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DESIGN OF AN OPTIMIZING CONTROL STRATEGY FOR CRUDE OIL BLENDING OPERATIONS

DISEÑO DE UNA ESTRATEGIA DE CONTROL OPTIMIZANTE PARA OPERACIONES DE MEZCLADO DE PETROLEO CRUDO

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Abstract

This paper presents the design of a control strategy for blending of crude oil. The strategy is based on the application of a non-linear "*bias-update*" optimizing controller to a multi-stage blending process. The goal is to meet contractual quality (i.e. density) and to minimize raw material costs. The proposed control strategy 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.

Keywords: optimizing control, bias-update, crude oil blending.

Resumen

Se presenta el diseño de una estrategia de control para un sistema multi-etapa de mezclado de petróleo crudo. La estrategia está basada en la aplicación de un control optimizante con "actualización de sesgo". El objetivo es satisfacer las condiciones de calidad contractual (en este caso, la densidad de la mezcla final) minimizando el costo de materia prima. La estrategia propuesta es comparada en simulación con su contraparte lineal. Los resultados muestran que el considerar la actualización en el sesgo permite que el controlador no lineal logre cumplir con los requerimientos de calidad, mientras que el controlador lineal lleva al sistema a producir una mezcla con una calidad degradada.

Palabras clave: control optimizante, actualización de sesgo, mezclado de petróleo crudo.

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. Sánchez *et al.* (2003) established that, by complying with 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 to compensate for model mismatches. Singh *et al.* (1997) improved the formulation with a non-lineal 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.

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and min. 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 to be considered in future works.

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 operates in a continuous mode and is constructed using three types of nodes: blending nodes (identified by the label "blender") with two inputs and one output, splitting nodes (circles) and separation nodes affecting the composition of the streams with one input and two outputs. No recycles are considered. The storage/load section operates in batch mode. It is composed of a number of storage tanks that receive the blended crude oil and 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 Non-linear model of density

The non-linear behavior of the density in crude oil blends is captured using the theory of excess contributions (Smith and Van Ness, 2000). The density is modeled as the arithmetic average of crude densities plus a non-linear contribution.

$$\rho_M(t) = \sum_{i=1}^{m} \frac{f_i(t)\rho_i}{f_M} + \sum_{j=1}^{m-1} \prod_{i=j+1}^{m} \frac{\pi_{i,j} f_i(t) f_j(t)}{f^2{}_M} \qquad (1)$$

The term $\pi_{i,j}$ is the interaction coefficient between components *i* and *j* and is obtained empirically as a function of the densities and fractions of the components (Sanchez *et al*., 2003). Even when crude oils have similar compositions, the non-ideal contribution may have an important impact for control purposes as it is shown in this work.

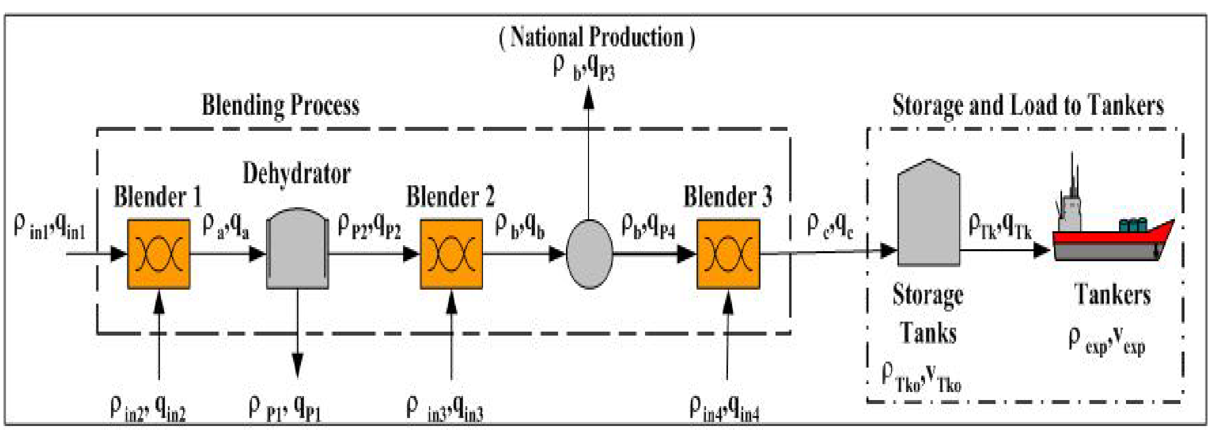

Fig. 1. A typical crude oil blending process.

