IMPROVING EFFICIENCY OF PROCESS CONTROL OPTIMIZATION FOR BATCH CHEMICAL SYSTEMS

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INTRODUCTION

Control projects take time to do right. In recent days, many companies are cutting back on the number of staff while increasing the automation and engineer involvement in operations. Long-term control optimization projects can be difficult to prioritize when compared to daily manufacturing troubleshooting and high profile capital work. A recent mentorship with Gregory K. McMillan, a distinguished ISA controls expert, helped boost the priority of control work at this plant, but it was still difficult to fit in controls projects with so many other immediate demands. The purpose of this report is to support engineers in the identification and completion of controls projects when they are unable to focus solely on controls.

Based on the capabilities of a relatively new chemical engineer in a small batch manufacturing plant, this paper will follow the troubleshooting experience for improving the temperature control of a batch stripping column, with an emphasis on ways to delegate and automate time-intensive tasks. The report will briefly cover the definition, initial findings, cost justification and selection of new instruments, strategies implemented and final results of the control project. Recommendations from Mr. McMillan are included and noted, along with personal interpretations and implementation.

IDENTIFYING THE PROBLEM

TOOLS TO FIND BATCH CONTROL ISSUES

As a process or control engineer, time can often be better spent justifying and prioritizing projects other people have identified than spent finding new problems. Often, projects are developed through time-intensive batch-to-batch variation, quality, and failure studies, but experience has shown that
behind almost every successful optimization project, there is an operator, technician, or process expert pushing to have some problem solved. Recent examples at this plant include mechanics replacing a cycling valve too often and production planning having difficulty scheduling a certain product’s hours consistently. Prioritizing and working on their projects rather than digging for new ones has several benefits that will help complete projects, including firsthand experience with the issue, operations-level buy in on the new controls, and the timely completion of cost justification work to obtain upper level support.

The control issues discussed in this paper were making life difficult for operations personnel, as the lack of good control was necessitating constant operator intervention. After an initial bout of troubleshooting, detailed below, the justification to fix this inconvenience for operations mushroomed into a complex cost-driven effort to improve batch-to-batch variability and enable other high value improvement projects.

From this example, process and process control engineers should see the benefit of gleaning project ideas, but it can be done in several ways. Informal discussions with operations about equipment operability are ideal times to discover ideas, but digressions into off topics can drain an engineer’s schedule. One tool this plant uses is a formalized issue log, or complaint database, which helps engineering learn about and act on control and logic issues in a timely fashion. A set meeting to dig into and discuss progress on these issues ensures face to face communication and reduces misinformation.

DETERMINING SCOPE AND GOALS FOR TROUBLESHOOTING

Once a solid problem is identified, the real project work begins. Best practices for time management include goal setting, and that should be the first step in controls work as well. How much variability can the equipment successfully handle? For batch processes, what products are really important to control, and during which phases? Most importantly, when can the project be considered complete?

For the batch stripping column example, there are three temperatures across the packed bed column, all of which showing oscillations and spikes during normal operation. A high top bed temperature could put unwanted material in the overhead condenser, so it trips the safety interlocks and holds the batch. The very bottom temperature is also important, as it determines if material being stripped out is falling back into the batch, extending cycle time only during the stripping phase. These temperatures are both important, and need to be kept within limits, but do not actually need to be controlled at set point or be brought to set point quickly. The controls for this column just need to reject the natural heat load disturbances from the batch as time progresses. The temperature across the middle of the bed is helpful information for operations, but does not need to be controlled, greatly reducing the scope of the project.
Product wise, there are two types of products run on this equipment: products that need a high reflux, and products that need a low reflux. Through discussions with operations and initial data gathering, focus for the project shifted to the low reflux products. The controls for high reflux were adequate to meet the temperature goals stated above and did not need to change, which further reduced the scope of work.

**INITIAL TROUBLESHOOTING**

**PHYSICAL SYSTEM**

For background into this issue, the stripping column in question is a packed-bed column that sits directly on top of the reactor, so while there is a condenser, decanter, and top reflux, there is no reboiler. The liquid bottoms stream falls directly back into the reaction mixture. The column is heated by the reactor vapor stream directly, with loads that vary widely through the reaction and stripping phases.

