3
Heat Flow (kJ/h)	-4.718e+009	-4.487e+008	-3.918e+008	-1.866e+009	-8.166e+008
Name	37	15	16	17	38
Vapour Fraction	0.9407	0.0041	0.2709	1.0000	0.7877
Temperature (C)	-18.00	9.442	-37.00	-38.96	-34.00
Pressure (kPa)	4790	625.7	126.5	116.5	4740 *
Molar Flow (kgmole/h)	6.000e+004	8614	8614	8614	6.000e+004
Mass Flow (kg/h)	1.277e+006	3.799e+005	3.799e+005	3.799e+005	1.277e+006
Liquid Volume Flow (m3/h)	3665	749.7	749.7	749.7	3665
Heat Flow (kJ/h)	-4.815e+009	-1.050e+009	-1.050e+009	-9.344e+008	-4.930e+009
Name	28	19	46	49	24
Vapour Fraction	1.0000	0.0041	1.0000	0.2565	0.1806
Temperature (C)	-38.96	9.442	21.00	-34.00	-19.00
Pressure (kPa)	116.5	625.7	4600	4450 *	253.1
Molar Flow (kgmole/h)	2.272e+004	1.248e+004	1.198e+005	1.198e+005	3.415e+004
Mass Flow (kg/h)	1.002e+006	5.503e+005	2.977e+006	2.977e+006	1.506e+006
Liquid Volume Flow (m3/h)	1977	1086	8366	8366	2972
Heat Flow (kJ/h)	-2.464e+009	-1.521e+009	-9.892e+009	-1.088e+010	-4.161e+009
Name	27	21	25	22	26
Vapour Fraction	0.2709	1.0000	1.0000	0.0041	0.0041
Temperature (C)	-37.00	0.4383	-20.13	9.442	9.442
Pressure (kPa)	126.5	467.9	243.1	625.7	625.7
Molar Flow (kgmole/h)	2.272e+004	1.248e+004	3.415e+004	5.687e+004	2.272e+004
Mass Flow (kg/h)	1.002e+006	5.503e+005	1.506e+006	2.508e+006	1.002e+006
Liquid Volume Flow (m3/h)	1977	1086	2972	4949	1977
Heat Flow (kJ/h)	-2.768e+009	-1.327e+009	-3.669e+009	-6.929e+009	-2.768e+009
Name	45	44	42	41	39
Vapour Fraction	1.0000	1.0000	1.0000	1.0000	1.0000
Temperature (C)	80.10	38.00	10.00	21.25	-30.00 *
Pressure (kPa)	4610	2690	990.0	1000	450.0 *
Molar Flow (kgmole/h)	1.198e+005	1.198e+005	1.198e+005	1.198e+005	1.198e+005
Mass Flow (kg/h)	2.977e+006	2.977e+006	2.977e+006	2.977e+006	2.977e+006
Liquid Volume Flow (m3/h)	8366	8366	8366	8366	8366
Heat Flow (kJ/h)	-9.456e+009	-9.658e+009	-9.739e+009	-9.675e+009	-9.926e+009