2.2 Model of the storage and load to tankers section

Given the required shipment (volume $V_{\exp}$ and density $\rho_{\exp}$) and initial conditions in storage tanks (V_{Tko}, ρ_{Tko}), the required total volume (V_c) to complete the shipment and its density (ρ_c) are calculated by a steady state mass balance around the storage/load section and the nonlinear mixing rule (1):

$$\rho_c = \frac{m_{\exp}^2\,\rho_{\exp} - m_{Tko}\,\rho_{Tko}(m_{\exp} - \pi\,m_c)}{m_c(m_{\exp} - \pi\,m_{Tko})} \quad (2)$$

$$V_c = \frac{V_{\exp}\,\rho_{\exp} - V_{Tko}\,\rho_{Tko}}{\rho_c} \quad (3)$$

Where *m* represents the crude oil mass and π is the experimental parameter of the binary interaction coefficient for the excess properties of the mixing rule. The required flowrate q_c is obtained dividing V_c by the expected loading time.

2.3 Model of the blending section

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

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

where

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

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

$$q_{\mathrm{B}} = \sum_{k=1}^{n} q_{ink} \quad (4)$$

$$\rho_{\mathrm{B}} = \sum_{k=1}^{n} \frac{\rho_{ink}\,q_{ink}}{q_{\mathrm{B}}} + \sum_{j=1}^{n-1} \prod_{k=j+1}^{n} \frac{\pi_{j,k}\,q_{inj}\,q_{ink}}{q_{\mathrm{B}}^2} \quad (5)$$

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

$$q_{P2} = \beta\, q_a \quad (6)$$

$$\rho_{P2} = \frac{\rho_a - \alpha\rho_{P1}}{\beta} \quad (7)$$

Where β is the output/input mass fraction ratio of crude oil and α 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, the discrete-time formulation of the optimizing controller proposed by Alvarez *et al.* (2002) for a blending node is given by:

$$\min_{\mathbf{q_{in}}} \ \mathbf{c}^T \mathbf{q}_{\mathbf{in}k} \quad k \geq 1 \quad (8)$$

s.t.

- Input flow availability

$$\mathbf{q}_{\mathbf{in}_{\min}} \leq \mathbf{q}_{\mathbf{in}k} \leq \mathbf{q}_{\mathbf{in}\max} \quad (9)$$

- Mass balance in node

$$\sum_{p=1}^{n} q_{(in,p)k} = q_B \tag{10}$$

- Density of blend

$$\rho_{B\min} \leq \frac{\bar{\boldsymbol{\rho}}_{\mathbf{in}}^{T}\mathbf{q}_{\mathbf{in}k}}{q_B} + \eta_k \leq \rho_{B\max} \tag{11}$$

with

$$\eta_k = \rho_{B,k-1} - \frac{\bar{\boldsymbol{\rho}}_{\mathbf{in}}^{T}\mathbf{q}_{\mathbf{in},k-1}}{q_B}, \quad \eta_1 = 0 \tag{12}$$

η_k is the estimated blending modeling error at sampling step k (i.e. bias-update). Density vector $\bar{\boldsymbol{\rho}}_{\mathbf{in}}$ is the nominal value of vector $\boldsymbol{\rho}_{\mathbf{in}}$. 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. Alvarez *et al*., (2000) applied this controller to the blending of gasolines and they found that the numerical sensitivity was highly dependant to the problem being solved. In our case, even when the nonlinearities are of smaller magnitude, the sensitivity is still an important issue. This is illustrated with the following two-dimensional example. The input streams are S_1 = {2.0, 9.0, 33.5, 0.1833} and S_2 = {1.0, 8.0, 30.0, 0.1795}. The desired output flow from the blender is $q_{out} = q_{out_{\min}} = q_{out_{\max}}$ = 10.0, with ρ_{min} = 31.5 and ρ_{max} = 33.0. Thus, the optimization problem to solve is:

$$min\ [0.1833\ q1_k + 0.1795\ q2_k] \tag{13}$$

s.t.

