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# MODELING OF CONTAMINATION IN PIPELINE BATCH TRANSFERS

**Alexandre Alvim Gomes (Graduate student)**

Universidade Federal Fluminense, Department of Mechanical Engineering - 24210-240 - Niterói - RJ - Brazil

[alexandre@mec.uff.br](mailto:alexandre@mec.uff.br)

**Felipe Bastos de Freitas Rachid**

Universidade Federal Fluminense, Department of Mechanical Engineering - 24210-240 - Niterói - RJ - Brazil

[rachid@mec.uff.br](mailto:rachid@mec.uff.br)

**José Henrique Carneiro de Araujo**

Universidade Federal Fluminense, Department of Computer Science - 24210-240 - Niterói - RJ - Brazil

[jhca@dcc.ic.uff.br](mailto:jhca@dcc.ic.uff.br)

**Abstract.** *In this paper a model for predicting the contaminated mixing volume arising in pipeline batch transfers is presented. The proposed technique represents an improvement over the existing methods since it takes into account time-dependent flow rates and accurate concentration-varying axial dispersion coefficients. After validating the model by comparing its predictions with a field test and other classical procedures, numerical simulations carried out have shown that the mixing phenomenon takes place continuously during the transfer with a decreasing time rate.*

**Keywords.** *mixing volume evolution, batch transfer, pipeline.*

## 1. Introduction

The batch transfer of petroleum products in long pipelines is a current practice all over the world. Unless mechanical separators such as scrapers or pigs are employed (Hara et al, 1979), there will be a certain amount of mixing between products which is called interface contamination. Since the use of separators requires a somewhat complex operation, it is in general avoided, especially when intermediate pump-station manifolds are present.

From the operational viewpoint, the occurrence of contamination implies additional costs associated with shipping the mixture back to refinery for later reprocessing. Thus, the minimization of interface contamination should always be pursued.

During the planning and execution of batch transfers, the interface contamination analysis is carried out in order to establish the best sequence to pump the fluids (to optimize the program of the product sequence aiming to minimize the mixing zones) as well as to control and monitor the spread of the mixing zone, as it travels along the pipe (Benke and Gravert, 1989).

When the specific weights of the products are similar, the contamination process takes place mainly due to turbulent dispersion of matter and has been currently estimated by a number of models (Aunicky, 1970; Austin and Palfrey, 1964; Levenspiel, 1958; Ovádi and Török, 1977; Sjenitzer, 1958; Smith and Schulze, 1948; Maron and Galianov, 1970; Botros, 1984; Netchval et al, 1972) of semi-empirical nature. However, the majority of these models is based on several simplifying assumptions that in some cases seem to be inadequate. As a consequence, they do not give satisfactory results as would be expected.

This paper presents a new methodology to evaluate the interface contamination which takes into account the variation of the flow rate during the transfer and an accurate concentration-dependent axial dispersion coefficient. A comparison with experimental data shows the proposed technique is the most accurate in predicting contaminated mixing volumes among several other methods used in the literature: (Sjenitzer, 1958),

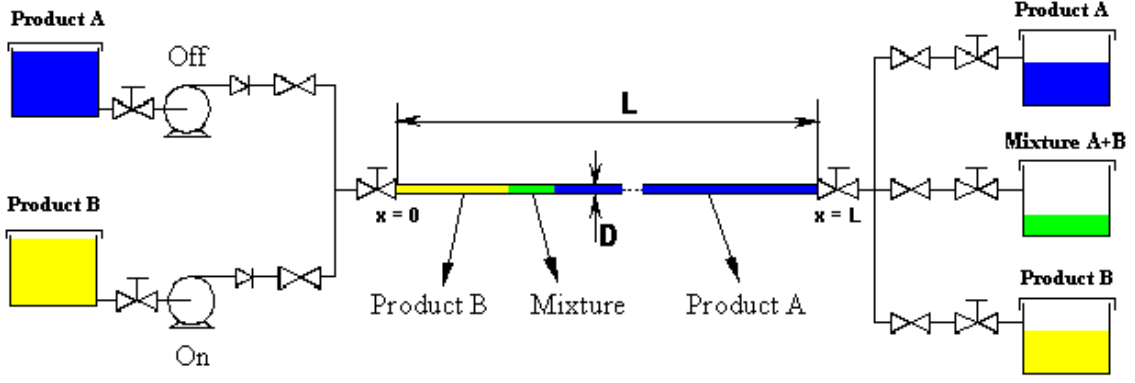


Figure 1: Batch transfer schematic

(Austin and Palfrey, 1964), (Levenspiel, 1958), (Ovádi and Török, 1977), (Smith and Schulze, 1948; Smith and Schulze, 1948), (Aunicky, 1970) and (Netchval et al, 1972). After its validation the model is employed to investigate the way the mixing volume evolution takes place as the transfer is processed.

## 2. Problem Statement

Consider the schematic pipeline installation shown in Fig 1 which is used to sequentially pump two distinct products designated as “A” and “B”. The pipeline has a constant diameter  $D$  and a length  $L$ , which is measured from the junction of the pump discharge lines ( $x = 0$ ) until the receiving point at the other pipeline end. By an appropriate valve switching, at time instant  $t = 0$ , the pumping of fluid “A” is interrupted and, at the same time, pumping of fluid “B” is started so that fluid “B” begins to push fluid “A” along the line. Let  $C_i(x, t) \in [0, 1]$ , with  $i \in \{A, B\}$ , be the time-averaged mean concentration of fluid  $i$  within the mixture at the cross-section of the pipeline at an axial coordinate  $x \in [0, L]$  and time  $t$ . In mathematical form,  $C_i(x, t)$  is defined as

$$C_i(x, t) := \frac{1}{2\Delta t A} \int_A \left[ \int_{t-\Delta t}^{t+\Delta t} \widehat{C}_i(x, r, \theta, \tau) d\tau \right] r dr d\theta \quad (1)$$

in which  $r$  and  $\theta$  stand for the radial and circumferential coordinates,  $A$  is the internal cross-sectional area of the pipe and  $2\Delta t$  is a suitable time interval sufficiently long to cut out turbulent fluctuations and sufficiently short to capture unsteady changes in the concentration field  $\widehat{C}_i(x, r, \theta, t)$ .

