Myke King

Process Control

A Practical Approach

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Myke King

Whitehouse Consulting, Isle of Wight, UK

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Preface

So why write yet another book on process control? There are already many published, but they are largely written by academics and intended mainly to support courses taught at universities. Excellent as some of these books are in meeting that aim, the content of many academic courses has only limited relevance to control design in the process industry. There are a few books that take a more practical approach but these usually provide only an introduction to the technologies. They contain enough detail if used as part of a wider engineering course but not enough for the practitioner. This book aims more to meet the needs of industry.

Most engineers responsible for the design and maintenance of control applications find daunting much of the theoretical mathematics that is common in the academic world. In this book we have aimed to keep the mathematics to a minimum. For example, Laplace transforms are only included so that the reader may relate what is in this book to what will be found in most theoretical texts and in the documentation provided by many DCS (distributed control system) vendors. They are not used in any of the control design techniques. And while we present the mathematical derivation of these techniques, to show that they have a sound engineering basis, the reader can skip these if too daunting and simply apply the end result.

The book aims to present techniques that have an immediate practical application. In addition to the design methods it describes any shortcuts that can be taken and how to avoid common pitfalls. The methods have been applied on many processes on a wide range of controllers. They should work!

In addition to providing effective design methods, this book should improve the working practices of many control engineers. For example, the majority still prefer to tune PID (proportional, integral, derivative) controllers by trial and error. This is time-consuming and rarely leads to controllers performing as well as they should. This might be because of a justified mistrust of published tuning methods. Most do have serious limitations. This book addresses this and offers a method proven to be effective in terms of both controller performance and engineering effort.

DCS include a wide array of control algorithms with many additional engineer-definable parameters. The DCS vendors are poor at explaining the purpose of these algorithms with the result that the industry is rife with misinterpretation of their advantages and disadvantages. These algorithms were included in the original system specification by engineers who knew their value, but this knowledge has not passed to the industry. The result is that there are substantial improvements that can be made on almost every process unit, surpassing what the control engineer is even aware of – let alone knows how to implement. This book addresses all the common enhancements.

This book takes a back-to-basics approach. The use of MVC (multivariable controllers) is widespread in industry. Control engineering staff and their contractors have invested thousands of man-hours in the necessary plant testing and commissioning. Improving the basic controls is not usually an option once the MVC is in place. Improvements are likely to change the process dynamics and would thus involve substantial re-engineering of the MVC. Thus poor basic control remains the status quo and becomes the accepted standard to the point where it is not addressed even when the opportunity presents itself. This book raises the standard of what might be expected from the performance of basic controls.

Before MVC, ARC (advanced regulatory control) was commonplace. MVC has rightly replaced many of the more complex ARC techniques, but it has been used by too many as the panacea to any control problem. There remain many applications where ARC outperforms MVC; but appreciation of its advantages is now hard to find in industry. The expertise to apply it is even rarer. This book aims to get the engineer to reconsider where ARC should be applied and to help develop the necessary implementation skills.

However due credit must be given to MVC as a major step forward in the development of APC (advanced process control) techniques. This book focuses on how to get the best out of its application, rather than replicate the technical details that appear in many text books, papers and product documentation.

The layout of the book has been designed so that the reader can progress from relatively straightforward concepts through to more complex techniques applied to more complex processes. It is assumed that the new reader is comfortable with mathematics up to a little beyond high school level. As the techniques become more specific some basic knowledge of the process is assumed, but introductory information is included – particularly where it is important to control design. Heavily mathematical material, daunting to novices and not essential to successful implementation, has been relegated to the end of each chapter.

SI units have been mainly used throughout but, where important and practical, conversion to imperial units is given in the text. Methods published in non-SI units have been included without change if doing so would make them too complex.

The book is targeted primarily for use in the continuous process industry, but even predominantly batch plants have continuous controllers and often have sections of the process which are continuous. My experience is mainly in the oil and petrochemicals industries and, despite every effort being taken to make the process examples as generic as possible, it is inevitable that this will show through. However this should not be seen as a reason for not applying the techniques in other industries. Many started there and have been applied by others to a wide range of processes.

It is hoped that the academic world will take note of the content. While some institutions have tried to make their courses more relevant to the process industry, practitioners still perceive a huge gulf between theory and practice. Of course there is a place for the theory. Many of the modern control technologies now applied in the process industry are developed from it. And there are other industries, such as aerospace, where it is essential.

