CONTROL LOOPS

Practical exercises to provide an understanding of the differences between the control systems and the parameters that regulate them and to see how the latter affect their operation.
INTRODUCTION

In this day and age it has become necessary to seek maximum efficiency and speed in executing any process, obtaining the desired results with a minimum investment and the highest quality. For this reason, control systems have become very important, and considerable time and resources have been dedicated to studying them to ensure their efficient start-up and operation. Thus, for a process to function within the required margins, an adequate control system must be implemented, regulated in accordance with the parameters that govern it. Within this context, a distinction is established between different types of controllers:

- Proportional (P) controllers.
- Proportional-integral (PI) controllers.
- Proportional-derivative (PD) controllers.
- Proportional-integral-derivative (PID) controllers.

The function of a controller is to detect changes in a certain process variable (PV) through a sensor. Since they have been assigned a set point (SP) and a margin of variation in the variable, they can act on a specific variable that must be changed (OP), in order to reach the set point and keep the system stable. As mentioned above, depending on the type of process and the variables to be controlled, this will be accomplished more efficiently depending on the type of controller installed.

A proportional controller is entrusted with changing the OP in proportion to the error detected, understood as the distance (difference) between the set point and the value of the variable measured. Thus, its fundamental parameter is the $K_c$ or gain. The higher the gain, the greater the change occurring in the manipulated variable (OP) to enable it to correct the error as quickly as possible. However, this may lead to undesirable oscillations around the SP when the errors are small, due to the extremely high gain giving rise to very sudden changes that eventually make the control loop unstable.

To improve this, a PI controller can be used, in which an integral time (Ti) is entered in order to approach the errors in small increments, thus improving the accuracy of the controller and obtaining a zero error. It should be considered that the longer the Ti, the less the integral control exerted, as the integral control will take longer to occur, whereas if it is short, the control will be clearly influenced by the integral action (it will occur within a shorter time). However, as mentioned above, an integral control that is too fast (lower Ti) may cause oscillations, delaying the arrival at the set point. This occurs because if the integral control is exerted too quickly, it will constantly "exceed" the set point, oscillating around it.

Within this context, PD control, in which the derivative time is added to the control (Td), makes it possible to make the approach in each control step in accordance with the error trend (whether below or above the SP), in the direction of the error and anticipate it; this means that it is faster in reaching the minimum error effectively right from the start. In turn, the PID attempts to include the benefits of both controls, in order to obtain a zero error more quickly, but it is not always the most appropriate, as derivative control is often affected by constantly-changing processes or by signal noise, which prevents it from "predicting" the error trend correctly. As always, everything comes at a price and has its advantages and disadvantages, so it is important to identify the parameters that are most critical for controlling the process and use them to exert better control over it.
CONTROL LOOPS

OBJECTIVES

The main objectives of this set of exercises is to study the stability of the controllers by changing their parameters and verifying that the control system is able to reach the set point and seeing how each parameter is affected and what margins or control strategies are the most critical in ensuring the correct control of a specific variable.

CASE STUDY DESCRIPTION

To open the Control loop case, in the Cases menu, go to the Open cases, and in it, go to Control Loops.

Figure 1. Base case started up and showing the process data generated by simulation

The Control loop case consists of a single gas-liquid separator in which some of the working conditions can be changed (temperature, pressure or Cv -size- of the valves) and the main controller parameters already described, in which it is also possible to change its action (direct or reverse) and its operating mode (off, manual or automatic), as shown in Figure 2. A direct operation means that in the presence of an increase in the PV above the SP, the controller must also apply an increase in the OP. On the contrary, in a reverse action, the OP must be reduced in the presence of an increase in the PV with respect to the SP.

We should bear in mind that in the event of requiring a more proportional and less integral control, we will have to considerably increase the value of Ti in the table, to give priority to the Kc value. On the other hand, no derivative control will be used, given that, as a rule, this control is not often used in
industry, and therefore $T_d = 0$; but if necessary, it can easily be changed in the same table of each controller.

It is seen that a cascade control exists, formed by the level controller LIC-001, which uses the flow rate controller FIC-002 to maintain its set point. So only the mode and action of the master controller (LIC-001) can be changed, and only the internal parameters of its slave (FIC-002) can be changed. It is thus possible to regulate the liquid outlet and the tank level.

![Figure 2. Modifiable parameters of controller FIC-001](image)

The pressure controller PIC-001 is used to regulate the gas outlet and it will thus be affected by the pressure and temperature in the tank, which, in turn, will depend on temperature controller TIC-001, which regulates the heat inlet expressed as power. The valve percentage is not its aperture angle, but the heating power expressed as a percentage of the maximum available power being used.

