Improving Refinery Performance: Process and Control Information from Step Testing

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Abstract

Step testing is typically only used during advanced control implementation to understand the effects of one process variable on other process variables. Not only can step testing prove valuable in understanding interactions between process variables when implementing model predictive control (MPC), it can also be used to identify control valve deficiencies, understand overall process dynamics, and assist with tuning control loops. This paper will demonstrate how step testing helped improve process control and process performance around a crude desalter, hydrocracker hydrogen quench, and gasoline blender.

Introduction

This paper is intended to encourage process control engineers to understand the dynamics of their processes and use this information to improve process control performance. Step testing is a simple and inexpensive tool that every control engineer should utilize.

A step test is when you place a controller in manual and change the output a small percentage so as not to significantly upset the process and then observe the response in the process variable. As shown on the right in Figure 1, the response can be the most common type, a first-order plus dead time response. There are many types of responses that could be observed such as integrating, second order response, inverse response and others. This paper will address the first-order plus dead time response.

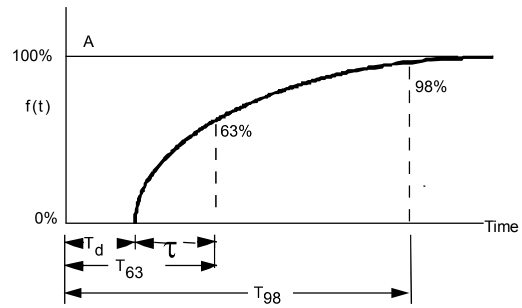

Figure 1: First order with deadtime response

Traditional Step Testing and Model Predictive Control (MPC)

In order to implement MPC on a process, one must first go through a series of moves on each manipulated variable to determine the effects on all controlled variables and the time-to-steady-state. The manipulated variables should be moved in both positive and negative directions from the current operating point to observe the process response. The data collected in the manipulated variables are fit to algorithms in order to generate a dynamic model that best represents the current process operation around the tested operating conditions (design conditions versus turndown rates).

Control Valve Deficiencies

Before beginning any process control optimization program or implementing MPC, one should first look at field device performance and deficiencies. A simple and inexpensive first phase is to execute step tests on each manipulated variable to determine the performance of the final control element. The size of the output step should be large enough to observe the response above any process noise that may be present, yet small enough to minimize process upsets.

To test the performance of a control valve, put the controller in manual and begin with small steps in one direction. Ensure you understand the process before executing step tests to prevent unwanted process upsets (for example, do not reduce hydrogen quench flow to cause a hydrocracker reactor temperature runaway; start with increasing hydrogen and then changing direction back to normal hydrogen use). Start with small steps such as 0.25%, 0.50% or 1% depending on sensitivity of the process. If you do not see a change in flow/temperature/pressure, continue making step changes until you do see a change. It is not uncommon to see control loops requiring an output change in excess of 2% to see changes from the final control element movement (Figure 2 above shows a limit cycle due to a sticky valve). Under-performing field devices add process constraints that cannot be tuned out or compensated with MPC.

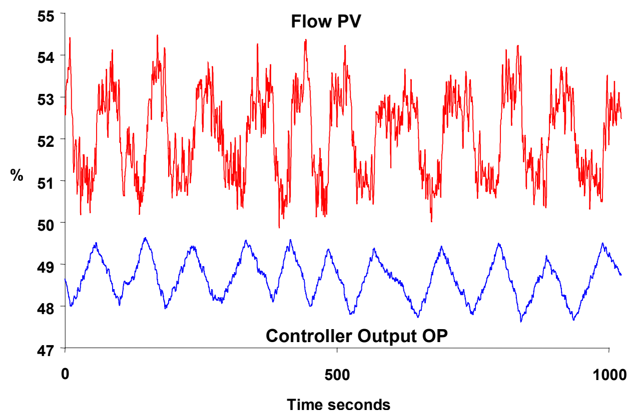

Figure 2: Limit cycle in a flow loop due to a sticky valve

Overall Process Dynamics and Assistance in Tuning Loops

Step tests can be a useful tool in determining overall process dynamics and time-to-steady-state. This information can be used in MPC and to assist in non-MPC tuning. The primary process dynamic information attained from a step test is steady-state process gain, dead time, and the time constant. Typically these parameters are an average of several step tests around the normal operating range.

The steady-state gain can be calculated by taking the measured process response before and after the step test and then dividing by the output change. The time constant Tau is the time required to reach 63.2% of the final process change.

The dead time is the amount of time after the step test is made until the process actually begins to move. See Figure 3.