The debate is what should be taught as part of chemical engineering. Very few chemical engineers benefit from the theory currently included. Indeed the risk is that many potentially excellent control engineers do not enter the profession because of the poor image that theoretical courses create. Further, those that do follow a career in process control, can find themselves working in an organisation managed by a chemical engineering graduate who has no appreciation of what process control technology can do and its importance to the business.

It is the nature of almost any engineering subject that the real gems of useful information get buried in amongst the background detail. Listed here are the main items worthy of special attention by the engineer because of the impact they can have on the effectiveness of control design.

- Understanding the process dynamics is essential to the success of almost every process control technique. These days there is very little excuse for not obtaining these by plant testing or from historically collected data. There are a wide range of model identification products available plus enough information is given in Chapter 2 for a competent engineer to develop a simple spreadsheet-based application.

- Often overlooked is the impact that apparently unrelated controllers can have on process dynamics. Their tuning and whether they are in service or not, will affect the result of steptests and hence the design of the controller. Any changes made later can then severely disrupt controller performance. How to identify such controllers, and how to handle their effect, is described in Chapters 2 and 8.

- Modern DCS include a number of versions of the PID controller. Of particular importance in the proportional-on-PV algorithm. It is probably the most misunderstood option and is frequently dismissed as too slow compared to the more conventional proportional-on-error version. In fact, if properly tuned, it can make a substantial improvement to the way that process disturbances are dealt with – often shortening threefold the time it takes the process to recover. This is fully explained in Chapter 3.

- Controller tuning by trial and error should be seen as an admission of failure to follow proper design procedures, rather than the first choice of technique. To be fair to the engineer, every published tuning technique and most proprietary packages have serious limitations. Chapter 3 presents a new technique that is well proven in industry and gives sufficient information for the engineer to extend it as required to accommodate special circumstances.

- Derivative action is too often excluded from controllers. Understandably introducing a third parameter to tune by trial and error might seem an unnecessary addition to workload. It also has a poor reputation in the way that it amplifies measurement noise, but, engineered using the methods in Chapter 3, it has the potential to substantially lessen the impact of process disturbances.

- Tuning level controllers to exploit surge capacity in the process can dramatically improve the stability of the process. However the ability to achieve this is often restricted by poor instrument design, and, often it is not implemented because of difficulty in convincing the plant operator that the level should be allowed to deviate from SP (set-point) for long periods. Chapter 4 describes the important aspects in sizing and locating the level transmitter and how the conventional linear PID algorithm can be tuned – without the need even to perform any plant testing. It also shows how nonlinear algorithms, particularly gap control, can be set up to handle the situation where the size of the flow disturbances can vary greatly.

- While many will appreciate how signal conditioning can be applied to measurements and controller outputs to help linearise the behaviour, not so commonly understood is how it can be applied to constraint controllers. Doing so can enable constraints to be approached more closely and any violation dealt with more quickly. Full details are given in Chapter 5.

- Many engineers are guilty of installing excessive filtering to deal with noisy measurements. Often implemented only to make trends look better they introduce additional lag and can have a detrimental impact on controller performance. Chapter 5 gives guidance on when to install a filter and offers a new type that actually reduces the overall process lag.

- Split-ranging is commonly used to allow two or more valves to be moved sequentially by the same controller. While successful in some cases the technique is prone to problems with linearity and discontinuity. A more reliable alternative is offered in Chapter 5.

- Feedforward control is often undervalued or left to the MVC. Chapter 6 shows how simple techniques, applied to few key variables, can improve process stability far more effectively than MVC.

- A commonly accepted problem with MVC is that, if not properly monitored, they become over-constrained. In fact, if completely neglected, they are effectively fully disabled – even though they may show 100 % up-time. Chapter 8 offers a range of monitoring tools, supplementary to those provide by the MVC vendor, which can be readily configured by the engineer.

- There are many examples of MVC better achieving the wrong operating objective; unbeknown to the implementer they are reducing process profitability. Rather than attempt to base the cost coefficients on real economics they are often adjusted to force the MVC to follow the historically accepted operating strategy. Some MVC are extremely complex and it is unlikely that even the most competent plant manager will have considered every opportunity for adopting a different strategy. Chapter 12 shows how properly setting up the MVC can reveal such opportunities.

- There are literally thousands of inferential properties, so called 'soft sensors', in use today that are ineffective. Indeed many of them are so inaccurate that process profitability would be improved by decommissioning them. Chapter 9 shows how many of the statistical techniques that are used to assess their accuracy are flawed and can lead the engineer into believing that their performance is adequate. It also demonstrates that automatically updating the inferential bias with laboratory results will generally aggravate the problem.