$$[2.0\ 1.0] \leq [q1_k\ q2k] \leq [9.0\ 8.0] \tag{14}$$

$$q1_k + q2_k = 10.0 \tag{15}$$

$$31.5 \leq \frac{33.5 f1_k + 30.0 f2_k}{10.0} + \eta_k \leq 33.0 \tag{16}$$

The operation time is 10 h. The density measurement is taken every hour. Using q_{1_k} and q_{2_k} as axes of a two-dimensional graph, the non-linear and linear solutions for k = 10 are shown in Fig. 2 and Fig. 3 respectively.

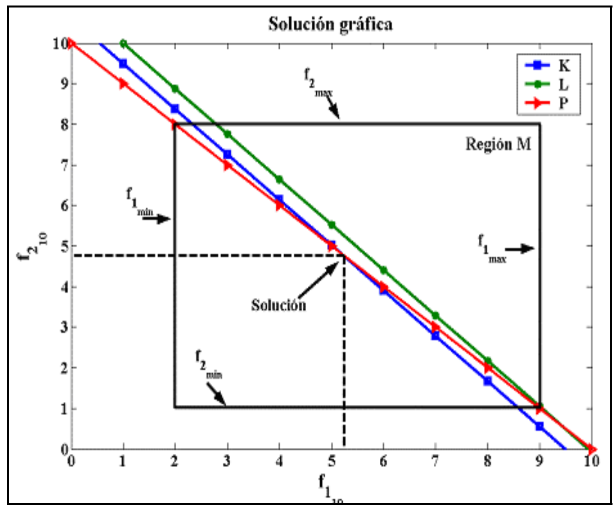

Fig. 2. Solution with η = -0.2.

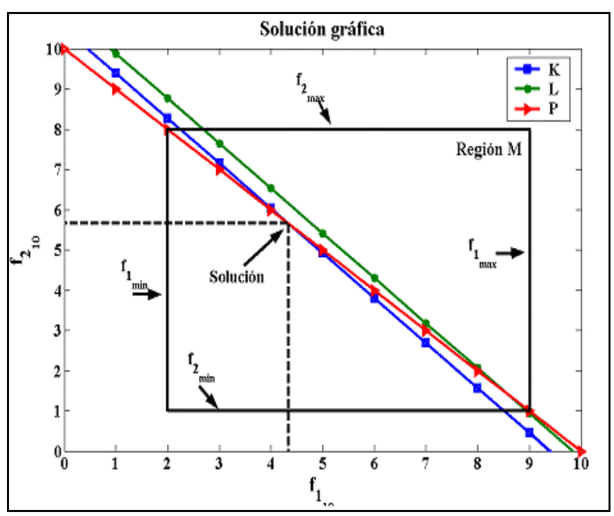

Fig. 3. Solution with η = 0.

The input flow restriction, Eq. 12, is given by region **M**. The mass balance, Eq. 15, and minimum and maximum density restrictions based on Eq. 15 are shown by lines **P**, **K** and **L** defined as follows.

$$\mathbf{M} := \{\mathbf{f}_k \in \Re^2 \; | \mathbf{f}_{\min} \le \mathbf{f}_k \le \mathbf{f}_{\max},\ \mathbf{f}_k \ge 0\} \quad (17)$$

$$\mathbf{P} := f_M - f1_k = f2_k \quad (18)$$

$$\mathbf{K} := \frac{f_M(\rho_{M\min} - \eta_k) - f1_k\,\bar{\rho}_1}{\bar{\rho}_2} = f2_k \quad (19)$$

$$\mathbf{L} := \frac{f_M(\rho_{M\max} - \eta_k) - f1_k\,\bar{\rho}_1}{\bar{\rho}_2} = f2_k \quad (20)$$

The nonlinear solution is given by the intersection of lines K and P, i.e. $[q_{1_{10}}\ q_{2_{10}}]$ = [5.2 4.8], in fig. 2 with ρ_{10} = 31.50, blend cost c_M = 1.81 and η_{10} = -0.20. Its linear counterpart, i.e. $[q_{1_{10}}\ q_{2_{10}}]$ = [4.3 5.8] and ρ_{10} = 31.19, is shown in fig.3. Since the distance between lines K and L are very narrow, a small change in the calculation of the non-linear contribution may give rise to a very different result.

