MODELING THE DISPERSION PHENOMENON IN BATCH TRANSFER OPERATIONS BY THE THEORY OF STRUCTURED MIXTURE

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ABSTRACT

This paper investigates the design of a model based on the theory of structured mixture that retains the character of one-dimensional models from the literature, simultaneously being able to estimate the mixing volume with consistency. The determination of the mixing volume in batch transfer operations based on this novel approach can facilitate the understanding of which parameters affect their growth, and even the optimization of multiproduct pipeline operations, in order to reduce the mixing volume. Among other things, the proposed study can generate considerable indirect economic impacts, by providing transported goods with higher quality control, and enabling more efficient planning of the pipeline operations with regard to implementation of pumping stops, thereby providing operational flexibility and reliability.

KEYWORDS

continuum mechanics; mixing volume; batch transfer; theory of mixtures; multiproduct pipeline operation

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1. INTRODUCTION

The transportation of oil products in long pipelines is a common practice in the petroleum industry and often involves sequential displacement of at least two dissimilar products, or different purity grades of one product in the same line. This type of operation is called batch transfer, and can be in practice performed in two different ways. The first is to use physical separators, usually pigs or scraper balls, in order to prevent the mixing process and subsequent product contamination at interfaces. In this type of operation, physical separators should be properly aligned and connected with each pumping station, thereby requiring skilled labor and time to perform these tasks. In particular, the operation with pigs or scrapers in pipelines where there are numerous intermediate pumping stations may become technically unfeasible. In the second way, commonly used when operational and economic issues arise, no physical separators are used, which ultimately enables the formation of a mixing zone at the products interfaces, as illustrated in Figure 1. After covering the pipeline, the mixing zone should be adequately quantified in terms of degradation of products transported or stored in a specific mixing tank, for subsequent re-processing at the refinery in situations where there is incompatibility of products (Botros, 1984; Coimbra, 1981; Drew and Passman, 1998; Tepedino et al., 2007).

Figure 1. Schematic representation of a batch transfer of two products – generically labeled as products A and B – and a pipeline in the absence of physical separators.

In this case, it is initially assumed that the product identified as A is already being pumped when, at a given moment, product B is inserted in line. After the adequate alignment of tanks, the pump responsible for the displacement of product A is turned off by setting on the pump connected to the tank that stores product B. From this moment on, product B starts to push product A through the line by initiating the formation of a

mixing zone at the products interface. Due to this type of arrangement, product B is called a fluid shifter while product A is called the displaced fluid.

From an economic standpoint, the formation of mixing zones in batch transfers implies additional costs associated with transportation of the mixture back to the refinery and its re-processing. Aiming to reduce the costs associated with the batch transfer, the volume of the mixing zone should be as small as possible, especially when this mixture can not be diluted with other products, for an example when the displaced products are diesel and gasoline (Tepedino et al., 2007). From an operational viewpoint, the prediction of the mixing zone extent through mechanical models is an important factor, not only at the implementation stage but also when planning the transfer, allowing, *a priori*, to monitor the volume required to pack the mixture and identify key issues responsible for the growth of the mixing zone during the transfer.

In order to better characterize the peculiarities associated with the batch transfer in the absence of physical separators, it should be noted that this operation involves the displacement of a large variety of products over distances that can reach hundreds of kilometers. Thus, the existence of more than one interface along the line, the injection and withdrawal of products in intermediate pumping stations, different diameters and different topographic dimensions, as well as the occurrence of shutdowns and accidental stops of the pumping operations are key factors that could significantly affect the mixing volume. It should be emphasized that there are no one-dimensional models available in the literature that allow to adequately quantify these parameters and their effects on batch transfer operations. Therefore, the methodology used in this article, based on a software developed in Fortran®, is founded on the structured mixing theory, retaining the character of one-dimensional models and simultaneously being able to estimate the mixing volume or the length of the mixing zone with consistency. In this paper, we aim to show the influence of the speed with which the batch transfer is conducted and also the influence of the pipeline extension on the mixing volume.

2. MATERIALS AND METHODS

In order to better characterize the phenomenon in question, one must consider the batch transfer in which product A is the displaced fluid and product B is the shifting fluid, as previously characterized in Figure 1. Let t and s be independent variables that respectively represent the time and axial position along the centerline of the pipeline, upstream the pumping station and downstream the receiving station, where L is the total length of the duct. This implies that $C_i(s, t) \in [0, 1]$, with $i \in \{A, B\}$ being the average time of concentration of fluid i inside the mixture at the cross section of the duct in s and t . Mathematically, $C_i(s, t)$ is defined by:

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

where r and θ refer to radial and circumferential coordinates, A is the cross-sectional area of the duct and $2\Delta t$ is a time period, long enough to eliminate the fluctuations of the turbulence and small enough to capture changes in the concentration camp $\hat{C}_i(s, r, \theta, t)$, with $C_A + C_B = 1$.