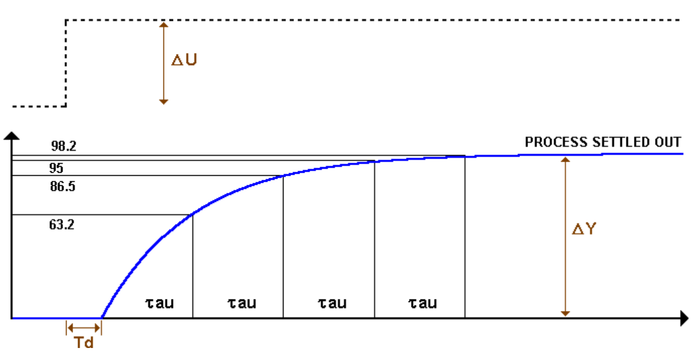

Figure 3: Typical first order step response

Once an engineer understands the process dynamics, model-based tuning such as Lambda Tuning can be applied to set the speed of response for each controller based on operating objectives and priorities. "Lambda Tuning" refers to tuning methods where the control loop speed of response is a selectable tuning parameter; the closed loop time constant is referred to as "Lambda" (λ). Lambda Tuning originated with Dahlin [1] in 1968; it is based on the same IMC theory as MPC [2, 3]. It is model-based and uses pole-zero cancellation to achieve the desired closed loop performance. A recommended Lambda starting point to ensure robustness is three times the larger of Tau or Deadtime. The time for the loop to reach setpoint is approximately four times the selected Lambda value.

<u>Self-Regulating Process (First-order with deadtime model, Classical PID) [3, 4]</u>

τ = time constant (from step test)
T_d = dead time (from step test)
K_p = steady state process gain (from step test)
λ = closed loop time constant (Lambda – user defined speed of response)

<u>Controller Settings</u>

$T_r = \tau$
$K_c = (1 / K_p) * T_r / (\lambda + T_d)$

Crude Desalter (Decoupling Loop Interactions without MPC)

For a refiner, a difficult task is to tune the flow and pressure controllers around a desalter so that they do not fight one another. These controllers have long time constants and long delay times before action sees the measured variable. Tuning-by-feel often results in de-tuning one controller to the point it is similar to being in manual operation. By understanding the process dynamics of each control loop and the overall process operating objectives, one can ensure the loops can be tuned to maximize loop performance while minimizing the tendency of the two loops to fight each other in closed loop control.

Here are the dynamics for one refiner's desalter.

	PT Downstream Pressure	FT Upstream Flow
PVC Pressure Control Valve	Gain 0.87 %/% Delay 15 sec. Tau 20 sec.	Gain 0.24 %/% Delay 15 sec. Tau 30 sec.
FVC Flow Control Valve	Gain -0.4 %/% Delay 11 sec. Tau 30 sec.	Gain 0.2 %/% Delay 25 sec. Tau 35 sec.

One refiner had difficult interactions between the pressure and flow controller around their desalter that resulted in a pressure change of –7 psig from setpoint when the flow increased 1% from normal. This was an operation issue when decreasing flow rate since the pressure increased and lifted the relief valve on the desalter vessel. A standard procedure for the board operator was to manually manipulate the pressure control valve when reducing feed rates to avoid lifting the relief valve. Figure 4 shows the 20-psig pressure change related to a 3% increase in feed rate. The tuning was changed with a 30-second Lambda value for the pressure controller and a 240-second Lambda value for the flow controller. The end result is displayed in Figure 5 with a 3% change in feed rate, yet the pressure only changed 6-psig (a 67% reduction in variation from setpoint). Flow rates could be changed and the pressure controller could remain in automatic without lifting the relief valve.

