An Optimizing Controller for Crude Oil Blending Operations

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Abstract
This paper presents the application of a “bias-update” optimizing controller to the blending of crude oil. The “bias-update” is carried out based on a non-linear mixing model. The goal is to meet contractual quality (i.e. density) and to minimize raw material costs. The proposed controller is compared in simulation to its linear counterpart. Results show that the non-linear controller meets the quality requirements whilst the linear controller produces degraded blends at a lower cost.

Keywords: crude oil blending, optimizing control

1. Introduction
Blending operations have been recently recognized by the Mexican petroleum industry as an opportunity for applying and demonstrating the benefits of novel techniques for automated operation and control. Sanchez et al. (2003) established that by meeting contractual quality conditions, increments of up to 0.30 USD/bbl could be achieved for a type of crude oil representing 13.6% of national exports. Since crude oil properties may vary considerably, real-time optimizing controllers have been proposed previously for calculating the optimal operating conditions. Forbes et al. (1994) introduced the notion of a “bias update” scheme. Singh et al. (1997) improved the formulation with a non-linear model and including a stochastic model for perturbations. Coordination control has been also used (Chang et al., 1998). Alvarez et al. (2002) studied the “bias update” for gasoline blends, establishing sufficient conditions for stability and convergence. They also showed that this scheme can be interpreted as a feedback linear-integral regulator acting on the modeling error.

This work presents the application of the “bias-update” optimizing controller proposed by Alvarez et al. (2002) to a crude oil blending system, typical of those used in the domestic petroleum industry. The controller takes into account design (max. and min.

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flowrates) and operation (raw materials availability and physical properties). In section 2, the crude oil blending process to be used as an exemplar is described. It is also presented how the process was modeled in order to incorporate the proposed control scheme. Section 3 discusses the controller formulation. Section 4 presents dynamic simulation results comparing the performance of the proposed controller against its linear counterpart. The paper closes in section 5 discussing practical aspects of the proposed scheme.

2. The Crude Oil Blending Process

The layout of a typical crude oil blending process is shown in fig. 1. It is composed of two sections: the blending section and the storage/load section. The blending section is constructed using three types of nodes: blending nodes (identified by the label “blender”), splitting nodes (circles) and separation nodes with storing capacities affecting the composition of the streams. The storage/load section may be composed of a number of storage tanks that feed the tankers. The optimization objective is to establish the combination of input streams with constant flow rate and minimum cost satisfying the volumes and physical properties (in this case, the density) of the shipping program. No other costs are currently considered.

2.1 Model of the storage and load to tankers section

Given the required shipment (volume $V_{\text{exp}}$ and density $\rho_{\text{exp}}$) and initial conditions in storage tanks ($V_{\text{Tk0}}, \rho_{\text{Tk0}}$), the required total volume to complete the shipment and its density are calculated by a steady state mass balance and a nonlinear mixing rule based on excess contributions (Smith and Van Ness, 00):

\begin{align}
\rho_c &= \frac{m_{\text{exp}}^2 \rho_{\text{exp}} - m_{\text{Tk0}} \rho_{\text{Tk0}} (m_{\text{exp}} - \pi m_c)}{m_c (m_{\text{exp}} - \pi m_{\text{Tk0}})} \quad (1) \\
V_c &= \frac{V_{\text{exp}} \rho_{\text{exp}} - V_{\text{Tk0}} \rho_{\text{Tk0}}}{\rho_c} \quad (2)
\end{align}

Figure 1. A typical crude oil blending process.
$m$ represents the crude oil mass and $\pi$ is the experimental parameter of the binary interaction coefficient for the excess properties of the mixing rule. The required flowrate $q_i$ is obtained dividing $V_c$ by the expected loading time.