Table A continued

Name		7	12	20	23	48
Vapour Fraction		0.0041	0.0041	0.0667	0.0041	0.5073
Temperature	(C)	9.442	9.442	0.3000	9.442	-18.00
Pressure	(kPa)	625.7	625.7	477.9	625.7	4500
Molar Flow	(kgmole/h)	1.900e+004	6702	1.248e+004	3.415e+004	1.198e+005
Mass Flow	(kg/h)	8.378e+005	2.955e+005	5.503e+005	1.506e+006	2.977e+006
Liquid Volume Flow	(m3/h)	1654	583.3	1086	2972	8366
Heat Flow	(kJ/h)	-2.315e+009	-8.166e+008	-1.521e+009	-4.161e+009	-1.058e+010
Name		47	43	35	32	51
Vapour Fraction		0.9400	1.0000	1.0000	1.0000	0.0000
Temperature	(C)	3.000	81.03	20.94	14.61	5.000
Pressure	(kPa)	4550	2700	4890	467.9	101.5
Molar Flow	(kgmole/h)	1.198e+005	1.198e+005	6.000e+004	7.219e+004	4.921e+006
Mass Flow	(kg/h)	2.977e+006	2.977e+006	1.277e+006	3.183e+006	8.865e+007
Liquid Volume Flow	(m3/h)	8366	8366	3665	6282	8.883e+004
Heat Flow	(kJ/h)	-1.009e+010	-9.376e+009	-4.661e+009	-7.597e+009	-1.412e+012
Name		52	40	50	6	55
Vapour Fraction		0.0000	1.0000	0.0000	0.0041	0.0000
Temperature	(C)	9.000	-30.13	4.998 *	9.442	18.00
Pressure	(kPa)	96.46 *	440.0	111.5 *	625.7	96.32 *
Molar Flow	(kgmole/h)	4.921e+006	1.198e+005	4.921e+006	8.835e+004	4.311e+005
Mass Flow	(kg/h)	8.865e+007	2.977e+006	8.865e+007	3.896e+006	7.766e+006
Liquid Volume Flow	(m3/h)	8.883e+004	8366	8.883e+004	7689	7782
Heat Flow	(kJ/h)	-1.411e+012	-9.926e+009	-1.412e+012	-1.076e+010	-1.233e+011
Name		54	58	57	60	61
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	0.0000
Temperature	(C)	5.000	25.00	5.000	5.000	9.000
Pressure	(kPa)	101.3	96.32 *	101.3	101.3	96.32 *
Molar Flow	(kgmole/h)	4.311e+005	1.814e+005	1.814e+005	2.068e+005	2.068e+005
Mass Flow	(kg/h)	7.766e+006	3.269e+006	3.269e+006	3.726e+006	3.726e+006
Liquid Volume Flow	(m3/h)	7782	3275	3275	3733	3733
Heat Flow	(kJ/h)	-1.237e+011	-5.179e+010	-5.207e+010	-5.934e+010	-5.928e+010
Name		3	53	59	56	
Vapour Fraction		0.0000	0.0000	0.0000	0.0000	
Temperature	(C)	10.00	4.995 *	4.995 *	4.995 *	
Pressure	(kPa)	635.7	126.3 *	126.3 *	126.3 *	
Molar Flow	(kgmole/h)	8.835e+004	4.311e+005	2.068e+005	1.814e+005	
Mass Flow	(kg/h)	3.896e+006	7.766e+006	3.726e+006	3.269e+006	
Liquid Volume Flow	(m3/h)	7689	7782	3733	3275	
Heat Flow	(kJ/h)	-1.076e+010	-1.237e+011	-5.934e+010	-5.207e+010	

All stream names (numbers) refer to the HYSYS flow sheets. A flow sheet with stream numbers included is shown in figure B.3..