Depending on the product, reflux is provided by the condensed and pumped overhead material or by material straight off of a header, which give the control valve(s) differing head pressures. Temperatures are measured in three places vertically along the bed, along with bottom and top vapor space measurements. Reflux flow is dispersed on top of the bed by a distributor, and is currently controlled by small and large split range valves based off of the indicated flow meter.

The control for this system when the project was identified was a slave split range flow controller cascaded to an average master temperature controller. The average was based off the middle and bottom temperature in the packed bed in order to help catch both a swift top temperature rise and an equally abrupt bottom temperature plummet.

In practice, there was a considerable amount of dead time (about 5 minutes) between reflux increasing and the bottom temperature decreasing. This means that the average temperature still read high several minutes after the top temperature was back in acceptable range. This resulted in the bottom temperature dropping too low and stripped material being dropped back into the reactor.

In addition, the system was not handling temperature spikes well. The bottom temperature would start to increase, but the interlocked top temperatures would not see it for several minutes. By the time the average temperature reached a high enough value to open the reflux up, the temperature controller was
demanding 100% output from the slave flow controller. The column would overheat, or almost overheat, then respond by dumping a deluge of reflux into the column and start the cycle all over again.

[Diagram of Stripping Column]

**SOFTWARE SETUP AND DATA CAPTURE**

The above summary of the starting process control and its issues could have been obtained in several ways, some more hands on than others. Conventional wisdom dictates that watching a process run live is the best way to know how it works. In reality, however, it is often impossible to sit through an entire batch and watch interactions. Batches run too long, meetings and urgent tasks interfere, or the really important step of the process occurs consistently around 3am. So for the process summary, the data historian on the plant’s distributed control system did most of the work, along with some clarification discussions with operations.

At first the information garnered from the process historian was almost useless. There were large data filters, substantial data compression on key temperatures and flows, and not everything needed was historized. Mr. McMillan offered the following guidelines for compression, filter time and update time:

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“Compression should be less than 1/5 the control band (allowable error in the controlled variable), the filtering should be less than 1/10 the dead time, and the update time interval should be less than 1/5 the dead time. Some typical numbers:

<table>
<thead>
<tr>
<th>Loop</th>
<th>Compression</th>
<th>Filter Time</th>
<th>Update Time</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow</td>
<td>0.1% [of range]</td>
<td>0.5 sec</td>
<td>1 sec</td>
</tr>
<tr>
<td>Gas Pressure</td>
<td>0.1%</td>
<td>0.5 sec</td>
<td>1 sec</td>
</tr>
<tr>
<td>Liquid Pressure</td>
<td>0.1%</td>
<td>0.05 sec</td>
<td>0.1 sec</td>
</tr>
<tr>
<td>Vessel Level</td>
<td>0.2%</td>
<td>5 sec</td>
<td>10 sec</td>
</tr>
<tr>
<td>Distillate Level</td>
<td>0.02%</td>
<td>1 sec</td>
<td>2 sec</td>
</tr>
<tr>
<td>Temperature</td>
<td>0.02%</td>
<td>5 sec</td>
<td>10 sec</td>
</tr>
</tbody>
</table>

The distributed control system used on site offered a way to bulk export, edit and re-import Excel files of historian setup data, which saved considerable time during the editing and downloading of the new values.

To aid in troubleshooting, batch logic can be altered to identify phase changes in the continuous historian. For example, temperature tuning and set points often changed between phases, but these changes were not captured in batch report as they were done automatically. To identify the first and most important phase change more readily, the flow controller was programmed to fully stroke prior to getting its set point. This helped the control valve unstick at the phase change, and gave an easy trigger for gathering and exporting historian data to statistical software for analysis.

**FINDING HARDWARE LIMITATIONS**

Initial attempts to bring this loop back into control using conventional loop tuning (as outlined in the ISA booklet *Good Tuning: A Pocket Guide*) and even manual intervention failed miserably. The flow was not acting consistently, and operators were complaining that the valves were leaking through, as the column would cool off even with the valves shut. If the hardware and instruments would not hold, then all control attempts would be useless, so they became the focus for the next step in the investigation.