Based on the past descriptions, the beginning of the sequential transfer can be described in terms of  $C_B$  by,

$$C_B(x = 0, t = 0) = 1 \text{ and } C_B(0 < x \leq L, t = 0) = 0 \quad (2)$$

or in terms of  $C_A$  by,

$$C_A(x = 0, t = 0) = 0 \text{ and } C_A(0 < x \leq L, t = 0) = 1 \quad (3)$$

since, for all  $(x, t)$ , the following relationship must hold:

$$C_A + C_B = 1. \quad (4)$$

During the passage of products through the pipeline a mixing zone is formed at the boundary of the two adjacent products. Such a zone of contaminated material increases in length as it travels along the line and can be delimited by the interval  $[x_f, x_i] \subset [0, L]$ , with  $x_f = x_f(t)$  and  $x_i = x_i(t)$ , such that for all  $x \in (x_f, x_i)$ ,  $1 = C_B(x = x_f, t) > C_B(x, t) > C_B(x = x_i, t) = 0$  or  $0 = C_A(x = x_f, t) < C_A(x, t) < C_A(x = x_i, t) = 1$ . When the beginning of the mixing zone reaches the pipeline end, that is  $x_i(t) = L$ , the flow is directed towards the mixture tank. Finally, when the end of the mixture zone reaches the pipeline end ( $x_f(t) = L$ ), the fluid flow is conducted to the tank containing product “B”. In practice, small degrees of contamination are allowed to take place at the beginning and at the end of the mixing zone, which are expressed in terms of the maximum admissible concentration  $C_{BA}$  and  $C_{AB}$ , respectively. Once these values (which are not necessarily equal) have been chosen in such a way that the products technical specifications are not altered, the contaminated volume, which should be stored in the tank for later reprocessing, is defined as:

$$V_c(x = L) = \int_{t_{BA}}^{t_{AB}} \widehat{Q}(x = L, t) dt \quad (5)$$

in which  $\widehat{Q}(x = L, t)$  denotes the volumetric flow rate at the receiving point (that is, the position  $x = L$ ),  $t_{BA}$  is the time instant associated with the arrival of the beginning of the contaminated zone with concentration  $C_{BA}$  at  $x = L$  and  $t_{AB}$  is the time instant associated with the arrival of the end of the contaminated zone with concentration  $C_{AB}$  at  $x = L$ . In other words,  $t_{BA}$  and  $t_{AB}$  are such that  $C_B(x = L, t = t_{BA}) = C_{BA}$  and  $C_A(x = L, t = t_{AB}) = C_{AB}$ , respectively.

### 3. Theoretical Formulation

It has been shown by several investigators (Taylor, 1954; Tichacek et al, 1957; Atesmen et al, 1971) that, under the assumptions of axisymmetric and incompressible developed turbulent fluid flow, the intermixing phenomenon of products "B" and "A" sequentially transported in a pipeline can be described for long-dispersion times by the following convection-diffusion equation,

$$\frac{\partial C_i}{\partial t} + u(t) \frac{\partial C_i}{\partial x} = \frac{\partial}{\partial x} [\mathcal{K}(C_i) \frac{\partial C_i}{\partial x}] \text{ for } i \in \{A, B\} \quad (6)$$

in which  $u(t)$  is the time-dependent bulk-average velocity and  $\mathcal{K}(C_i)$  is the effective axial dispersion coefficient which accounts for the relative motion of the fluid flow with respect to the bulk-average velocity as well as for molecular and eddy diffusion in the axial direction.

Equation (6) is the basis for a number of models available in the literature used to estimate mixing volumes, which consider as additional simplifying assumptions that  $\mathcal{K}$  does not depend on  $C_i$  (the products have similar kinematic viscosities) and that  $u$  is constant. However, in some practical cases, these assumptions seem to be inadequate what in turn may endanger the accuracy of the models' predictions.

During a batch transfer, different stretches of the pipeline will be subjected to varying hydraulic conditions inducing in this way a slightly time-dependent flow rate. By assuming that the inertia of the liquids is negligible, the specific weights  $\gamma_i$ , for  $i \in \{A, B\}$ , do not differ significantly from each other, the flow regime remains the same along the whole pipeline, the suction line of the pump is short, minor hydraulic losses can be neglected in the pipe system and, finally, that the mixture process has little effect on the overall balance of linear momentum, such a principle can be stated for any time instant as:

$$P_0 - \left[ \frac{f_B x_m + f_A (L - x_m)}{D} \right] \frac{u^2}{2g} - \Delta z \gamma_A - P_L = 0 \quad (7)$$

in which

$$\frac{dx_m}{dt} = u. \quad (8)$$

In the above equations  $x_m(t)$  stands for the current spatial position of the conventional half-length mixture (that is, the material coordinate in which  $C_B(x = x_m(t), t) = C_A(x = x_m(t), t) = 0.5$ ),  $P_0$  and  $P_L$  represent, respectively, the pressures at the pump discharge ( $x = 0$ ) and at the receiving point ( $x = L$ ),  $\Delta z$  is the topographical level difference between the pipeline positions  $x = L$  and  $x = 0$ ,  $g$  is the gravitational acceleration and  $f_i$ , for  $i \in \{A, B\}$ , is the Darcy-Weisbach friction factor in the stretch of the line occupied by the fluid  $i$ . For the sake of simplicity, the friction factor can be expressed as:

$$f_i = \frac{h}{(Re)_i^m} \quad \text{with} \quad (Re)_i = \frac{uD}{\nu_i}, \quad (9)$$

where  $h$  and  $m$  are constants which depend on the pipe's roughness and  $\nu_i$ , for  $i \in \{A, B\}$ , are the kinematic viscosities of the products.