- Simple monitoring of on-stream analysers, described in Chapter 9, ensures that measurement failure does not disrupt the process and that the associated reporting tools can do much to improve their reliability and use.

- Compensating fuel gas flow measurement for variations in pressure, temperature and molecular weight requires careful attention. Done for accounting purposes, it can seriously degrade the performance of fired heater and boiler control schemes. Chapter 10 presents full details on how it should be done.

- Manipulating fired heater and boiler duty by control of fuel pressure, rather than fuel flow, is common practice. However it restricts what improvements can be made to the controller to better handle process disturbances. Chapter 10 shows how the benefits of both approaches can be captured.

- Fired heater pass balancing is often installed to equalise pass temperatures in order to improve efficiency. Chapter 10 shows that the fuel saving is negligible and that, in some cases, the balancing may accelerate coking. However there may be much larger benefits available from the potential to debottleneck the heater.

- Compressor control packages are often supplied as 'black boxes' and many compressor manufacturers insist on them being installed in special control systems on the basis that DCS-based schemes would be too slow. Chapter 11 describes how these schemes work and, using the tuning method in Chapter 3, how they might be implemented in the DCS.

- A common failing in many distillation column control strategies is the way in which they cope with changes in feed rate and composition. Often only either the reboiler duty or the reflux flow is adjusted to compensate – usually under tray temperature control. Chapter 12 shows that failing to adjust both is worse than making no compensation. Other common misconceptions include the belief that column pressure should always be minimised and that the most economic strategy is to always exactly meet all product specifications.

- There are many pitfalls in executing an advanced control project. Significant profit improvement opportunities are often overlooked because of the decision to go with a single supplier for the benefits study, MVC, inferentials and implementation. Basic controls, inferentials and advanced regulatory controls are not given sufficient attention before awarding the implementation contract. The need for long-term application support is often underestimated and poor management commitment will jeopardise the capture of benefits. Chapter 13 describes how these and many other issues can be addressed.

Gaining the knowledge and experience now contained in this book would have been impossible if it were not for the enthusiasm and cooperation of my clients. I am exceedingly grateful to them and indeed would welcome any further suggestions on how to improve or add to the content.

Myke King
July 2010, Isle of Wight

About the Author

Myke King is the founder and director of Whitehouse Consulting, an independent consulting organisation specialising in process control. He has over 35 years experience working with over 100 clients from more than 30 countries. As part of his consulting activities Myke has developed training courses covering all aspects of process control. To date, around 2000 delegates have attended these courses. To support his consulting activities he has developed a range of software to streamline the design of controllers and to simulate their use for learning exercises.

Myke graduated from Cambridge University in the UK with a Master's degree in chemical engineering. His course included process control taught as part of both mechanical engineering and chemical engineering. At the time he understood neither! On graduating he joined, by chance, the process control section at Exxon's refinery at Fawley in the UK. Fortunately he quickly discovered that the practical application of process control bore little resemblance to the theory he had covered at university. He later became head of the process control section and then moved to operations department as a plant manager. This was followed by a short period running the IT section.

Myke left Exxon to co-found KBC Process Automation, a subsidiary of KBC Process Technology, later becoming its managing director. The company was sold to Honeywell where it became their European centre of excellence for process control. It was at this time Myke set up Whitehouse Consulting.

Myke is a Fellow of the Institute of Chemical Engineers in the UK.

1

Introduction

In common with many introductions to the subject, process control is described here in terms of layers. At the lowest level is the process itself. Understanding the process is fundamental to good control design. While the control engineer does not need the level of knowledge of a process designer, an appreciation of how the process works, its key operating objectives and basic economics is vital. In one crucial area his or her knowledge must exceed that of the process engineer, who needs primarily an understanding of the *steady-state* behaviour. The control engineer must also understand the *process dynamics*, i.e. how process parameters move between steady states.

Next up is the field instrumentation layer, comprising measurement transmitters, control valves and other actuators. This layer is the domain of instrument engineers and technicians. However the control engineer needs an appreciation of some of the hardware involved in control. He or she needs to be able to recognise a measurement problem or a control valve working incorrectly and must be aware of the accuracy and the dynamic behaviour of instrumentation.

Above the field instrumentation is the DCS and process computer. These will be supported by a system engineer. It is normally the control engineer's responsibility to configure the control applications, and their supporting graphics, in the DCS. So he or she needs to be well-trained in this area. In some sites only the system engineer is permitted to make changes to the system. However this does not mean that the control engineer does not need a detailed understanding of how it is done. Close cooperation between control engineer and system engineer is essential.