3.1. The non-linear blending formulation

Using the mass balance equations type and the non linear mixing rule for each node in Fig. 1, the following expressions are found for the desired flowrate q_c and density ρ_c:

$$q_c = \beta\left(q_{in1,k} + q_{in2,k}\right) + q_{in3,k} + q_{in4,k} - q_{P3} \quad (21)$$

$$\rho_{c,k} = \frac{(\bar{\rho}_{in1} - \alpha\rho_{P1})q_{in1,k} + (\bar{\rho}_{in2} - \alpha\rho_{P1})q_{in2,k}}{q_c} + \frac{\bar{\rho}_{in3}q_{in3,k} + \bar{\rho}_{in4}q_{in4,k} + \Theta_k}{q_c} \quad (22)$$

where Θ_k is the non-linear part of the blending model.

$$\Theta_k = \left(\frac{\pi_1 q_{in1,k} q_{in2,k}}{q_a} + \frac{\pi_2 q_{P2} q_{in3,k}}{q_b}\right)\left(1 - \frac{q_{P3}}{q_b}\right) - \left(\begin{array}{l}(\bar{\rho}_{in1} - \alpha\rho_{P1})q_{in1,k} + (\bar{\rho}_{in2} - \alpha\rho_{P1})q_{in2,k} + \\ \bar{\rho}_{in3}q_{in3,k}\end{array}\right)\frac{q_{P3}}{q_b} + \left(\frac{\pi_3 q_{P4} q_{in4,k}}{q_c}\right) \quad (23)$$

With this information and assuming constant flow rate, Ecs. 10 and 11 of the optimal controller formulation take the following form:

$$\beta\left(q_{in1,k} + q_{in2,k}\right) + q_{in3,k} + q_{in4,k} = q_c + q_{P3} \quad (24)$$

$$\rho_{c_{\min}} \le \frac{(\bar{\rho}_{in1} - \alpha\rho_{P1})q_{in1,k} + (\bar{\rho}_{in2} - \alpha\rho_{P1})q_{in2,k} +}{q_c} \frac{\bar{\rho}_{in3}q_{in3,k} + \bar{\rho}_{in4}q_{in4,k}}{q_c} + \psi_k \le \rho_{c_{\max}} \quad (25)$$

where k is the time period and ψ_k is the non-linear *bias update* term obtained as:

$$\psi_k = \rho_{c,k-1} - \frac{(\bar{\rho}_{in1} - \alpha\rho_{P1})q_{in1,k-1} + (\bar{\rho}_{in2} - \alpha\rho_{P1})q_{in2,k-1} +}{q_c} \frac{\bar{\rho}_{in3}q_{in3,k-1} + \bar{\rho}_{in4},q_{in4,k-1}}{q_c}, \quad \psi_1 = 0$$

The optimization problem was solved with Matlab's *linprog* routine.

4. Simulation results

Considering the formulation presented in section 3, the following data taken form real operation is employed to compare performances of the non-linear ($\eta \ne 0$) and linear ($\eta = 0$) controllers. The shipping requirements are $V_{\exp} = 350{,}000\,\mathrm{m}^3$ and $\rho_{\exp} = 32.2\,°\mathrm{API}$. The initial conditions in the storage tanks are $V_{\mathrm{Tko}} = 100{,}000\,\mathrm{m}^3$ and $\rho_{\mathrm{Tko}} = 32.9\,°\mathrm{API}$. The input stream properties are given in table 1. With $\alpha = 0.0915$, $\beta = 0.9085$, and $\pi = $ -1.302 .