The pump is supposed to be centrifugal and to be operating at constant speed so that its curve can be expressed as:

$$P_0 = (a - bQ^{2-m})\gamma_B \quad (10)$$

in which  $Q$  stands for the volumetric flow rate through the pump. If only one pump is employed, then  $Q = \pi D^2 u / 4$ .

By assuming that the pressure at the receiving point is held constant during the transfer, Eqs. (7) to (10) can be combined and solved for  $x_m$ , along with the initial condition  $x_m(t = 0) = 0$ , to finally obtain the bulk-average velocity as a function of time:

$$u(t) = NL \left[ \frac{3-m}{2-m} ANt + B^{\frac{3-m}{2-m}} \right]^{-\frac{1}{3-m}} \quad (11)$$

in which

$$\begin{aligned} A &= \alpha_B - \alpha_A \\ B &= \alpha_A + b\gamma_B \\ N &= \frac{4}{\pi D^2 L} (a\gamma_B - \Delta z\gamma_A - P_L)^{\frac{1}{2-m}} \\ \alpha_i &= \frac{h2^{3-2m}\pi^{m-2}\nu_i^m\gamma_i L}{gD^{5-m}} \quad \text{for } i \in \{A, B\}. \end{aligned}$$

Since  $N > 0$  and  $m < 2$ , an analysis of (11) reveals that the bulk-average velocity can increase as well as decrease along the transfer. It will increase (decrease) if  $A < 0$  ( $A > 0$ ), what implies the following inequality in terms of the fluids' properties  $\nu_A^m\gamma_A > \nu_B^m\gamma_B$  ( $\nu_A^m\gamma_A < \nu_B^m\gamma_B$ ).

When the following dimensionless parameters for space and time are considered,

$$y = \frac{x}{D} - \tau, \quad (12)$$

$$\tau = \frac{1}{D} \int_0^t u(t') dt', \quad (13)$$

Eq. (6) can be put in a dimensionless and more convenient form:

$$\frac{\partial C_i}{\partial \tau} = \frac{\partial}{\partial y} \left[ K \frac{\partial C_i}{\partial y} \right] \quad \text{for } i \in \{A, B\} \quad (14)$$

in which  $K = \mathcal{K}/Du$  stands for the dimensionless effective axial dispersion coefficient. As reported in many analytical studies,  $K$  presents a strong dependence not only on the Reynolds number but also on the mean velocity and eddy diffusivity profiles in the turbulent core and pipe wall region. Also, a somewhat less important dependence of  $K$  on the Schmidt number is shown to take place at lower turbulent Reynolds numbers.

Since Taylor (Taylor, 1954) first conceived this approach in 1954, several different correlations have been proposed for  $K$  ((Tichacek et al., 1957); (Yablonski et al., 1960); (Maron and Galianov, 1970); (Wasan and Dayan, 1970); (Krantz and Wasan, 1974)). The main difference among them is due to the mean velocity and eddy diffusivity profiles adopted for evaluating  $K$ . The correlation proposed by Krantz and Wasan (Krantz and Wasan, 1974) is the most precise since the mean velocity and eddy diffusivity distributions used to compute  $K$  satisfy the equations of motion and boundary conditions in the wall region, providing a smooth and continuous transition to the universal mean velocity profile valid in the turbulent core. Based on these considerations, the Krantz and Wasan's correlation for  $K$  will be adopted in this paper. For the sake of simplicity, we consider that  $K$  has only the following simpler functional dependence on the Reynolds number of the mixture  $Re$ ,

$$K = c(Re)^d, \quad \text{with } Re = \frac{uD}{\nu} \quad (15)$$

where  $\nu$  is the kinematic viscosity of the mixture and  $c$  and  $d$  are constants. For a fixed Schmidt number, Eq. (15) can be interpreted as being a linearization of the correlation proposed by Krantz and Wasan (Krantz and Wasan, 1974) in a  $\log K \times \log Re$  plot depicted in their paper.

The functional dependence of  $K$  on  $C_i$ , which is disregarded in several other models (e.g., (Taylor, 1954), (Levenspiel, 1958); (Sjenitzer, 1958); (Aumicky, 1970); (Ovádi and Török, 1977)), appears when the kinematic viscosity of the mixture is expressed in terms of the viscosities and concentrations of the products in the mixture, such as (Gambill, 1959):

$$\nu^{\frac{1}{3}} = C_A\nu_A^{\frac{1}{3}} + C_B\nu_B^{\frac{1}{3}}. \quad (16)$$

For practical purposes, the parameters  $c$  and  $d$  of (15) are determined for a Reynolds number calculated by using the value of  $u$  at  $t = 0$  obtained by (11) along with the mixture viscosity  $\nu$  evaluated on the basis of a 50%-50% blend of "A" and "B" through (16).

