Stripping Vessel/Pot Level Control Automation

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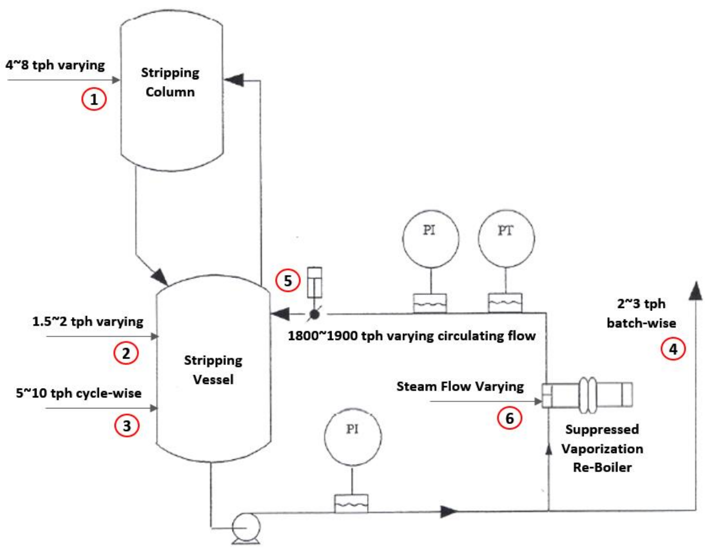
Abstract – This article elaborates briefly the level control automation of a stripping vessel equipped with a suppressed vaporization re-boiler with multiple inlet/outlet streams – some of which feed the system in a sporadic manner batch-wise or cycle-wise – along with a very high recirculating flowrate. Altogether, this can generate a lot of disturbance in the system and a proper level control is indispensable while accounting of process dynamics. Here, a DCS based PI control is discussed which post-implementation resulted in significant process control improvement and almost nullified manual intervention.

INTRODUCTION

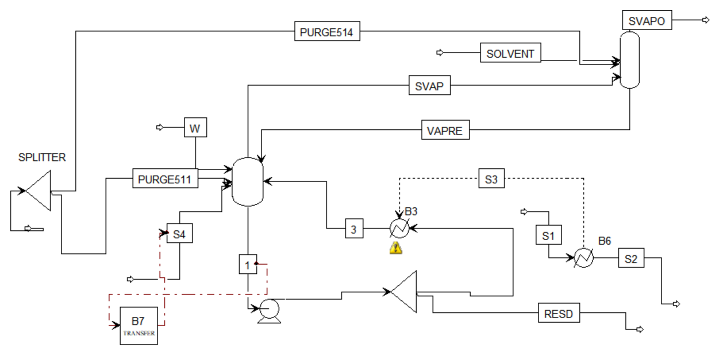
The whole system comprises of a stripping vessel which is fed with a scrubbing liquor stream (25% Acetic Acid, 75% Water) that varies on continuous basis, a purge stream (88% Acetic Acid with remaining Water, Benzoic Acid, Trimellitic Acid, etc.) that is fed batch-wise once after every 10-15 minutes, an outlet stream that is fed to an evaporator that operates batch wise once in every 24-hrs, the vessel also receives a scrubbed stream (rich in Acetic Acid) from an overhead stripping column. The whole vessel content is circulated at a very high nominal flow rate of ~1900 tph via suppressed vaporization re-boiler that vaporizes/flashes the volatile contents at the recirculated liquid inlet route to the stripping vessel.



Saturated steam at 1.8~2 barg is fed to the re-boiler and in view of the aforementioned change in feed inlet/outlet flux, steam demand changes which is manipulated manually as per the change in stripper vessel accumulation and in some cases when coarse control become indispensable, stream: 1 and stream: 2 have to be varied as well.