As seen, there are four independent control systems that are, however, associated with each other, as the perturbations affecting the set point of one of them will affect the process and hence, the other controllers. This can be seen in the graphs and in the tables set out next to each controller with their most relevant parameters.
GRAPHICS

They allow us to observe the behaviour of the different controllers: the OP of each one is shown, so that we can see how each one affects its manipulated variable, in order to adapt to changes in the process, which are shown through the SP and PV.

The behaviour of each controller is shown below in the presence of a change in the tank level set point. It is clearly seen that the cascade controllers, whose function is to regulate the level, react with strong oscillations in the time to reach the SP, although the oscillations indicate that they find it difficult to quickly reach a stable error, and the oscillations will decrease until they reach a zero error.

Figure 3. Graph showing the OP of each controller

Figure 4. Graph showing the SP of each controller
As expected, we can see that the speed with which the SP changes and how the PV adapts to each, always depending on the action of the OP. Thus the correlation among the 3 parameters is clear, and will help us see the set points that facilitate process controllability, and on changing the control parameters, how they will affect the manipulated and controlled variables of the process. One significant example of this is cascade control, in which it is seen that the SP changes in accordance with the variations required by the master controller OP.

![Figure 5. Graph showing the PV of each controller](image)

Figure 5 also shows how process variables such as pressure or temperature are affected by the actions of the controller. In this way, if the set point of the temperature controller is changed, there is an immediate response in the evolution of those parameters and it also affects the outgoing flow rates, as seen in Figures 6 and 7 below. It should be noted that the flow rates perceive the change earlier and to a greater extent than the temperature sensor. Therefore this gives us an idea of how quickly the variables can change, and which controls must be faster.

This situation arises because the temperature has risen to 100 °C and, since all the liquids are organic, the increase in temperature makes the volatile component flow rate rapidly increase, with all the gas having a temperature of 100 °C. This leads to a drastic fall in the liquid flow rate and a considerable increase in the gas flow rate. Would the facility be prepared to deal with this situation, the outcome of which would be emptying the tank? How could this problem be solved? This is where the present exercises are useful.

It is thus of the essence to understand what is being done, in order to foresee how the process will react to a change in the controller and vice versa, in order to adapt the action of the controller in the best way to ensure that the process being executed stabilises within the correct margins.
Figure 6. Graph showing the temperature and pressure variables inside the tank

Figure 7. Graph showing the outgoing liquid and gas flow rates
SCENARIO 1. STEADY STATE

Start up the case by pressing the green **Run** button under the **Runtime Module** menu and open the three graphs located next to this button. Then wait for the graphs to show the process variables in the steady state.

In this base case, the parameters of each controller (gain (Kc), integration time (Ti) and derivation time (Td)) correspond to the values that produce the correct functioning of the control system. To do the exercises correctly, it should be known that the integral control eliminates the steady state error, but this may give rise to important oscillations, with the result being that the systems takes longer to stabilise. In turn, derivative control does not, in fact, have a significant relation to the error in the steady state and it is only used to give a better initial response.

Within this context, integral control can be understood as the memory of a past error. For this reason, if, after an action, the error continues to be present, the integral control reacts by increasing the action, meaning that if it is close to the set point and with a high gain, the oscillations will be greater, as the error remains and the integral control continues to operate until it manages to reduce it.

The following exercises are proposed for the purpose of verifying the correct operation of the controllers. Take note of the final values (steady state) of the foregoing variables if you see that they do not change in Table 1.1 below.
Q1.1: If the SP of FIC-001 is changed to 11,000 kg/h, how long will it take for the process to become stable again?

Q1.2: If the SP of LIC-001 is changed to 25%, will there be an over impulse in any of the controllers? Is this important?

At the beginning, it was seen that on increasing the temperature, the system eventually becomes unstable due to the emptying of the tank. Imagine that the process conditions require a higher temperature (90ºC) to make the global process more economical.

Q1.3: What would you do to keep the system stable with the initial filling of 50%? What fundamental parameter must be controlled, and how?
### Table 1.1. Results of changing the SP of TIC-001

<table>
<thead>
<tr>
<th></th>
<th>Stabilisation time</th>
<th>Value of the highest over impulse</th>
<th>Existence of error in the steady state</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Base case</strong></td>
<td>3 minutes</td>
<td>None</td>
<td>No</td>
</tr>
</tbody>
</table>

### Table 1.2. Results of changing the SP of LIC-001

<table>
<thead>
<tr>
<th></th>
<th>Stabilisation time</th>
<th>Value of the highest over impulse</th>
<th>Existence of error in the steady state</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Base case</strong></td>
<td>8 minutes</td>
<td>55%</td>
<td>No</td>
</tr>
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</table>
SCENARIO 2. CHANGE IN THE GAIN OF THE $K_c$ CONTROLLER

The proportional control is the simplest, and therefore it will be very useful in controlling processes and in particular, when using the other parameters, to achieve the stabilisation times and the necessary minimum of variability for the process. Changing the gain of the controller is a very useful tool to ensure its correct operation.