One of the effects of considering  $K$  as being dependent on  $C_i$  is the asymmetric shape of concentration distribution profile, as reported by Netchval et al. (Netchval et al., 1972) and pointed out by Austin and Palfrey (Austin and Palfrey, 1964) and Haisan (Haisan, 1997) in field tests. In a more practical sense, Botros (Botros, 1984) has also shown, by using a suitable approximation, that errors up to 20% in mixing volumes could be committed when  $K$  is assumed not to depend on  $C_i$ .

The success of the models based on (14) in predicting the intermixing volumes relies heavily on the accuracy of the correlation used for  $K$  and on its implicit dependence on  $C_i$  and  $u(t)$  through the Reynolds number. Variations in either concentration or the bulk-average fluid velocity can promote changes in the Reynolds number capable to alter substantially the effective axial dispersion coefficient.

The approach considering time dependent fluid velocity and concentration dependent dispersion coefficient was originally proposed by Netchval et al. in 1972 (Netchval et al., 1972). Nevertheless, Netchval's model has

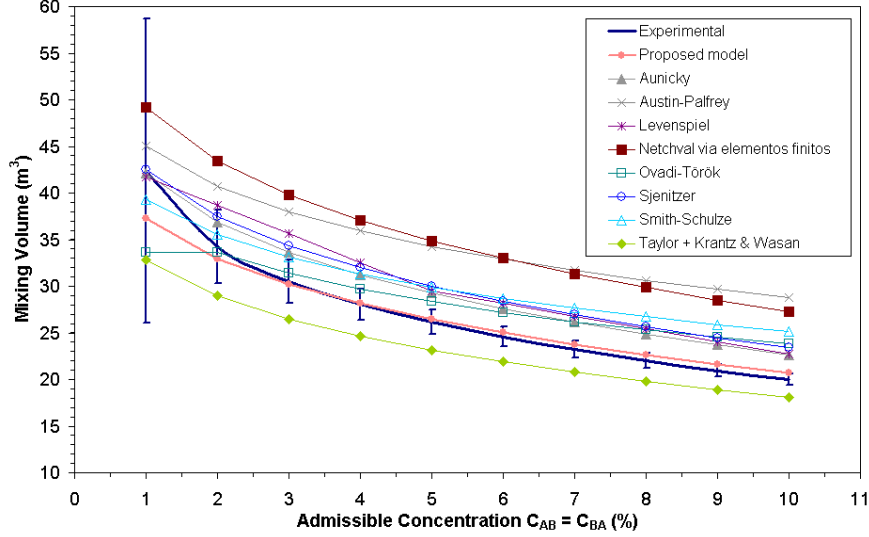


Figure 2: Comparison among mixing volume predictions

two weak points which severely compromises its application to practical cases. The first is due to the poor correlation used for  $K$  proposed by Yablonski et al. (Yablonski et al, 1960). The second one is related to the approximating analytical solution employed to solve the problem, which is restricted to situations for which the ratio of the products' kinematic viscosities can not exceed 8.

In view of the simple form adopted for  $K$  in (15), it is possible to impose the change of variables,

$$Z = \frac{y}{2\sqrt{\eta}} \quad (17)$$

in which

$$\eta = \int_0^\tau K_j(\tau') d\tau' \quad \text{with } K_j = cRe_j^d, \quad \text{for } j \in \{A, B\}, \quad (18)$$

and transform the problem described by Eq. (6) (or 14)), along with (2) or (3), into the following simpler boundary-value problem:

$$\frac{d}{dZ} \left[ (1 + \lambda_i C_i)^{-3d} \frac{dC_i}{dZ} \right] + 2Z \frac{dC_i}{dZ} = 0 \quad \text{for } i \in \{A, B\} \quad (19)$$

with

$$\begin{cases} C_B(Z \rightarrow -\infty) = 1 \\ C_B(Z \rightarrow +\infty) = 0 \end{cases} \quad \text{if } i = B \quad (20)$$

or

$$\begin{cases} C_A(Z \rightarrow -\infty) = 0 \\ C_A(Z \rightarrow +\infty) = 1 \end{cases} \quad \text{if } i = A \quad (21)$$

in which

$$\lambda_B = \left( \frac{\nu_B}{\nu_A} \right)^{\frac{1}{3}} - 1, \quad (22)$$

$$\lambda_A = \left( \frac{\nu_A}{\nu_B} \right)^{\frac{1}{3}} - 1. \quad (23)$$

Usage of Eq. (19) with  $i = A$  or  $i = B$  is carried out depending on the condition: if  $\nu_B \geq \nu_A$ , then  $i = B$  and  $j = A$ ; if  $\nu_A > \nu_B$ , then  $i = A$  and  $j = B$ .

Once the solution  $C_B(Z)$  of (19) with (20) is determined, Eq. (4) is used to evaluate  $C_A(Z)$ . Similarly, if  $C_A(Z)$  is computed through (19) with (21) then  $C_B(Z)$  is evaluated by (4). Whatever the case  $i = B$  or  $i = A$  is, the dimensionless time instants  $\tau_{BA} = \tau(t_{BA})$  and  $\tau_{AB} = \tau(t_{AB})$ , associated to the arrival of the beginning and of the end of the mixing volume at a generic position  $x = x^*$ ,  $x^* \in (0, L]$  can be written from (17) as:

$$\tau_{BA} = \frac{x^*}{D} - 2Z_{BA} \sqrt{\eta(\tau_{BA})} \quad (24)$$

$$\tau_{AB} = \frac{x^*}{D} - 2Z_{AB} \sqrt{\eta(\tau_{AB})} \quad (25)$$

in which the parameters  $Z_{AB} = C_B^{-1}(C_{AB})$  and  $Z_{BA} = C_A^{-1}(C_{BA})$  represent the values of  $Z$  for which the concentration assumes the pre-set admissible tolerances at the mixture zone extremities. Finally, based on (24), (25) and using the definition of  $\tau$  in (13), the contaminated mixing volume given by (5) at  $x = x^*$  can be expressed as