The lowest layer of process control applications is described as *regulatory control*. This includes all the basic controllers for flow, temperature, pressure and level. But it also includes control of product quality. Regulatory is not synonymous with basic. Regulatory controls are those which maintain the process at a desired condition, or SP, but that does not mean they are simple. They can involve complex instrumentation such as on-stream analysers. They can employ 'advanced' techniques such as signal conditioning, feedforward, dynamic compensation, overrides, inferential properties etc. Such techniques are often described as *advanced regulatory control (ARC)*. Generally they are implemented

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within the DCS block structure, with perhaps some custom code, and are therefore sometimes called 'traditional' advanced control. This is the domain of the control engineer.

There will be somewhere a division of what falls into the responsibilities between the control engineer and others working on the instrumentation and system. The simplistic approach is to assign all hardware to these staff and all configuration work to the control engineer. But areas such as algorithm selection and controller tuning need a more flexible approach. Many basic controllers, providing the tuning is reasonable, do not justify particular attention. Work on those that do requires the skill more associated with a control engineer. Sites that assign all tuning to the instrument department risk overlooking important opportunities to improve process performance.

Moving up the hierarchy, the next level is *constraint control*. This comprises control strategies that drive the process towards operating limits, where closer approach to these limits is known to be profitable. Indeed, on continuous processes, this level typically captures the large majority of the available process control benefits. The main technology applied here is the *multivariable controller* (*MVC*). Because of its relative ease of use and its potential impact on profitability it has become the focus of what is generally known as *advanced process control* (*APC*). In fact, as a result, basic control and ARC have become somewhat neglected. Many sites (and many APC vendors) no longer have personnel that appreciate the value of these technologies or have the know-how to implement them.

The topmost layer, in terms of closed loop applications, is *optimisation*. This is based on key economic information such as feed price and availability, product prices and demand, energy costs etc. Optimisation means different things to different people. The planning group would claim they optimise the process, as would a process support engineer determining the best operating conditions. MVC includes some limited optimisation capabilities. It supports objective coefficients which can be set up to be consistent with process economics. Changing the coefficients can cause the controller to adopt a different strategy in terms of which constraints it approaches. However those MVC based on linear process models cannot identify an unconstrained optimum. This requires a higher fidelity process representation, possibly a rigorous simulation. This we describe as *closed-loop real-time optimisation* (*CLRTO*) or more usually just *RTO*.

Implementation should begin at the base of the hierarchy and work up. Any problems with process equipment or instrumentation will affect the ability of the control applications to work properly. MVC performance will be restricted and RTO usually needs to work in conjunction with the MVC. While all this may be obvious, it is not necessarily reflected in the approach that some sites have towards process control. There are sites investing heavily in MVC but which give low priority to maintaining basic instrumentation. And **most** give only cursory attention to regulatory control before embarking on implementation of MVC.

2

Process Dynamics

Understanding process dynamics is essential to effective control design. Indeed, as will become apparent in later chapters, most design involves performing simple calculations based solely on a few dynamic parameters. While control engineers will commit several weeks of round-the-clock effort to obtaining the process dynamics for MVC packages, most will take a much less analytical approach to regulatory controls. This chapter aims to demonstrate that process dynamics can be identified easily and that, when combined with the design techniques described in later chapters, it will result in controllers that perform well without the need for time-consuming tuning by trial-and-error.

2.1 Definition

To explore dynamic behaviour, as an example, we will use a simple fired heater as shown in Figure 2.1. It has no automatic controls in place and the minimum of instrumentation – a temperature indicator (TI) and a fuel control valve. The aim is to ultimately commission a temperature controller which will use the temperature as its *process variable* (PV) and the fuel valve position as it *manipulated variable* (MV).

Figure 2.2 shows the effect of manually increasing the opening of the valve. While the temperature clearly rises as the valve is opened, the temperature trend is somewhat different from that of the valve. We use a number of parameters to quantify these differences.