Re-open the base case with its initial values. Press Run like before and wait for the values to stabilise. Open all the graphs to see how each part of the system is affected.

Q2.1: What happens when the gain of controller FIC-001 has a value of 10 on changing the SP to 9,000 kg/h?

Is it possible to start to control a process with such a high gain and then moderate it? In other words, sometimes, when making certain changes, we might consider an important action at the start that leads to a considerable increase in the variable of interest, such as the temperature, in order to reach the desired value, so that the controller gain is initially increased and then reduced.

Q2.2: Describe what happens when, in the base case, we start with $K_c = 8$ in TIC-001 and then the temperature is changed to 5ºC. Are there any over impulses? Is there any improvement when it is restored to its value of 1?

In Q2.1 the change in gain was certainly very brusque. However, some variables or controllers are more affected by more minor changes.
It is therefore clear that, depending on the controller to which the strategy is applied or different gain values, different results may be obtained that could be more or less desirable. There will be process points in which an excessive gain may be lethal to it, and others in which it may help at certain times, in all cases returning to the stable point. Thus, it is essential to select values that are coherent and realistic for the process, as otherwise we could make it worse.

So it is advisable to start designing control systems with small values and gradually see how each variable is affected and what other parameters can be combined to ensure their responsible is adequate in terms of magnitude and time. However, a process may also be inefficient when it fails to act with sufficient force.

This demonstrates the importance of knowing which controllers and variables are affected, and which are the primary ones in terms of control. Proportional control will always act by reducing error. As a general rule, controllers with the lowest possible number of parameters must be used, as every time one is added, this increases the complexity in stabilising control, and a control that is excessively adjusted is very specific in the process in question, which is less general. The integral control recommended in cases with small sequential changes is set out below, as it is better in delimiting errors and also in processes dominated by idle time (the time a control takes to respond to a change in a system variable).
SCENARIO 3. CHANGES IN THE CONTROLLER INTEGRAL TIME, Ti

In this scenario we study the function of the integral parameter of the controller. Bear in mind that integral control gives a zero error, and therefore greater the precision, but at the cost of having to invest more time.

To start with, change the set points again to return to the base case. Now change the flow rate to 8,000 kg/h at the inlet.

**Q3.1**: If an integral time of 150 minutes and Kc of 1 are entered in FIC-001 (changes to proportional control), do you think the stabilisation time will be shorter or longer than the one observed above? What happens?

**Q3.2**: If the process makes it necessary to delimit the error and the initial Ti is required (0.25 minutes) but we want to make it faster, what must be done? Why?

When we re-open the case with its default values, we see that there are controllers with very little integral control, such as the pressure controller. So we might think that it would be a good idea to make this parameter more accurate, given that we are separating gas from liquid due to the different volatility of the mixture. We will now study the behaviour. Start up the integrator and increase the SP of the controller by 0.2 bar.
Q3.3: Describe what you see. What happens when the integral time is reduced to 0.1 min? What is the explanation?

It was seen that it is wrong to think that simply increasing the integral control (reducing Ti) would be sufficient to improve a result, as it could exceed the grade of accuracy. We will try to correct the system again.

Q3.4: How could the system be stabilised and what parameters would you change to reduce the oscillations still observed with the default values?

Return to the base case for the next test. Locate the least accurate variable, or the one with the response that is most difficult to stabilise, and which thus requires greater integral control. Which of the variables can be changed most accurately?

Q3.5: Identify a variable whose Ti can be changed to improve it and say which values were obtained and what the best one was. Are there any problems? Describe them.

To find out the response to question Q3.5 try a Ti of up to 1.5 or a minimum of 0.002 to see its evolution, and find the optimal value, based on what is required. The oscillations in the variables over time are mitigated, and everything depends on the time available to reach a sufficiently steady state from an approximate perspective.
SCENARIO 4. CHANGES IN THE DERIVATIVE TIME OF THE CONTROLLER, Td

We will now study the best situation in which to use derivative control and why, and learn how to change it in order to suit our purposes.

As indicated above, start with the base case and restore all the variables to the values of the base case. We have seen that variables such as the liquid flow rate of products show great oscillations and are at the mercy of the control, and it is difficult for them to reach a stable error. So it might be normal to think about improving their action in order to anticipate the error and have fewer oscillations. Enter a change of 1,000 kg/h in the incoming flow rate.