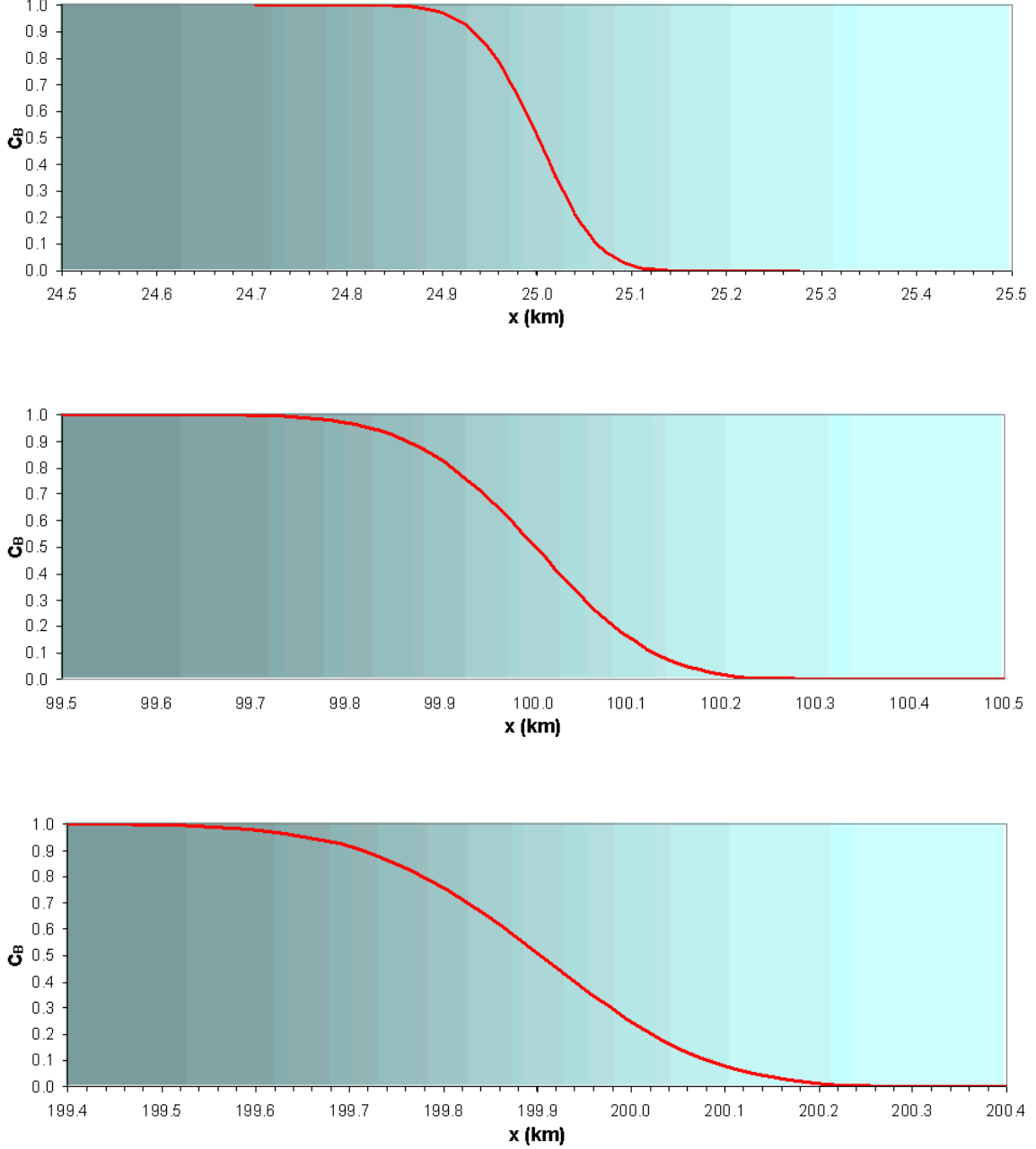


Figure 3: Mixing zone profile at three subsequent time instants during the transfer

$$V_c(x = x^*) = 2 \frac{D}{x^*} \left( Z_{BA} \sqrt{\eta(\tau_{BA})} - Z_{AB} \sqrt{\eta(\tau_{AB})} \right) V_p \tag{26}$$

in which  $V_p = \pi D^2 x^* / 4$  stands for the pipeline volume in the pipeline stretch  $[0, x^*]$ .