APPENDIX B: Flow sheets

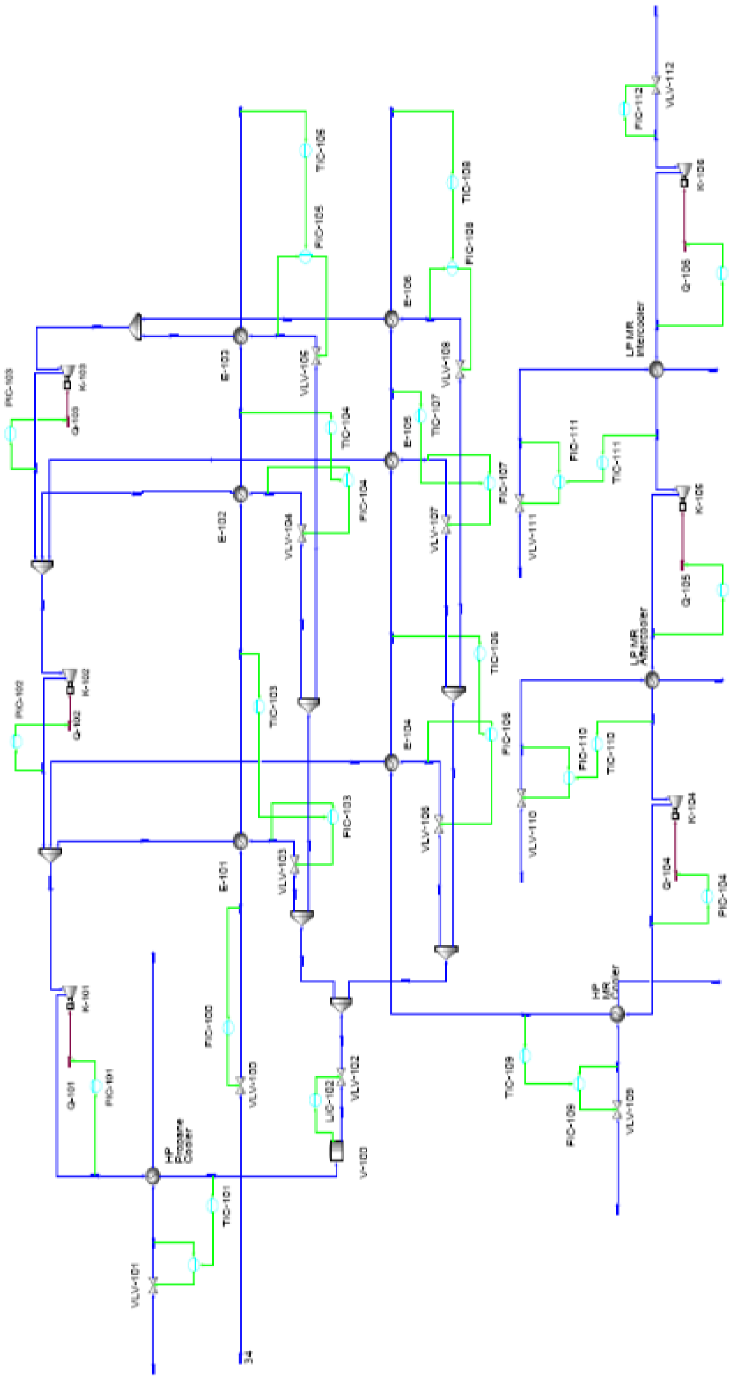

Figure B.1 Flow sheet of process with control scheme 1

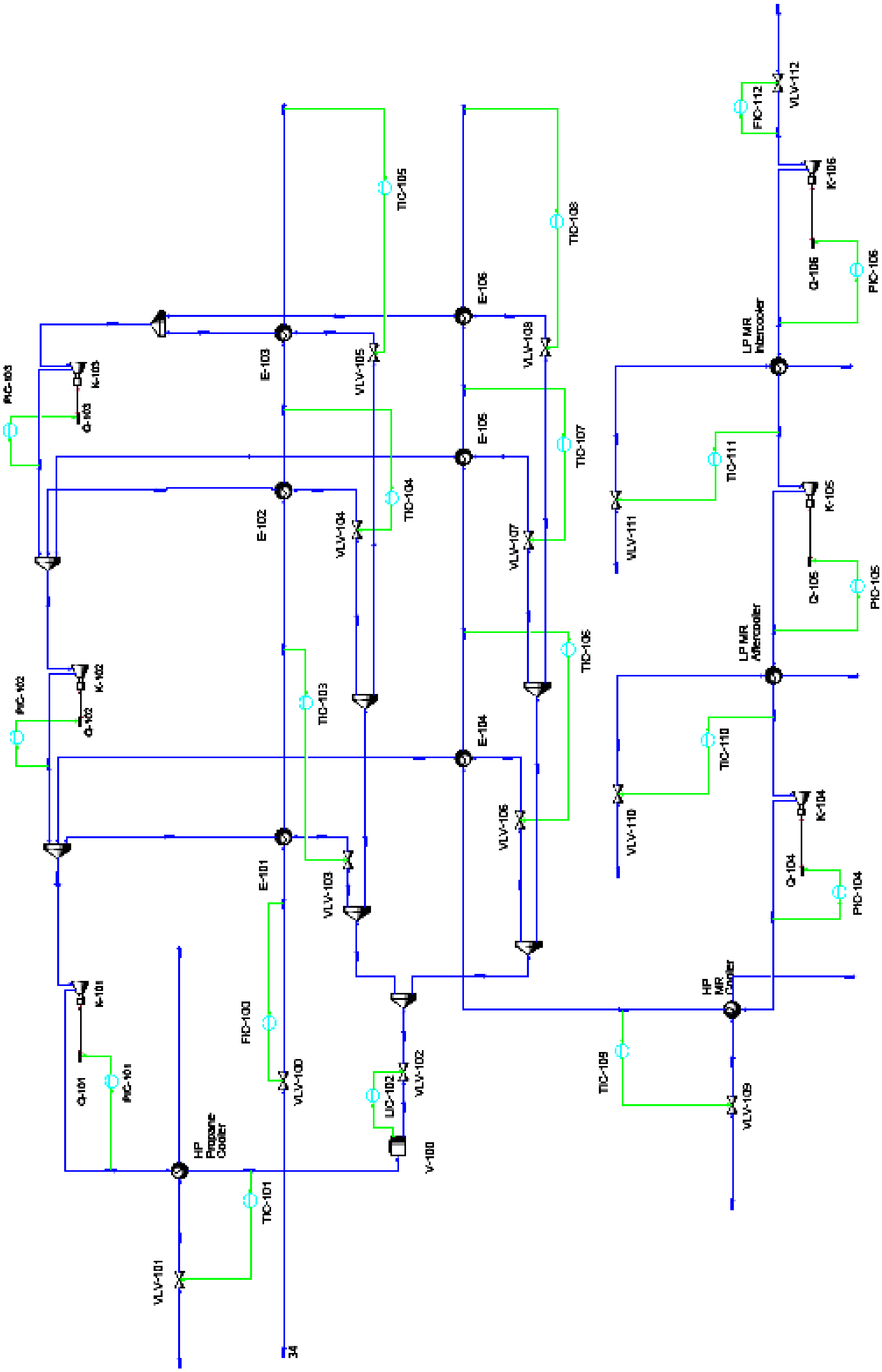

Figure B.2 Flow sheet of process with control scheme 2