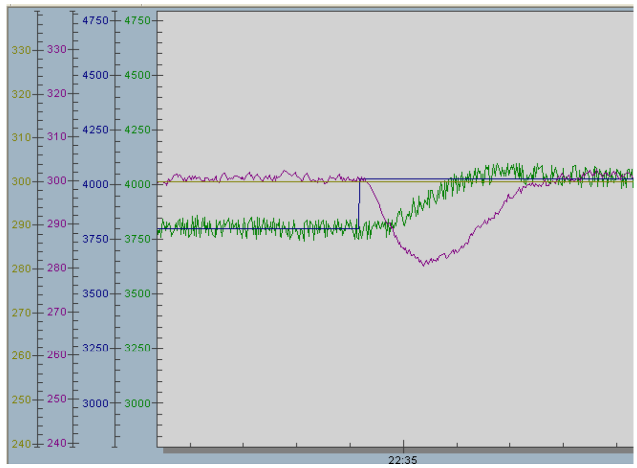

Figure 4: Desalter Flow and Pressure Interactions

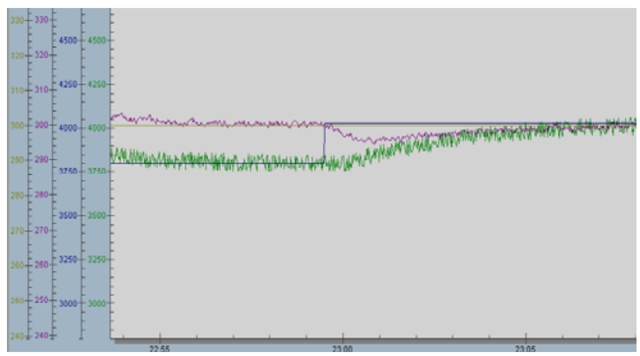

Figure 5: Desalter Flow and Pressure De-Coupled

A simple rule used in cascade control loops is to have the inner control loop tuned 5-10 times faster than the outer loop. The same rule is used to decouple interacting control loops not used in MPC. The most important control parameter, the desalter pressure, will be tuned 5-10 times faster than the crude flow controller.

Hydrocracker Hydrogen Quench (Detecting Hardware Deficiencies)

A refiner was experiencing +/-3 °F control on a reactor bed temperature control via a hydrogen quench. No matter what tuning was used, the performance of the controller could best control within +/-3 °F. As shown in Figure 6, the problem was not the tuning but the resolution of the control valve.

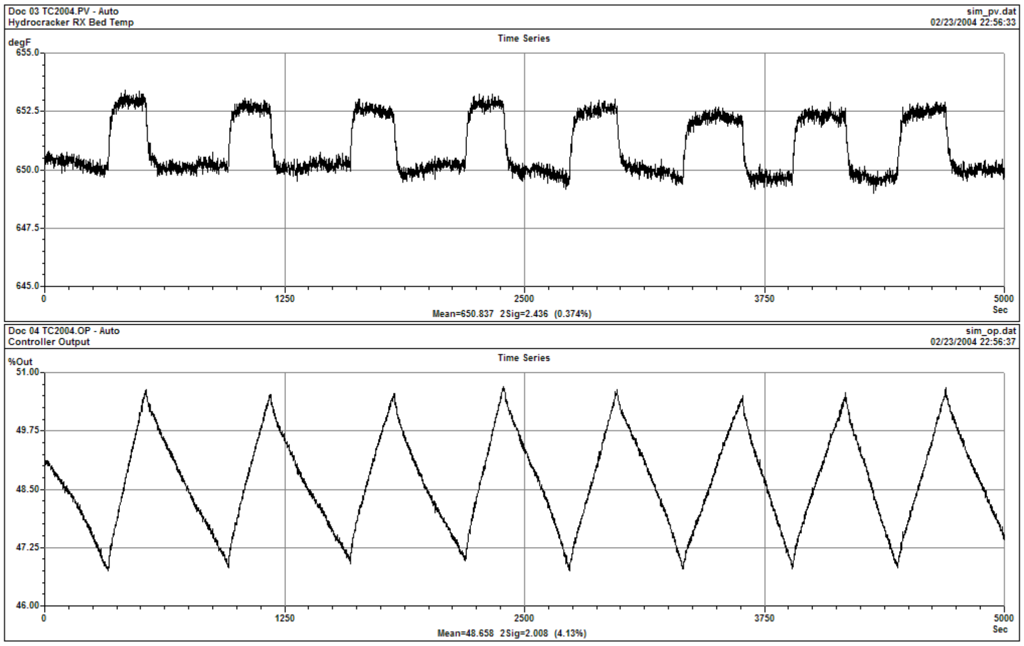

Figure 6: Limit cycle in Rx Bed Quench Temperature due to 3% valve stiction

The hydrogen quench control valve was repaired with less friction packing and a new two-stage positioner which resulted in much tighter control around the hydrocracker reactor bed temperature. The controller was re-tuned and the temperature control is observed below in Figure 7. The temperature set point was increased slightly with the improvement in control without additional hydrogen required.