#### 4. Results

Before analyzing the evolution of the contaminated mixing zone during the batch transfer, a comparison between its predictions and experimental data are presented next for different values of admissible concentrations  $C_{AB} = C_{BA} = 1, 2, \dots, 9, 10\%$ . To find an approximating numerical solution of the non-linear problem described by (19) with (20) or (21), a finite element method coupled with a Newton's scheme is used (Freitas Rachid et alii, 1999b). The experimental data refers to a transfer of gasoline/diesel, being the gasoline the leading fluid

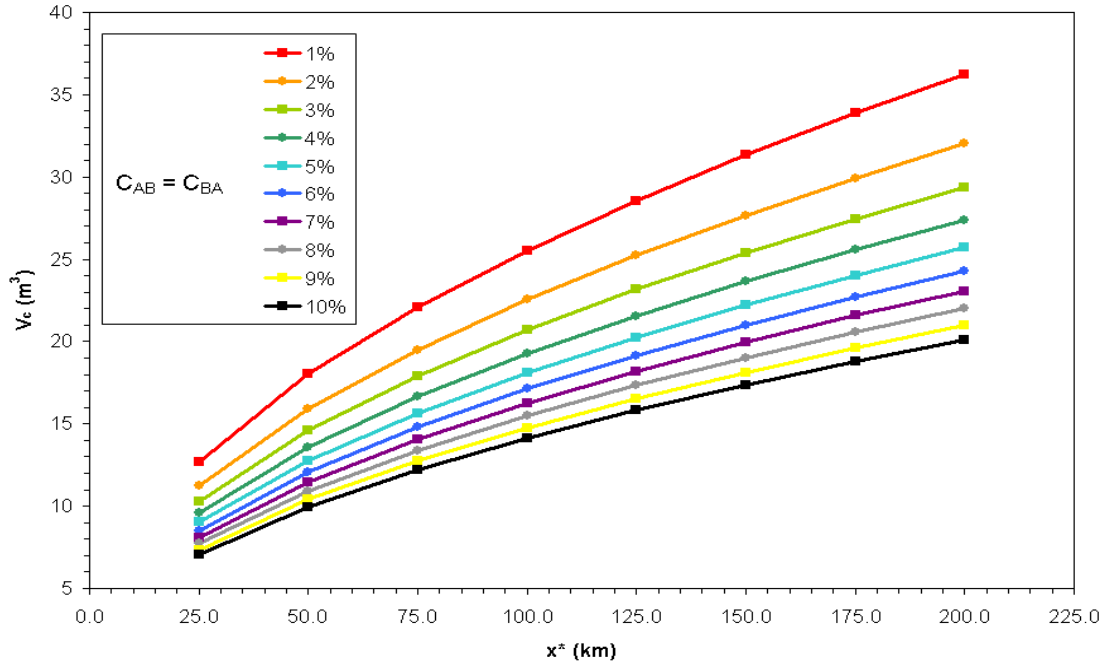


Figure 4: Mixing volume against admissible concentration at positions along the pipeline

( $\gamma_A = 734 \text{ kgf/m}^3$ ,  $\nu_A = 0.9 \text{ cSt}$ ) and the diesel the following product ( $\gamma_B = 833 \text{ kgf/m}^3$ ,  $\nu_B = 7.6 \text{ cSt}$ ). The transfer took place in a pipeline  $200 \text{ km}$  long having a nominal diameter of  $25.4 \text{ cm}$  (10 inches). The topographical difference level between the receiving point and the pump station is  $\Delta z = -895 \text{ m}$ .

The experimental mixing volume was evaluated by continuous monitoring the sonic velocity of the mixture at the receiving point,  $x = 199.9 \text{ km}$ . To do so, a clamp-on transit-time ultrasonic flow meter was used with an acquisition frequency of  $0.2 \text{ Hz}$  (Couto, 1998; Freitas Rachid et al, 1999a). The volumetric flow rate at the beginning of the transfer was  $245 \text{ m}^3/\text{h}$  and the pressure at the receiving point was  $P_L = 9.21 \text{ kgf/cm}^2$ . For these conditions, the initial Reynolds number of the mixture (50%-50% blend) is about  $1.3 \times 10^5$ , rendering the constants  $c = 0.803$  and  $d = -0.136$  in Eq. (15). The parameters  $a$  and  $b$  of Eq. (10) were determined from the pump curve and are equal to  $a = 378.8 \text{ m}$  and  $b = 5099.1 \text{ s}^{1.75}/\text{m}^{4.25}$ . For this pipeline, the following constants  $h = 0.3164$  and  $m = 0.25$  associated to the friction factor correlation (9) were used.

To better characterize the proposed model performance, Fig 2 shows the contaminated mixing volume as a function of admissible concentrations  $C_{AB} = C_{BA} = 1, 2, \dots, 9, 10\%$  for seven traditional methods currently used ((Sjenitzer, 1958), (Austin and Palfrey, 1964), (Levenspiel, 1958), (Ovádi and Török, 1977), (Smith and Schulze, 1948; Smith and Schulze, 1948), (Aunicky, 1970), (Netchval et al, 1972)) along with experimental data for  $x = 199.9 \text{ km}$ . Since the original method of Netchval et al. (Netchval et al, 1972) is not capable to deal with kinematic viscosity ratios  $\nu_A/\nu_B$  or  $\nu_B/\nu_A$  greater than 8, the predictions associated to this method displayed in Figs 2 was obtained by using the same finite element technique employed in (Freitas Rachid et al, 1999b). As expected, it can be seen in Fig 2 that the experimental mixing volume decreases as the admissible concentration increases. Such a trend is also observed for all the seven models, except to the Ovádi and Török (Ovádi and Török, 1977) which, as the authors say, is valid for  $C_{AB} = C_{BA} > 2\%$ . It can be clearly noted that the proposed method is the only one that presents the best agreement with experimental data over the whole range of concentration. The proposed model is the unique method whose predictions fits inside the uncertainty bars for any admissible concentration.