FLOW VARIATION ON VESSEL'S ACCUMULATION

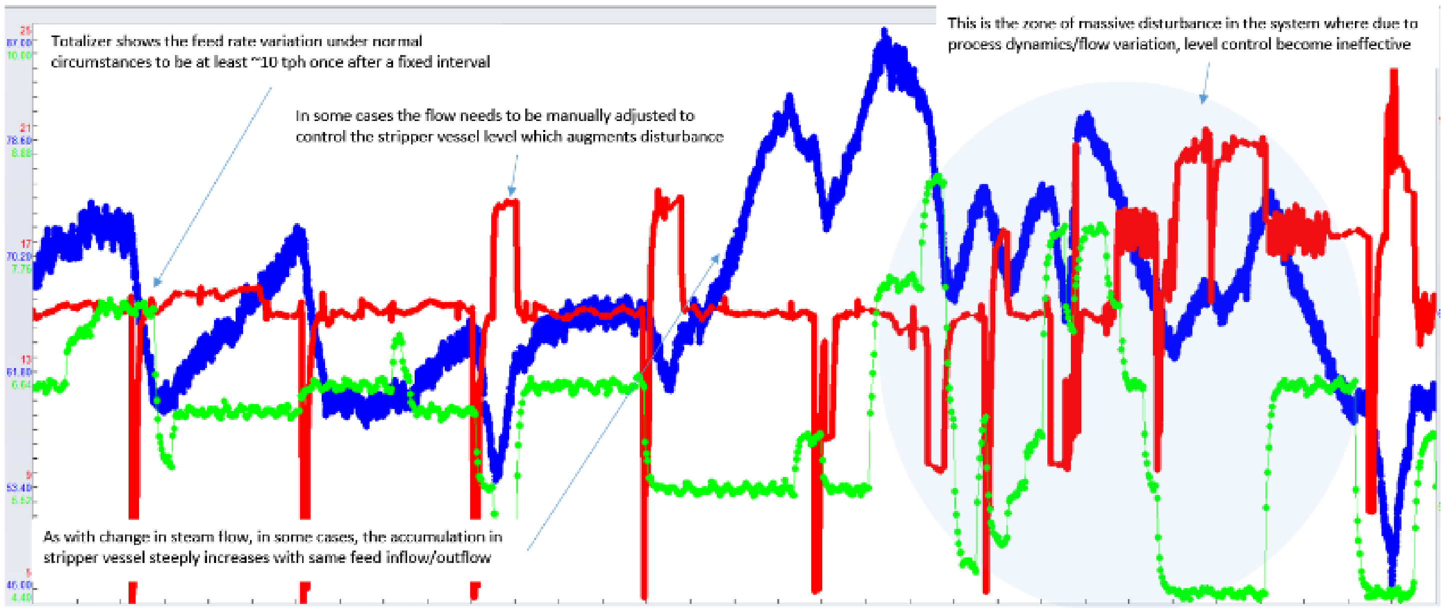
The nominal process data analysis showed that under normal circumstances at least ~10 tph flow variation once after a fixed time interval is evident. Aboveboard, the vessel's accumulation declines with the said aforementioned profile remained consistent, creating further disturbance when an outflow is lined-up batch-wise. A third nuisance is an inadvertently keeping low steam flow in contrast to demand which steeply increases the accumulation and subsequently triggers a chain-effect of cyclic-variations which become very cumbersome to control. All of the flows are controlled by respective independent DCS based master controllers and are not directly linked with change in vessel's level changing profile.

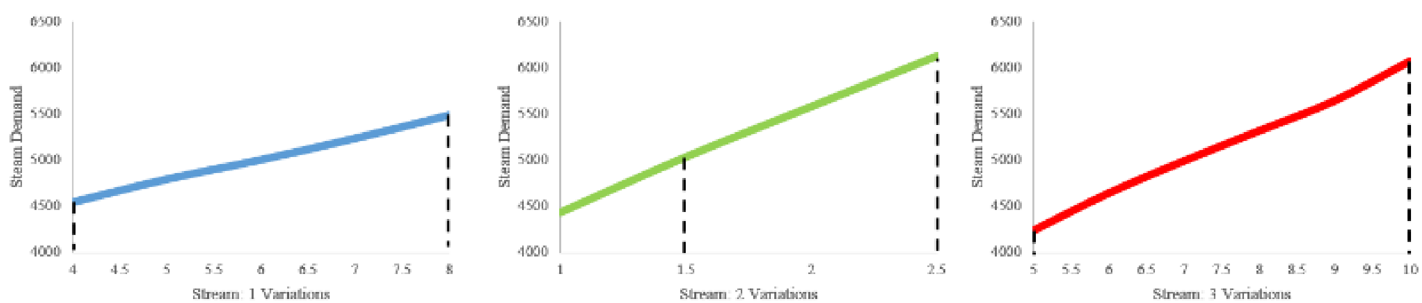


The simulation analysis explains different steam demand profiles depending upon the inlet composition i.e. stream: 2 can bring a change of ~1.1 tph steam demand per unit change of flow rate while for stream: 1, the change in steam demand is least i.e. 0.36 tph per unit change in flow rate.