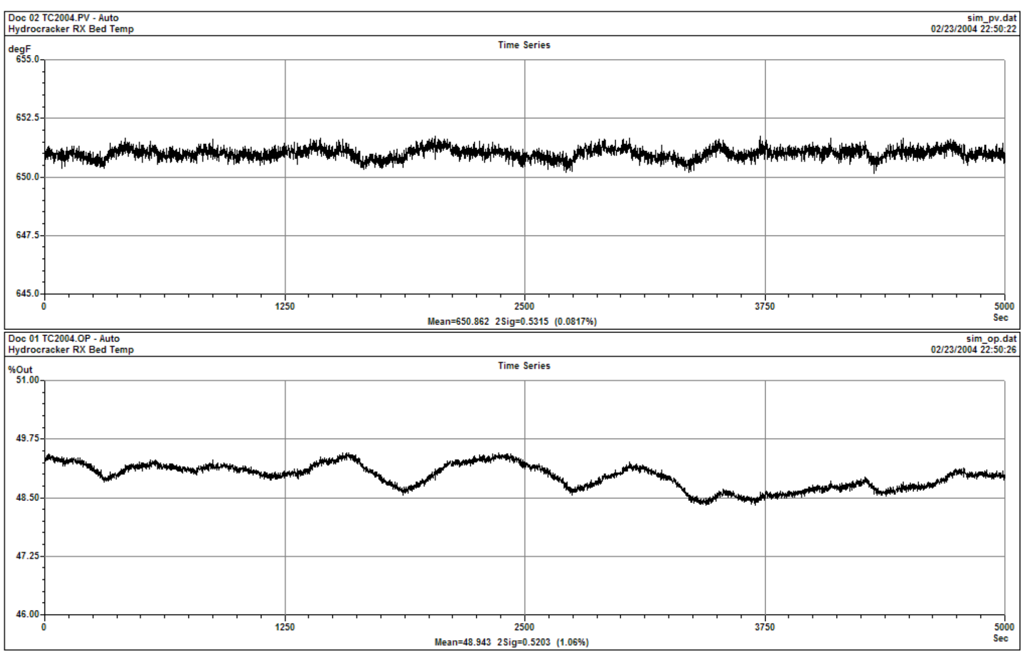

Figure 7: Reactor Bed Quench Temperature Control after improving valve performance

Gasoline Blending (Making Component Dynamics Perform the Same)

Many refiners use fudge factors on their gasoline blends based on experience to account for non-linear octane blending. These compensating factors only work well if no additional field device deficiencies or additional non-linear behaviors exist for each blend component controller. In this application, it is desirable for all of the blend components to have the same dynamic response to maintain a consistent blend even when the components are ramped up or down.

For example, here is one refiner's gasoline component blend control system along with a description of each component dynamics (determined from step testing):

Component	Line Size	Valve Size	Valve Type	Gain	Dead Time (sec.)	Time Constant (sec.)
n-Butane	3	3	Equal %	0.95	2.0	2.2
Isomerate	3	3	Linear	0.57	2.5	2.7
Reformate	4	3	Linear	0.73	3.0	3.0
Cat Gas	6	4	Equal %	0.87	2.3	3.3
Alkylate	3	3	Equal %	0.36	3.2	3.7
Back Pressure	6	6	Equal %	-1.10	2.3	2.7

Tight control of the back pressure controller is desirable to maintain a consistent upstream pressure on the blend flow valves which will stabilize the individual flow controllers. Applying Lambda Tuning, the Back Pressure controller will be tuned to a Lambda of 8 seconds (or 3 times the open loop time constant). To minimize interaction between the backpressure and the flow controllers, it is necessary to tune the flow controllers at least 5 times slower than the pressure controller. Here, each of the flow controllers will be tuned to the same Lambda of 40 seconds (5 times slower than the back pressure controller).

As a result of step testing, finding control element deficiencies and fixing them, understanding the process dynamics, and applying the operation objectives to determine desired speed of response for each controller, the primary benefits for this refiner were as follows:

- 80% reduction in the number of off-spec blends (already had a blend optimization package implemented)
- Cost avoidance related to additional lab samples for off-spec blends
- Better inventory management (less off-spec taking up storage)
- Demurrage avoidance (again due to off-spec blends)
- Octane target shifted from +0.2 above specification to +0.1
- Repeatable process control operation to learn non-linear octane blending characteristics
- Total octane and RVP giveaway reduced from approximately $250K/month to less than $50K/month.

Conclusions

Process control engineers have the ability to understand more about the dynamics of their processes and then use this information to improve process control performance. Step testing is a simple and inexpensive tool that every control engineer can utilize. It can be used to identify control valve deficiencies, understand overall process dynamics, and assist with tuning non-MPC control loops.

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