The greatest discrepancies between the proposed model and the experimental data take place in the range of concentrations  $1\% \leq C_{AB} = C_{BA} \leq 2\%$ . However, since the uncertainty associated to the mixing volume varies from  $\pm 16.32 \text{ m}^3$  (for  $C_{AB} = C_{BA} = 1\%$ ) to  $\pm 3.95 \text{ m}^3$  (for  $C_{AB} = C_{BA} = 2\%$ ), as seen in Fig 2, no comparison among the methods can be effectively done for  $C_{AB} = C_{BA} < 2\%$  inasmuch as their predictions are within the experimental uncertainty. The practical significance of the proposed method ability in accurately predicting mixing volumes is best highlighted by computing the largest and the smallest relative errors for the best and the worst estimates among the seven methods. The worst mixing volume estimates are given by the Netchval's method (Netchval et al, 1972), being the relative error in the interval  $[16.1\%, 36.5\%]$ . For  $C_{AB} = C_{BA} \geq 2\%$ , the Ovádi and Török (Ovádi and Török, 1977) method provides the best estimates with relative errors in the range  $[1.8\%, 20.6\%]$ .

To characterize the main features that render the proposed model the best performance, it is also depicted in

Fig 2 the mixing volume predictions obtained by using the Taylor’s formulation ( $K = constant$ ) (Taylor, 1954), being  $K = 0.1611$  evaluated by using the Krantz and Wasan (Krantz and Wasan, 1974) correlation for a Reynolds number of  $1.3 \times 10^5$  (corresponding to an initial flow rate of  $245m^3/h$  and a mixture kinematic viscosity based on a 50%-50% blend). As it can be seen in Fig 2 , the results obtained through this procedure (designated as Taylor + Krantz & Wasan) underestimates the mixing volume for the entire range of admissible concentration analyzed. By comparing the results of Taylor + Krantz & Wasan, Netchval et al., the proposed model and the experimental results, we can conclude that the best estimates of mixing volumes are achieved when both an accurate correlation for  $K$  and its functional dependence on the concentration as well as on the bulk-average velocity are properly accounted for in the model.

Once the model accuracy has been tested and validated, we use the model to investigate how the mixing zone evolution takes place as the transfer is carried out. Figure 3 presents the mixing zone at three distinct subsequent time instants, which corresponds approximately to the passage of the conventional half-length mixture to the spatial positions  $x^* = 25km$ ,  $x^* = 100km$  and  $x^* = 199.9km$ . Despite some beliefs, one can see that the mixing zone region increases continuously from the beginning to the end of the transfer. This statement can be best visualized whether one plots the mixing volume ( computed based on different admissible concentrations) at several distinct spatial positions along the pipe, as illustrated in Fig 4. Although the mixing volume increases continuously, we can see that it occurs with a reducing pace. Moreover, the lower the admissible concentration used to evaluate the mixing volume is, the larger the time rate increase becomes.

Another interesting way to illustrate the aforementioned results is plotting the mixing volume against the admissible concentrations for different spatial positions along the pipeline, as shown in Fig 5. It becomes clear in this figure that the mixing volumes evaluated with admissible concentrations raging from 6% to 10% do not differ markedly from each other, when they are computed at the earlier stages of the transfer.

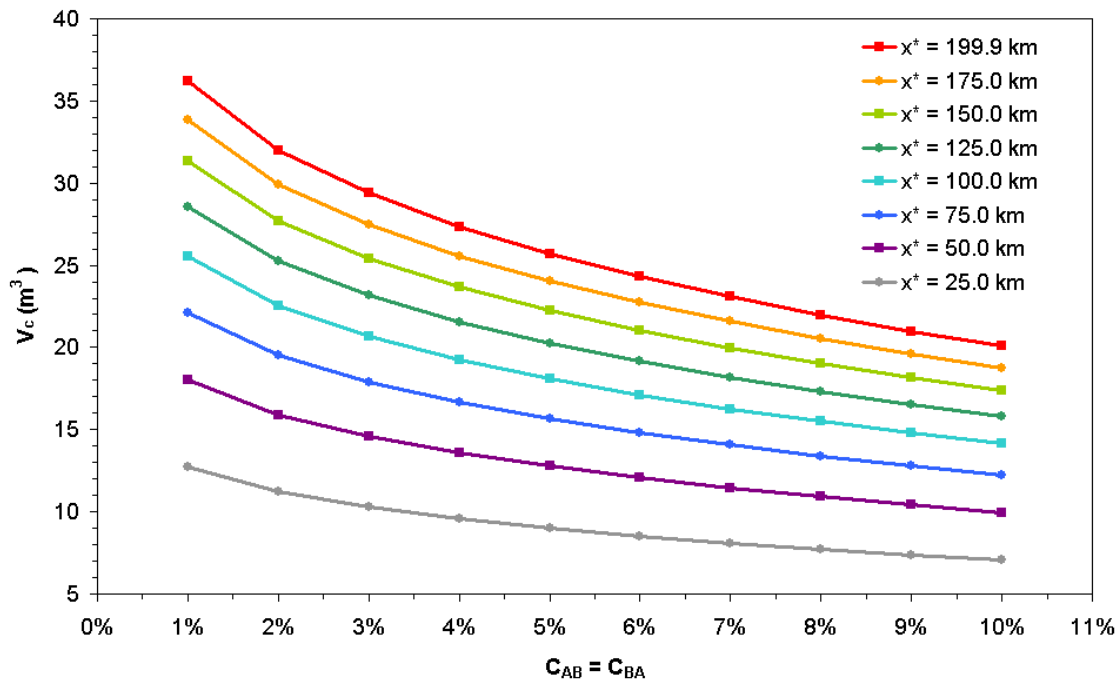


Figure 5: Mixing volume against spatial position for different admissible concentrations

### 5. Conclusions

A model has been proposed in this paper to evaluate mixing volumes in serial transport of petroleum products in pipelines. Its novel features are the incorporation of the flow rate variation with time and the use of a more precise effective dispersion coefficient, which is considered to depend on the concentration. A comparison with experimental data and with other seven traditional methods available in the literature shows that the proposed model is the most accurate over the entire range of admissible concentrations used to compute the mixing volume. It is shown that the success of the proposed model relies not only on the use of accurate correlations for the dispersion coefficient (in terms of Reynolds and Schmidt numbers) but also on its dependence on the concentration and time-varying bulk-average velocity. Finally, it is also demonstrated that the mixing volume continuously increases along the transfer with a decreasing pace.