Stream:1	Stream:2	Stream:3
tes steam/tes flow rate change		
237	1132	367

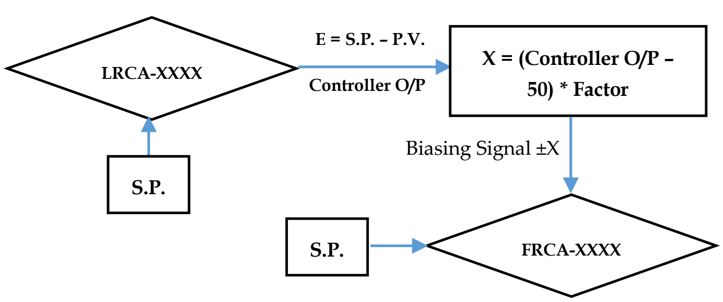
Therefore, accounting for known process dynamics, a constant change of at least 1.7~2 tph steam flow rate is required in order to cope up with the changing flowrates affecting accumulation in vessel. This doesn't account for the fact that when stream: 4 is lined-up then a nominal demand decreases as well by approximately ~0.25 tph. Such complex scenario can't be easily managed with a typical master-slave PID control concept citing level controller that manipulates the inlet/outlet feed flow rates. One master control can't manipulate multiple final control elements at once unless certain tuning weightage(s) are assigned to each slave controls. Despite of that, the control would remain essentially ineffective and a S.P. limiting biasing cascade control would verily serve the purpose effectively.



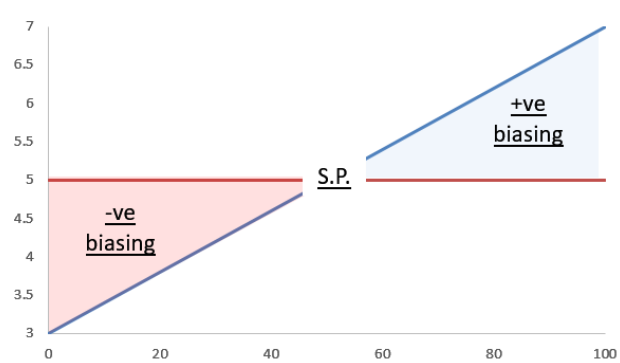


Process Control Logic

The developed cascade control is made simplified to direct the steam demand as per the stripper vessel's level accumulation instead of making flow rates part of it. A level controller is developed which served the role of Master and takes a S.P. followed by generating output based on the Error = S.P. – P.V. which acts as an input signal for a calculation block that divides the signal into three parts: (a) <50% = this is where the calculation block works in a receding way by generating negative values; (b) 50% = this is where S.P. and P.V. equalizes eventually generating a zero command; (c) 50%> = this is where the calculation block generates positive values. These respective values then go for the summation with given P.V. of slave controller which in this case is steam flow rate; therefore, depending upon the accumulation in vessel, the calculation block can do biasing on the fixed given steam S.P.



As from the process dynamics, we have already estimated that the theoretical steam requirement can rise up to ~1.7 tph (round-off to 2 tph) and in the same manner it can decrease as well; therefore, the biasing calculation block is factor adjusted accordingly so against a fixed S.P. it can be self-manipulated.

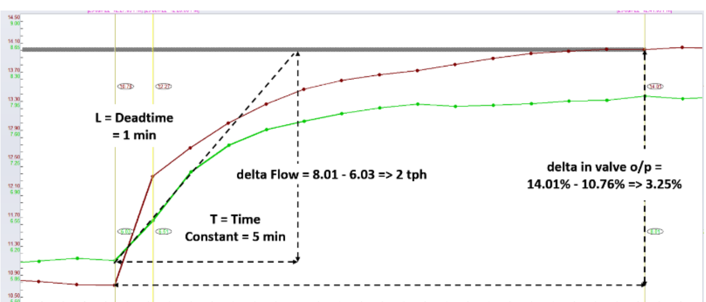


Process Control Tuning

Rule-based tuning is used via Zieglar-Nichols method in order to target less disturbances in amplitude of error among successive cycles against change in S.P. – P.V. since, the newly developed controller doesn't result in any change in physical parameter and it only results in generating output in the range of 0-100%, our main intent is to find its optimal rate of rise/fall. Therefore, for initial estimation purposes, the slave steam controller is manipulated on auto mode and below parameters were estimated as below. We usually do not employ derivative action in controller because it has the tendency to augment error resulting in noisy response especially in industrial application where the mass electrical system is connected/synced with large grids independent of captive power plant which affects electrical frequency.