Working first on the slave loop, data was pulled from the historian into statistical software for manipulation. The biggest tip off of issues is shown below, in the chart of percent output on the reflux flow controller to the measured reflux flow. This was a controller for split ranged valves with the following ranges:
0 – 27% OP = 0 – 100% small valve

14-100% OP = 0-100% big valve.

The first 10 percent of output on the flow controller, which is over one-third of the total output for the small valve of the set, is showing zero flow. The larger valve starts cracking open somewhere between 14 percent total output and 30 percent total output, depending on the stiction at the time. Even when the large valve is mostly open, it shows a large scattering effect from flow meter error. The decision to replace the larger valve was made, as the old one did not have any positioner feedback, and was a different brand than those in stock on site.

Concern then turned to the smaller, newer valve. Why was it showing zero flow almost across its whole range when it had a (confirmed in the field) flow? The flow meter vendor was contacted, who helpfully provided the original specifications and error limits for the 20-year-old flow meter. As it turned out, while the plant had changed products and required flows over the years, it had not upgraded all of the instruments, and the flow meter being used for control was woefully oversized. For the newer low reflux products, the entire reflux flow range is shaded in grey on the chart below. With maximum flow rates of less than two percent of the current total range for the flow meter, a new flow meter would be needed for any hope of control.

![Figure 2: % Output to Flow for Flow Controller](image-url)
COST JUSTIFICATION OF NEW INSTRUMENTS

SAFER AND FASTER

In tandem with the troubleshooting during the initial stages of the project, general cost justification was done in order to obtain and keep management support of the work. Value was based on rough estimations of reduced cycle time, which was then translated into a dollar value via additional capacity or reduced utilities, depending on the current market. The first cycle time reduction estimate, for decreased stripping time, was obtained from several historian graphs like the one below, where the top lines are the bottom column temperature and set point (which are tracking, as the measurement is not in control) and the bottom line is the output to the flow controller. Reflux flow decreases steadily until it nears the point where reliable flow measurement is lost.

While the actual amount of fluid fed back to the reactor could be calculated based on an assumed flow rate, with a reduced time based on lab data for typical stripping rates, this estimate was simplified for brevity. Basically, whenever the temperatures dropped below the vaporization point of the refluxed fluid, no stripping was taking place. During sold out conditions, this time value can be directly translated to dollars using an average batch size, cycle time, and $/unit value. In other times, the utility

Figure 3 Percent of Flow Range to Percent Error for Flow Meter
savings for steam, nitrogen, and fuel oil over that time period for an annual number of batches would serve for cost savings.

An almost identical calculation was done for the heat up chart of the batches, looking at top temperature limits instead of bottom. The improved control would reduce heat up time, as the new control scheme would respond fast enough to keep column top temperature in control and the batch out of hold mode. In addition, downtime from overhead plugging was also calculated.

AN “ENABLING” PROJECT

If data for the control issue is inconsistent or difficult to obtain, control work can also be justified as an enabling project. Specifically, better control usually enables further process optimization and savings by operating closer to set point or, in this case, much further from the gutter. By improving temperature control of the column, other ongoing projects with better defined economics can move forward. For this system, these enabled projects included efforts to increase heat up rates, reduce volatile organics to the oxidizer header in the overhead stream, and increase reliability. Each of these issues already had defined economic or environmental justification that was borrowed to keep the control work a priority.
SIZING, SELECTION, AND IMPLEMENTATION

SIMULATION VERSUS FIELD MEASUREMENT FOR VALVE SIZING

Troubleshooting work had shown that the instruments, specifically a large control valve and a flow meter, needed to be replaced. The justification work calculated what could be conscientiously spent. Finally, the equipment could be sized and purchased. Here are a few learnings from the selection of these instruments.

First, the vendor generally knows more than the engineer about the types of valves available and valve sizing software. The vendor will not, however, know enough about the system to be able to select a valve without help, so there is no reason to shortcut through this step unless delegation to a plant engineer is an option. Providing flow rates, temperatures, process fluids, and general E&I specifications can be done quickly, but accurate pressure drop calculations will make or break the abilities of the new control scheme. Always walk down an existing system to sketch out upstream and downstream piping, as drawings may not capture a two-foot section of half-inch line someone temporarily put in as a fix and forgot. The pump curves, turns, transitions, and pipe lengths can be put into a simulator such as Engineer’s Aide™ or a spreadsheet like Excel to calculate pressures and pressure losses upstream and downstream of the control valves. For assistance, reference Flow of Fluids through Valves, Fittings, and Pipe or Practical Piping Design.