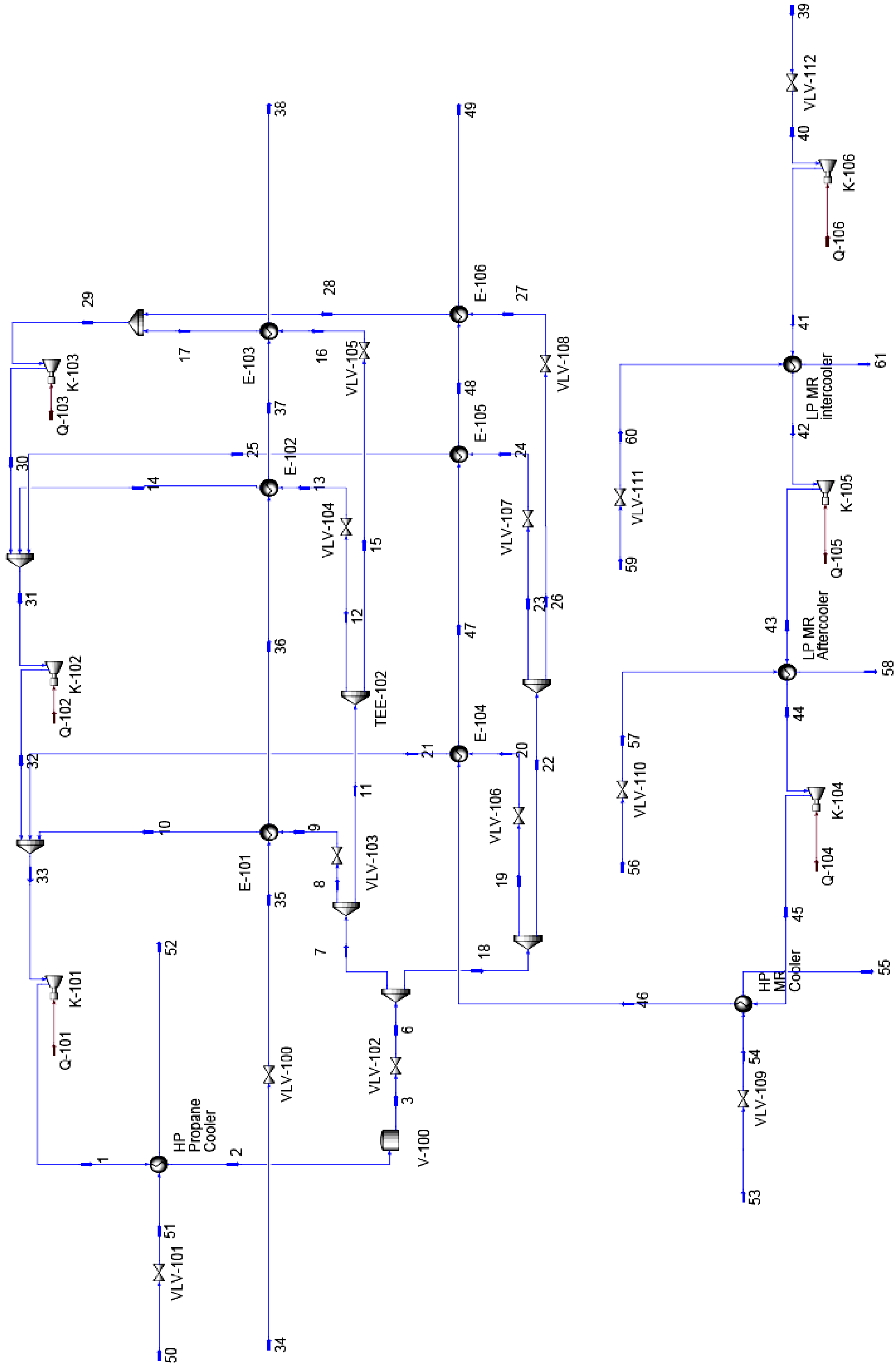

Figure B.3 Flow sheet with all stream numbers, and with controllers not shown. (For better picture quality one should see the HYSYS files)