Q4.1: Is it helpful to use a derivative time of 0.2 minutes in controller FIC-002? What do you expect to see?

Return to the initial values to respond to Q4.2. With the same flow rate:

Q4.2: What do you think is the variable that is most in need of derivative control? Justify this and describe what happens when a Td = 20 minutes is entered. Does it make sense?

Observe what happens when the Td value is restored to 0. Bear in mind that according to what was said up to now, the Td could be used to stabilise the system faster, so if this is done coherently, the whole process stands to gain, for, as already explained, a change in one controller has a repercussion on the others. Return to the base case.
Q4.3: Set the temperature controller to $T_d = 0.2$ minutes (changing the pressure to 5.4 bar). What happens? How long does the system take to become stable?

This could lead us to think that a significant increase in derivative control would give a more favourable result, but is this actually true?

Q4.4: What happens when the $T_d$ of TIC-001 is changed to 2 minutes?

Again, this shows the importance of knowing the process and changing the control parameters in a reasonable manner to achieve the objectives, as otherwise, an override could generate a negative effect. Derivative control is used less often and is usually the last element to be adjusted, since with a PI controller, it is normally sufficient. Derivative control can be used as support in refining the controller response grade under the specific process conditions, if its vital aspects and the most interesting ranges of action are known.
SCENARIO 5. CHANGES IN THE PROPERTIES OF THE VALVES, Cv

In this scenario, we propose to show that in implementing the control, not only must the mode and control action be considered, the elements of the process that intervene must also be taken into account. In this regard, it is necessary to define a fundamental element in any valve: Cv.

Cv is a valve design parameter that is expressed in USGPM (*United States Gallons Per Minute*) units and thus, by definition, it is the water flow in gallons per minute that flows through a completely open valve in 60 °F conditions with a pressure loss of 1 psia. Thus, it is a measurement of how large the valve is, which is essential in the control, as a change in its controller could have a greater or a smaller effect, depending on the size of the valve.

As already done above, start with the base case by restoring the variables to their original values, as shown in Figure 8.

<table>
<thead>
<tr>
<th>Q5.1: What happens when we enter the value $C_v = 500$ USGPM in valve PCV-001? Why?</th>
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<table>
<thead>
<tr>
<th>Q5.2: What parameters of controller PIC-001 would you change to obtain a better stabilisation value, generating a lower over impulse? Explain how you did it.</th>
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It is also possible to test how a change in the valve would affect the system. It is quite usual (especially in the simulation) to start with a valve aperture of 50%. In the first valve of the process, this is not the case.
Q5.3: What Cv value must be entered for valve FCV-001 in order to comply with what was mentioned above?

With approximately 100 USGPM the desired aperture angle is obtained.

This can also be done with other valves, such as the one next to the pressure controller PCV-001. We will now see how the system acts in the presence of changes of this type. Make the changes proposed above or any others you wish to make and observe the controller parameter graphs.

Q5.4: How do the OP of the controllers behave when the Cv is changed?

What happens when a valve is undersized? As inferred from the above, this will make it necessary to operate with a large aperture, and in addition, it will provoke sudden jumps to reach the SP, depending on the type of control. See whether this occurs:

Q5.5: How does the liquid flow rate branch react to Cv = 10 USGPM in LCV-001? What it if were 1 USGPM?

The best option would therefore be to find the best Cv values to allow the system to be managed with no problems under normal conditions, with a margin for preventing accidents, unexpected events or errors that might occur. All the above must be done in accordance with the sizing conditions already established in the plant, the equipment availability or the conditions imposed by the equipment.

In many cases, it should be considered that the system control is subsequent to the design in the plant, or that it is often implemented in already existing plants, and so it is common for the control system to be adapted to the type of valve and not vice versa, but if starting from scratch, it is always advisable to study both viewpoints in order to select the most appropriate general design.
In conclusion, apart from what we have learned about controllers, their parameters and their correct operation, it should be considered that the failure of one part will often inexorably lead to the failure of the system, for instance, the tank overfilling due to a failure in the liquid outlet valve. The simulation tools allow these contingencies to be detected and corrected, and an estimation of the time they could take to cause a disastrous outcome. For this reason, in control systems it is advisable to implement an integral, seamless control with several levels of protection and prevention. This will result in:

- Duplicate or triplicate vital controls.
- Remote and manual control by plant operators.
- High level alarms and very high level alarms for any main parameter that is being controlled.
- Sequential control loops: if an outlet valve fails, the inlet valve must close or an alternative outlet be enabled, etc.
- Emergency plan, depending on the alarm level.
- Prior controls performed by simulation and planning for emergency situations.
- Training for operators.