Solving Eq. 1 and 2, the blending requirements are $q_c = 242{,}200\,\text{kg/h}$ and $\rho_c = \rho_{c,\min} = 32.29\,°\text{API}$. The system behavior and controller response are shown for variations in the density of input streams: -5% in ρ_{in4} and -4% in ρ_{in3} at t=12 h; -7.5% and -10% in ρ_{in1} and ρ_{in2} respectively at t=22 h. Fig. 4 shows how the non-linear controller meets the contractual requirements while the linear controller drives the system to an off-spec density.

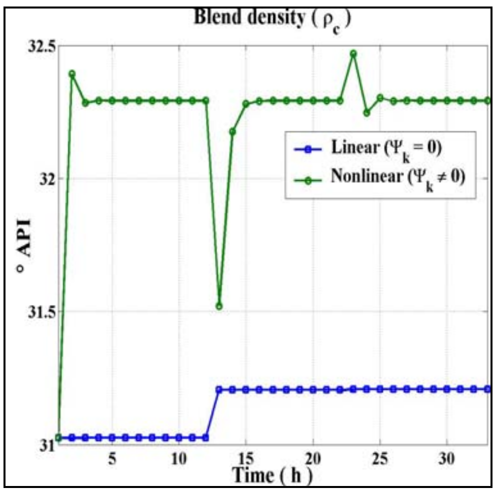

Fig. 4. Blend density behavior.

Table 1. Input streams properties.

	S_1	S_2	S_3	S_4
$q_{in,min}$ (kg / h)	100	100	100	100
$q_{in,max}$ (kg / h)	644,000	280,000	180,000	273,000
ρ_{in} °API	32.2	33.5	21.8	32.8
c_{in} ($ / kg)	0.1819	0.1833	0.1703	0.1826

It is interesting to note that although the linear controller does no meet the density requirements, it produces a cheaper blend than the non-linear controller as shown in Fig. 5. Two important points are the instantaneous value of the blend and the considerable change of input flowrates, Fig. 6, due to changes in the inlet stream properties. In practice, this may lead to challenging operational conditions.

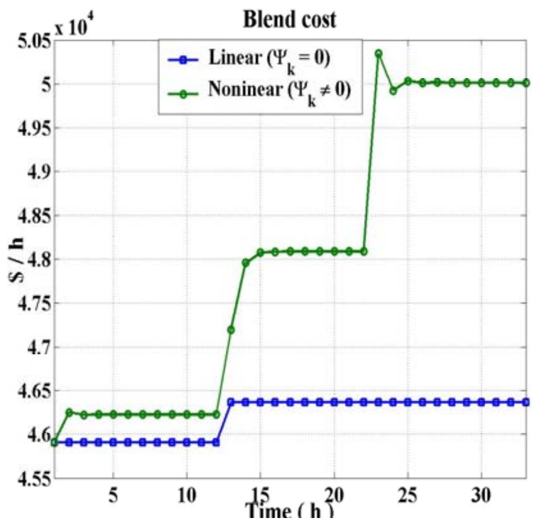

Fig. 5. Blend cost.

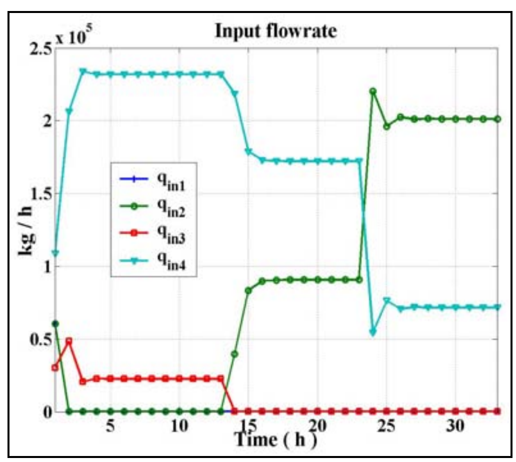

Fig. 6. Input flow rate behavior for the non-linear controller.

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. The next step in this work is to considering the control of the dehydration process and the measuring of the required flowrates. Since measuring large amounts of crude oil is very expensive, it is currently carried out only at delivery points. In order to implement this controller strategy, estimation techniques will be taken on board. This will open interesting issues regarding controller robustness and the effect of the operation parameters in the solution hyper region.

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