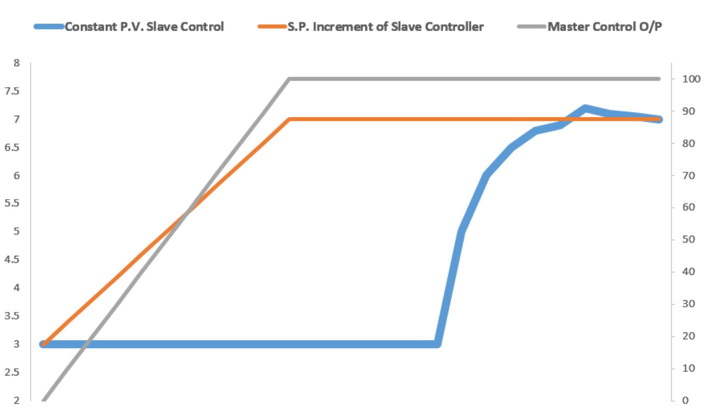
L = Approximate Dead Time = 1 minute
T = Time Constant = 5 minutes
Delta Flow Rate = 8.01 – 6.03 => 2 tph
Delta in Valve O/P = 14.01% - 10.76% => 3.25%
K = Delta Flow Rate/Delta Valve O/P => 0.61538 tph/%
a = K*L/T => 0.12308
For PI based DCS control:
Kp = 0.9/a => 7.31
Ti = 3 * L => 3

The gain and integral actions are updated in the slave controller accordingly and employed in the same manner to master controller based on process dynamics heuristics to observe initial response.



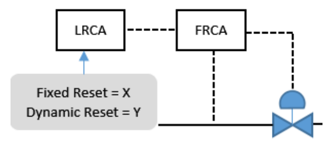
Anti-Reset Windup Control

For the cascade control, anti-reset windup is necessary to incorporate. Modern DCS system has built-in configuration that takes slave controller's PV as an input to the master controller. Considering our case, one of the method is to use a scalar block that scales the flow controller FRCA-XXXX for the range 0% - 100% and feeds back to level controller LRCA-XXXX. This phenomenon is alternatively known as External Reset Windup Control. Alternatively, proceed with biasing around a manually given S.P. so the change in S.P. in entirely governed by the calculation block that sets upper and lower limit as what we have incorporated in our logic (refer to graph above of



+ve/-ve biasing) eliminating the need to using this anti wind-up control insofar but not completely. The intent of this thing is to prevent master controller's integrating action when slave controller reaches a wind-up state. In the last graph, it can be seen that slave controller stuck at 3 tph (assuming some process upset or final control element problem) while the master controller increases to 100% with slave S.P. of 7 tph. After a certain time, the downstream slave controller suddenly responds and takes maximum S.P. causing the P.V. to steeply increase. Though, it doesn't happen invariably but chances are present. In our case, however, we don't see the need to utilize the dynamic external reset because it can slow down the controller's response even with high gain and low reset action i.e. seconds/repeat.

Here, both cases of external fixed/static and dynamic reset will be discussed. Mathematically, it can be explained as:



$$\Delta MVno = \Delta MVn + \Delta L/\tau 1 * (Fr - MVn\text{-}1)$$

Fr is the external reset limit which can be dynamic or static, MV is the controller output, the term (Fr – MVn-1) will become zero in case the external reset limit is not applied, ΔL is the change in master control LRCA.

$$\Delta MVn = K*Ks*(\Delta En + \Delta l/Ti * En),$$

$$En = PVn - SVn,\ \Delta En = En - En\text{-}1$$

In above equation, differential terms are taken zero and not mentioned. En denotes the error between S.P. and P.V. The final derived equation is shown below indicating that the controller output would be essentially dependent on the controller gain (K) multiplied by error and scaling gain (Ks) while external reset rate remained constant.