The test was begun with the process steady and sufficient time was given for the process to reach a new steady state. We observed that the steady state change in temperature was different from that of the valve. This difference is quantified by the *steady state process gain* and is defined by the expression

$$\text{process gain} = \frac{\text{change in temperature}}{\text{change in valve position}} \tag{2.1}$$

Process gain is given the symbol K_p. If we are designing controls to be installed in the DCS, as opposed to a computer-based MVC, K_p should generally have no dimensions. This

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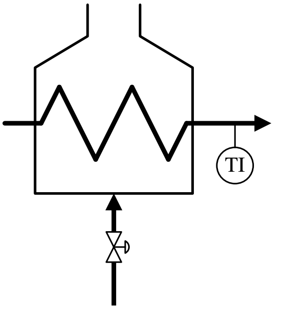

Figure 2.1 Process diagram

is because the DCS works internally with measurements represented as fractions (or percentages) of instrument range.

$$K_p = \frac{\Delta PV}{\Delta MV} \tag{2.2}$$

where

$$\Delta PV = \frac{\text{change in temperature}}{\text{range of temperature transmitter}} \tag{2.3}$$

and

$$\Delta MV = \frac{\text{change in valve position}}{\text{range of valve positioner}} \tag{2.4}$$

Instrument ranges are defined when the system is first configured and generally remain constant. However it is often overlooked that the process gain changes if an instrument is

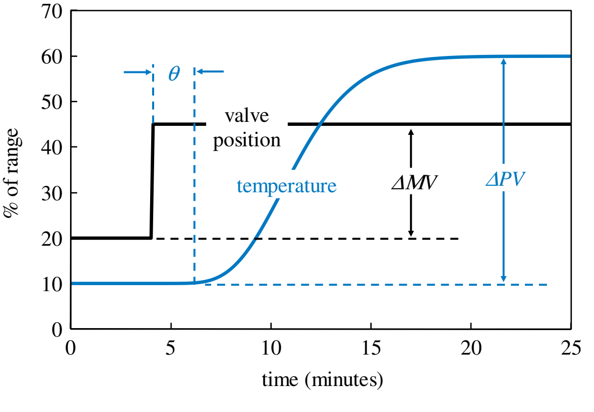

Figure 2.2 Process response

later re-ranged and, if that instrument is either a PV or MV of a controller, then the controller should be re-tuned to retain the same performance.

Numerically K_p may be positive or negative. In our example temperature rises as the valve is opened. If we were to increase heater feed rate (and keep fuel rate constant) then the temperature would fall. K_p, with respect to changes in feed rate, would therefore be negative. Nor is there is any constraint on the absolute value of K_p. Very large and very small values are commonplace. In unusual circumstances K_p may be zero; there will be a transient disturbance to the PV but it will return to its starting point.

The other differences, in Figure 2.2, between the trends of temperature and valve position are to do with timing. We can see that the temperature begins moving some time after the valve is opened. This delay is known as the *process deadtime*; until we develop a better definition, it is the time difference between the change in MV and the first perceptible change in PV. It is usually given the symbol θ. Deadtime is caused by *transport delays*. In this case the prime cause of the delay is the time it takes for the heated fluid to move from the firebox to the temperature instrument. The DCS will generate a small delay, on average equal to half the controller scan interval (ts). While this is usually insignificant compared to any delay in the process it is a factor in the design of controllers operating on processes with very fast dynamics – such as compressors. The field instrumentation can also add to the deadtime; for example on-stream analysers may have sample delays or may be discontinuous.

Clearly the value of θ must be positive but otherwise there is no constraint on its value. Many processes will exhibit virtually no delay; there are some where the delay can be measured in hours or even in days.

Finally the shape of the temperature trend is very different from that of the valve position. This is caused by the 'inertia' of the system. The heater coil will comprise a large mass of steel. Burning more fuel will cause the temperature in the firebox to rise quickly and hence raise the temperature of the external surface of the steel. But it will take longer for this to have an impact on the internal surface of the steel in contact with the fluid. Similarly the coil will contain a large quantity of fluid and it will take time for the bulk temperature to increase. The field instrumentation can add to the lag. For example the temperature is likely to be a thermocouple located in a steel thermowell. The thermowell may have thick walls which cause a lag in the detection of an increase in temperature. Lag is quite different from deadtime. Lag does not delay the **start** of the change in PV. Without deadtime the PV will begin changing immediately but, because of lag, takes time to reach a new steady state. We normally use the symbol τ to represent lag.

To help distinguish between deadtime and lag, consider liquid flowing at a constant rate (F) into a vessel of volume (V). The process is at steady state. The fraction (x) of a component in the incoming liquid is changed at time zero ($t=0$) to x_{new}. By mass balance the change in the quantity of the component in the vessel is the difference between what has entered less what has left. Assuming the liquid is perfectly mixed then

$$V.dx = F.dt.x_{new} - F.dt.x \tag{2.5}$$

Rearranging

$$\frac{V}{F}\frac{dx}{dt} + x = x_{new} \tag{2.6}$$