### 2.2 Model of the blending section

In this process section each stream is modeled as a set of properties:

$$S_x = \{q_x, q_{x_{\text{min}}}, q_{x_{\text{max}}}, \rho_x, c_x\}$$

where

- $q_x, q_{x_{\text{min}}, \text{max}}$ is the nominal, min. and max. flowrate of stream $x$, (kg/h).
- $\rho_x$ is the density of stream $x$, (°API)
- $c_x$ is the cost associated to stream $x$, ($/\text{kg}$)

Refer to fig. 1. for the identification sub-index of each stream. Based on mass balances, expressions are obtained for the output flowrate and density of each type of node. For the blender node these expressions are:

$$q_B = \sum_{k=1}^{n} q_{ink}$$

$$\rho_B = \frac{\sum_{k=1}^{n} \rho_{ink}q_{ink}}{q_B} + \sum_{j=1}^{n} \prod_{k=j+1}^{n} \pi_{j,k}q_{mj}q_{ink} q_{B}$$

For the case of the separator node, only water is considered to be separated. The corresponding expressions are:

$$q_{p2} = \beta q_a$$

$$\rho_{p2} = \frac{\rho_a - \alpha \rho_{p1}}{\beta}$$

Where $\beta$ is the mass fraction ratio of crude oil in the input/output to next stage streams and $\alpha$ is a constant representing the separated water in the dehydrator. In the case of the splitting node, the splitting ratio is given as an operation parameter.

### 3. The Blending Controller

Assuming that there is no mass accumulation in the blenders, the discrete-time formulation of the optimizing controller proposed by Alvarez et al. (2002) is given by:

$$\min_{q_{\text{in}}} c^T q_{\text{ink}} \quad k \geq 1$$

s.t.
• Input flow availability \[ q_{\text{in min}} \leq q_{\text{in}} \leq q_{\text{in max}} \]  
\[ (8) \]

• Mass balance in nodes \[ \sum_{p=1}^{n} q_{(in,p)k} = q_B \]  
\[ (9) \]

• Density of blend \[ \rho_{B\min} \leq \frac{P_{\text{in}}^T q_{\text{in}k}}{q_B} + \eta_k \leq \rho_{B\max} \]  
\[ (10) \]

with \[ \eta_k = \rho_{B,k-1} - \frac{P_{\text{in}}^T q_{\text{in},k-1}}{q_B}, \quad \eta_0 = 0 \]  
\[ (11) \]

\( \eta_k \) is the blending modeling error at sampling step \( k \). Density \( \rho_{\text{in}} \) is the nominal value of \( \rho_{\text{in}} \). The blend density \( \rho_{B,k} \) is measured at each sampling interval. The optimization problem is thus solved using a one-step delayed estimation of the blend properties.

### 3.1. The non-linear blending model

Using the equations of flowrate and density for each node in fig. 1, the following expressions are found for the desired flowrate \( q_c \) and density \( \rho_c \):

\[ q_c = \beta(q_{in1,k} + q_{in2,k}) + q_{in3,k} + q_{in4,k} - q_{p3} \]

\[ \rho_{c,k} = \frac{(P_{in1} - \alpha \rho_{p1}) q_{in1,k} + (P_{in2} - \alpha \rho_{p1}) q_{in2,k} + P_{in3} q_{in3,k} + P_{in4} q_{in4,k} + \Theta_k}{q_c} \]

where \( \Theta_k \) is the non-linear part of the blending model.

\[ \Theta_k = \left( \frac{\pi q_{in1,k} q_{in2,k}}{q_v} + \frac{\pi q_{p2} q_{in3,k}}{q_b} \left( 1 - \frac{q_{p1}}{q_b} \right) \right) - \left( (P_{in1} - \alpha \rho_{p1}) q_{in1,k} + (P_{in2} - \alpha \rho_{p1}) q_{in2,k} + P_{in3} q_{in3,k} + P_{in4} q_{in4,k} \right) \frac{q_{p1}}{q_b} + \left( \frac{\pi q_{p2} q_{in4,k}}{q_c} \right) \]

with \( \alpha \) and \( \beta \) as the dehydration constants of eqns. (5) and (6). With this information and assuming constant flow rate, eqns. (9) and (10) of the optimal controller formulation take the following form:

\[ \beta(q_{in1,k} + q_{in2,k}) + q_{in3,k} + q_{in4,k} = q_c + q_{p3} \]

\[ \rho_{\text{in min}} \leq \frac{(P_{in1} - \alpha \rho_{p1}) q_{in1,k} + (P_{in2} - \alpha \rho_{p1}) q_{in2,k} + P_{in3} q_{in3,k} + P_{in4} q_{in4,k} + \psi_k}{q_c} \leq \rho_{\text{in max}} \]

where \( k \) is the time period and \( \psi_k \) is the non-linear bias update term obtained as:
\[ \psi_k = \rho_{c,k+1} - \frac{(\rho_{c1} - \alpha \rho_{\text{in}1})q_{\text{in}1,k} + (\rho_{c2} - \alpha \rho_{\text{in}2})q_{\text{in}2,k} + \rho_{\text{in}3}q_{\text{in}3,k} + \rho_{\text{in}4}q_{\text{in}4,k}}{q_c}, \quad \psi_i = 0 \]

The optimization problem was solved with Matlab’s linprog routine.

4. Simulation Results

Consider the following example. The shipping requirements are \( V_{\text{exp}} = 350,000 \text{ m}^3 \) and \( \rho_{\text{exp}} = 32.2^\circ\text{API} \). The initial conditions in the storage tanks are \( V_{\text{tko}} = 100,000 \text{ m}^3 \) and \( \rho_{\text{tko}} = 32.9^\circ\text{API} \). The input stream properties are given in table 1.

<table>
<thead>
<tr>
<th></th>
<th>S_1</th>
<th>S_2</th>
<th>S_3</th>
<th>S_4</th>
</tr>
</thead>
<tbody>
<tr>
<td>( q_{\text{in},\text{min}} ) (kg/h)</td>
<td>100</td>
<td>100</td>
<td>100</td>
<td>100</td>
</tr>
<tr>
<td>( q_{\text{in},\text{max}} ) (kg/h)</td>
<td>644,000</td>
<td>280,000</td>
<td>180,000</td>
<td>273,000</td>
</tr>
<tr>
<td>( \rho_{\text{in}} ) (°API)</td>
<td>32.2</td>
<td>33.5</td>
<td>21.8</td>
<td>32.8</td>
</tr>
<tr>
<td>( c_{\text{in}} ) ($/kg)</td>
<td>0.1819</td>
<td>0.1833</td>
<td>0.1703</td>
<td>0.1826</td>
</tr>
</tbody>
</table>

Solving eqns. (1) and (2), the blending requirements are \( q_c = 242,200 \text{ kg/h} \) and \( \rho_c = \rho_{c,\text{min}} = 32.29^\circ\text{API} \). The system behavior and controller response are shown for variations in the density of input streams: -5% in \( \rho_{\text{in}1} \) and -4% in \( \rho_{\text{in}3} \) at \( t=12 \text{ h} \); -7.5% and -10% in \( \rho_{\text{in}1} \) and \( \rho_{\text{in}2} \) respectively at \( t=22 \text{ h} \). Figure 2 shows how the non-linear controller meets the contractual requirements while the linear controller drives the system to an off-spec density. It is interesting to note that although the linear controller does not meet the density requirements, it produces a cheaper blend than the non-linear controller as shown in figure 3. Two important points are the instantaneous value of the blend (figure 3) and the considerable change of input flowrates (figure 4) due to changes in the inlet stream properties. In practice, this may lead to challenging operational conditions.

Figure 2. Blend density behavior.  
Figure 3. Blend cost.
5. Conclusions

The proposed controller achieves the production requirements, whilst the linear controller fails short in quality but with cheaper mixtures. This opens the door for interesting trade-off considerations in establishing contractual conditions. Other issues to be discussed will be the controller robustness and the effect of the operation parameters in the solution hyper region.

Figure 4. Input flow rate behavior

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