$$MVn = Fr - K * Ks * En$$

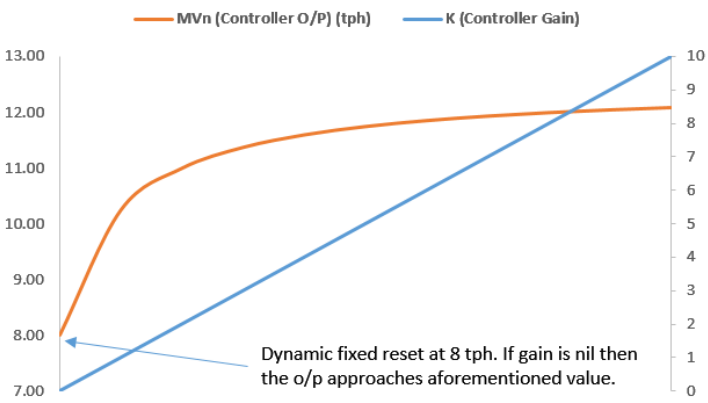
$$MVn = Fr - K * Ks * (PVn - SVn)$$

$$Kp = \text{Process Gain} = PVn/SVn$$

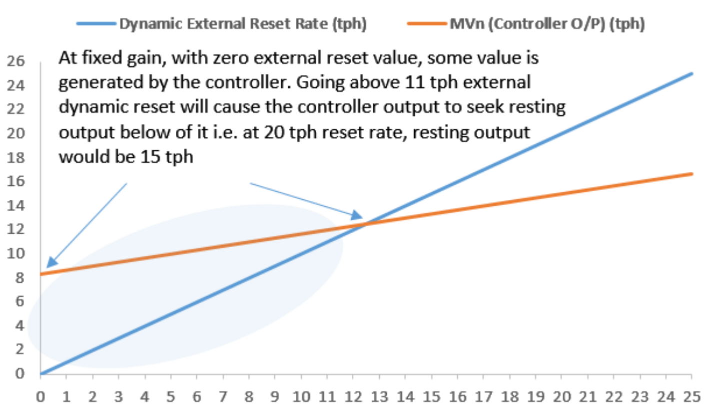
$$MVn = (Fr + K*Ks*SVn)/(1 + K*Ks*Kp)$$

Dynamic Reset Rate	Fr	8	tph
Slave Control	MVrange	25	tph
Master Control	PVrange	100	%
MVrange/PVrange	Ks	0.25	
Set Value Master	SVn	50	%
Process Gain (Assumed Linear)	Kp	4	%/tph

The next graph shows the relationship between varying controller gain and resultant controller response at constant/static external reset value. At low gain, the value essentially approaches its fixed external reset rate and increasing the controller gain will serve no purposes after a certain point. In absence of a biasing control, this signifies that, for the whole scaled 25 tph range, the controller output stabilizes prior instead of integrating further but this is dependent upon slave controller range i.e. like in this case as long as the external reset static rate is less than half of the slave controller range, the relationship goes inverse (the concept if direct/reverse acting.



Below graph depicts the relationship of varying/dynamic external reset rate on controller output at fixed controller's gain. It can be inferred that even at zero external reset rate, some controller output is generated. Going above 11 tph external dynamic reset will cause the controller output to seek resting output below of it i.e. at 20 tph reset rate, resting output would be 15 tph. This is basically a way to describe where the controller output would limit if external reset rate (linked with P.V.) sticks at a particular point. This is again dependent on the scaled flow rate range.



Above two cases are showing two important points: (a) It is utilizing the whole scaled range 0-25 tph which in our case, it is not required as the theoretical limit suggests a mere change of +2 tph/-2 tph is enough for fine control. (b) Utilizing dynamic/static reset would slow down the controller response had the change in P.V. remained minimal while the master controller has a constant error profile.

Conclusion

In our case, the biasing controller with designated limits works way better than utilizing whole flow span after incorporating anti-reset wind-up control. As can be seen in the final trend that minimal biasing above/below S.P. can effectively control the level with minimal or no variation in system. Below trend profile depicts the range value for LRCA (In Dark Brown) which has reduced by at least 7~9%. The steam consumption profile (In Dark Red) has reduced by at least ~0.9~1.2 tph on average basis.