For this system, the valves needed to operate well with two different head pressures, one from the pump and one from the reflux header system. In addition, new styles of flow meters that could handle the correct range of flow had a significantly higher pressure drop which had to be taken into account.

As a final test, pressure gauges were tied into sample ports just upstream and downstream of the current valve system. During downtime, various flow rates were tested and pressure drops compared to the calculated values. Once satisfied, the process information was rolled over to the vendors for final valve and meter recommendations.

CONTROL VALVES LEAK THROUGH

During the valve selection, a rather interesting discussion arose about the desired leak characteristics of the control valve. As mentioned previously, the operators had complained about apparent leak through, and it was asked if the control valve was the right place to fix leaks. This question was posed to Mr. McMillan, who had the following recommendation:
“All control valves [are] designed for throttling service leak. It is a matter of how much. There are leakage classifications to quantify how much leakage there is. The lower the leakage the higher the friction (stiction) from tighter seating and sealing surfaces as the valve closes. For isolation (essentially zero leakage), on-off block valve should be added that automatically closes when the control valve signal is 0% and automatically opens when the signal is greater than 0% plus some deadband (e.g. 0.2%). I also suggest the use of an automated on-off block valve for a safety instrumentation system (SIS) shutoff of streams rather than relying upon the control valve. A control valve should be viewed as a throttle valve (throttling flow and not completely stopping flow). An on-off block valve should be used for an isolation and SIS valve.”

Based on this information, a more relaxed leakage classification was chosen to reduce friction near the seat, and block valves were added prior to the inlet lines, tied to the flow controller output. Now, when the control valve closes, so does the block valve.

**HIGH TURNDOWN AND STAT/STOP OPERATIONS**

The last issue discovered during instrument selection and installation was that the instrument options for batch operations are limited. For this application at least, the reflux controller needed to control over a range with a 70:1 turndown, depending on product, and lines commonly run empty during the batch starts and stops. These complexities narrowed down the meter options and increased the price significantly. Bring these issues up early with the vendor to weed out unsuitable tools quickly.

**DEVELOPING A NEW STRATEGY**

**SELECTING THE BEST CONTROL POINT**

While waiting on the new instruments to arrive, a proposed new control scheme was being developed with advice from more experienced engineers, control modeling software, and historical data. First, a primary control measurement was selected. The average temperature was a pretty stable measurement, but did not exhibit a fast enough or large enough response to adequately control the system. Mr. McMillan offered the following advice on choosing stable temperatures in a packed bed column:

“The temperature showing the largest and most symmetrical change in temperature for a change in reflux is normally the best location [to control]. However, with a packed column, poor flow
distribution can cause localized hot spots and cold spots. ... If several temperature measurements in your column show a large and nearly symmetrical response but exhibit some temporary inconsistencies, you could try an average...

[To identify a measurement stable enough for control] first, the measurement needs to be representative of what you [are] trying to control and [be] significantly affected by the disturbances you [are] trying to reject. This is best seen by monitoring changes in the measurement for changes in the manipulated and disturbance streams. Second, the measurement should have a dead time less than 1/20 of the cyclic disturbance period otherwise the disturbance is too fast ... Third, the measurement threshold resolution must be less than ½ the allowable control error for the entire [process] measurement range (must have enough rangeability to handle turndown). Fourth, the measurement drift should be less than ½ the allowable control error between scheduled maintenance. Fifth, the noise must be small and fast enough to be filtered without appreciably slowing the loop’s ability to see disturbances.”

Based on these recommendations, the reflux was shifted up and down while the temperatures were monitored during a normal strip. The largest and most symmetrical response was observed in T2, as shown in the history chart below, so it was selected as the new control measurement. This temperature measurement is very close to the reflux inlet, so there is minimal dead time between a manipulated disturbance and measurement. It also meets the other stability standards specified. The reversals in temperature along the height of the column, such as where T3 is cooler than T2, indicate channeling at low flows, but no efforts will be made to correct this (yet).