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## References

- Hara, A., Hayashi, H., Tsuchiya, M., 1979, "Sphere separation system aids long-haul oil-product transport," *The Oil and Gas Journal*, Vol. 22, pp. 47–51.
- Benke, H. and Gravert, W., 1989, "Process Analysers for optimisation of reinjection of contaminated product in product pipelines," *Oil Gas European Magazine*, Vol. 1, pp. 42–44.
- Aunicky, Z., 1970, "The longitudinal mixing of liquids flowing successively in pipelines," *The Canadian Journal of Chemical Engineering*, Vol. 48, pp. 12–16.
- Austin, J. E., and Palfrey, J. R., 1964, "Mixing of miscible but dissimilar liquids in a serial flow in a pipeline," *Proc. Institution of Mechanical Engineers*, Vol. 178, Part1, No.15, pp. 377–395.
- Levenspiel, O., 1958, "How much mixing occurs between batches?," *Pipe Line Industry*, pp. 51–54.
- Ovádi, Z., and Török, E., 1977, "Examination of pipeline transport from the viewpoint of goods quality," *Kőolaj és Földgáz*, Vol. 10, No. 4, pp. 121–123. In Hungarian.
- Sjenitzer, F., 1958, "How much do products mix in a pipeline?," *The Pipeline Engineer*, pp. D31–D34.
- Smith, S. S., and Schulze, R. K., 1948, "Interfacial mixing characteristics of products in products pipe line - Part 1," *The Petroleum Engineer*, Vol. 20, pp. 94–104.
- Maron, V. I., and Galianov, A. I., 1970, "Mixture coefficients for liquid flows in pipes," *Izvestia Vischikh Utchebuikh Zavedenii Neft'i Gaz*, Vol. 2, pp. 74–78. In Russian.
- Botros, K. K., 1984, "Estimating contamination between batches in products lines," *The Oil and Gas Journal*, pp. 112–114.
- Netchval, M. V., Tugunov, P. I., and Slesáreva, V. G., 1972, "Mixture formation during the process of sequential pumping of petroleum products of varying viscosities," *Neftyanoy Khozyaistvo*, Vol. 50, No. 8, pp. 57–59. In Russian.
- Smith, S. S., and Schulze, R. K., 1948, "Interfacial mixing characteristics of products in products pipe line - Part 2," *The Petroleum Engineer*, Vol. 20, pp. 7–12.
- Taylor, G. I., 1954, "The dispersion of matter in turbulent flow through a pipe," *Proceedings of Royal Society, Series A*, Vol. 223, pp. 446–468.
- Tichacek, L. J., Barkelew, C. H., and Baron, T., 1957, "Axial Mixing in Pipes," *American Institute of Chemical Engineering Journal*, Vol.3, No. 4, pp.439–442.
- Atesmen, K. M., Baldwin, L. V., and Haberstroh, R. D., 1971, "The dispersion of matter in turbulent pipe flows," *ASME Journal of Basic Engineers*, pp. 161–177.
- Yablonski, V. S., Asaturyan, A. S., and Khizgilov, I. H., 1960, "Turbulent Diffusion in Pipes," *Inzhenierno-Fizicheski Zhurnal*, Vol. 3, No. 3, pp. 117–122. In Russian.
- Wasan, D. T., and Dayan, J., 1970, "The dispersion of solute in turbulent pipe flow of non-Newtonian fluids," *The Canadian Journal of Chemical Engineering*, Vol. 48, pp. 129–131.
- Krantz, W. B., and Wasan, D. T., 1974, "Axial dispersion in turbulent flow of power-law fluids in straight tubes," *Industrial Chemical Fundamentals*, Vol. 13, No. 1, pp. 56–61.
- Gambill, W. R., 1959, "How to estimate mixtures viscosities," *Chemical Engineering*, Vol. 66, pp. 151–152.
- Haisan, Q., 1997, "Contamination due to intermixing in pipelines," *Journal of Petrochemical Universities*, Vol. 10, No. 2, pp. 68–70. In Chinese.

- Couto, N. C., 1998, "Products separation in pipeline by using low-density foam pigs (Final Report)," Internal Report Petrobras - Petróleo Brasileiro S.A., CENPES/SUPEP/DIPILOT/SEPROT, Project 02.05.07. In Portuguese.
- Freitas Rachid, F. B., Carneiro de Araujo, J. H., Martins-Costa, M. L., and Costa Mattos, H. S., 1999a, "Development of models and operational procedures aiming to minimize the interfacial volume - Second Report," Internal Report, Project Petrobras-UFF. In Portuguese.
- Freitas Rachid, F. B., Carneiro de Araujo, J. H., Martins-Costa, M. L., and Costa Mattos, H. S., 1999b, "Development of models and operational procedures aiming to minimize the interfacial volume - Third Report," Internal Report, Project Petrobras-UFF. In Portuguese.
- Bischoff, K. B., and Levenspiel, O., 1962, "Fluid dispersion-generalization and comparison of mathematical models-II Comparison of models," *Chemical Engineering Science*, Vol. 17, pp. 257–264.