Figure 5 Reflux effect on column temperatures
ADDING FEEDFORWARD

While the top temperature responded quickly to changes in the reflux flow, the controller did not show a fast enough response to disturbances from the reactor at the base of the column. To handle this, a feed forward factor based on the bottom column temperature will be included in the flow controller. This bottom temperature should indicate changing load from the reactor to the column, while feedback response from the top column temperature will ensure tight reflux flow control.

First, the feed forward compensator block has to ensure the timing of the feed forward signal is correct. As Mr. McMillan stated, “if the feed forward signal arrives too soon, the initial response of the PV will be in the opposite direction of the final response from the disturbance. The result is an inverse response… If the disturbance arrives too late, a second disturbance is created from feed forward action. In both cases, an oscillation develops, and the error from the disturbance is increased.” For this system, the overall time delay is equal to the time delay between change in bottom temperature and change in control temperature minus the time delay between change in reflux and change in control temperature. Similarly, the overall time constant will be equal to the time constant between change in bottom temperature and change in control temperature minus the time constant between change in reflux and change in control temperature. These values can be determined using historical data, or by using control modeling software.

For other systems, Mr. McMillan offers the following guidelines:

\[ \text{TD2} = \text{TD1} (\text{time delay between change in bottom temp and change in control temp}) - \text{TD3} (\text{time delay between change in reflux and change in control temperature}) \]

\[ \text{TC2} = \text{TC1} (\text{time constant between change in bottom temp and change in control temp}) - \text{TC3} (\text{time constant between change in reflux and change in control temperature}) \]

If \( \text{TD2} > 0 \) then
   Feed forward dead time = TD2
else
   Feed forward dead time = 0
endif

If \( \text{TC2} > 0 \) then
   If \( \text{TD2} > 0 \) then
      Feed forward lead time = 0
   else
      Feed forward lead time = TD2
   endif
   Feed forward lag time = TC2

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else
  Feed forward lead time = TC2
  Feed forward lag time = 0.2 x TC2 (increase to filter out noise in T1)
endif

The actual set point of the reflux flow controller will be equal to the sum of the present value, a feedback correction, and a feed forward correction. The feedback correction is the normal PID response, which is trended and put on the operating screen. The feed forward correction will be the temperature change multiplied by a coefficient representing the “the difference in the process response to a change in the manipulated variable versus a change in a disturbance output.”

For this system, this means the predicted change in reflux flow needed based on an observed change in temperature to keep the top column temperature constant. This was initially estimated using basic thermodynamic calculations for the equipment, but needs refinement to bring the feedback correction as close to zero as possible. By setting up this coefficient as an input on the operating screen and trending it, operators watching the batches can do much of this work. This helps them understand what is happening behind the scenes with control, and they often can predict when a phase change will need to change the coefficient. After values are determined that keep the feedback control from working the majority of the time, program the coefficients into the batch phases.

CONSIDER CONTROL SOFTWARE AS A STARTING POINT

Many control systems now include or can interface directly with auto tuning software. This software can usually provide a variety of useful and automated information that is time consuming to obtain manually. For further information on the benefits and theory behind auto-tuning applications, Control Loop Foundation: Batch and Continuous Processes can be referenced.

For this process, automatic modeling was set up in the tuning software to measure process dead time, gain, and time constants for the slave flow loop any time the set point changed by three percent. These calculations ran automatically over several runs without affecting the process, and produced gain, reset and rate starting points for the entire range of flow. Thankfully, the values did not change greatly across the flow range, so default tuning could be downloaded into the flow controller itself.

Temperature set point changes did not occur as often, so to speed up model collection, broad temperature ramps were altered in the batch logic to a series of small three percent step changes. Each step change then produced a process gain, dead time, and time constant calculation for that section of the range without sacrificing much time on the batch. Instead of one model per batch, many were
produced. The resulting overall model was very dynamic across the range of flows, and was phase dependent. Work is ongoing to fully realize control specifications based off of this information.

FINAL VALIDATION AND RESULTS

At this time, the new instruments selected using the methods above are still on order, but the pre-planning done should ensure a smooth control transition upon arrival. The financial implications of this lack of control are now well understood by plant management, who are eager to see improvement. Their support, along with continued assistance from the ISA Mentorship program, instills confidence that new controls will significantly improve the cycle time for these columns.

REFERENCES