Under certain circumstances, at the beginning of the batch transfer, concentration can be represented as a leap in $s = 0$. For example, suppose that we work with the concentration of fluid B (C_B), and then jumps occur as seen in Figure 2, which illustrates the configuration of the mixing zone in terms of the concentration of product B at different positions on a continuous transfer duct. During transfer, a mixing zone is formed on the edge of two adjacent products. The extent of this region increases as the transfer takes place along the line. This phenomenon can be observed in Figure 2 at $t = t_1$ and $t = t_2$, when the conventional half of the mixture (the material point at which the concentrations are equal to 0.5) reaches $s = s_1$ and $s = s_2$ at the space positions.

For practical implementation, the mixing volume at the tube position $s = s_i$ can only be evaluated when the permissible concentrations of product B in A, at the beginning of the mixing zone, C_{BA} , and that of product A in B at the end of the mixing zone, C_{AB} , are specified as shown in Figure 3. Once these values (which are not necessarily equal)

Figure 2. Evolution of spatial mixing zone in a continuous transfer operation.

have been chosen so that the technical specifications of the products are not changed, the volume of mixture at $s = s_i$ can be calculated with the following expression:

$$V_i(s = s_i) = \int_{t_{BA}}^{t_{AB}} Q(s = s_i, t) dt \quad (2)$$

in which $Q(s = s_i, t)$ represents the volumetric flow at spatial location $s = s_i$, t_{BA} represents the initial time of the mixing zone with the concentration C_{BA} at position $s = s_i$, and t_{AB} is the final time of the mixing zone with concentration $(1 - C_{AB})$ at the position $s = s_i$.

Figure 3. Characterization of the mixing volume at position $s = s_i$ at $t = t_{BA}$ and $t = t_{AB}$.

The formation of the products mixing zone in continuous transfers (*i.e.*, transfers in which the pump does not stop) is predominantly governed by the phenomenon of mass dispersion. This phenomenon results from the combination of the diffusion process in the radial direction with the existence of a non-uniform velocity profile of the mixture in developed laminar or turbulent flow (Aunicky, 1970; Baptista et al., 2000a and 2000b; Baptista et al., 2001; Botros, 1984; Carvalho et al., 2002; Freitas Rachid et al., 2001a and 2001b; Freitas Rachid et al., 2002a and 2002b; Maron and Galianov, 1970; Netchval et al., 1972; Ovadi and Török, 1977; Sjenitzer, 1958; Smith and Schulze, 1948).

2.1 Balance equations

Generally, a mixture can be considered as an aggregate composed of various materials, defined as the constituents which may be present in different phases, such as fluid-fluid, solid-fluid, solid-solid, among others. To facilitate the understanding of key issues, consider a real mixture at a given time occupying a region R of the space, as shown in Figure 4.

Figure 4. Constituents composition in a real and idealized mixture.

In a sample of the mixture, extracted from the vicinity of position $x \in R \subset R^k$, " $k \in \{1, 2, 3\}$ " notes that the mixture can be composed of different constituents α , β and γ , that occupy separate regions of an infinitesimal volume dV . Differently from what occurs in real situations, this is accepted as a basic premise of the Theory of Mixtures, and the three constituents α , β and γ occupy fully and simultaneously the same infinitesimal volume dV in an idealized region $R^m \subset R^k$, similar to R , as seen in Figure 4. The interfaces among them observed in the volume element in R are no longer noticed in the idealized volume element R^m , since a homogenization process was conducted in the passage from R to R^m . Thus, the mixture in R^m can be interpreted as a current configuration in the usual sense of the classical theory of Continuum Mechanics, in which each spatial position $x \in R^m \subset R^k$,

Figure 5. Representation of multiple reference configurations.

Figure 6. Representation of a control volume within the duct.

" $k \in \{1, 2, 3\}$ " is occupied at any time for each particle of each constituent, with independent thermo-mechanical properties. Alternatively, each component can be represented in different references configurations R_i (in this case, with $i \in \{1, 2, 3\}$ for a mixture composed of three constituents α , β and γ), as shown in Figure 5.

Analogously to the idealized representation of a mixture illustrated in Figure 4, the region R^m in this study represents a control volume bounded by the inner wall of a straight tube of inner radius $\hat{R}(s, t)$, with $R(s) = \hat{R}(s, t = 0)$ designating the radius of the tube in a non-deformed configuration, and cross sections $s = a$ and $s = b$, with $b > a$. The internal cross section area of the duct is denoted by $A(s, t)$, with $A = \pi \hat{R}^2$. In this characterization, $s \in (0, L)$ (L is the length of the tube) represents the spatial coordinate measured along the centerline of the tube and t is time (see Figure 6).

The control volume illustrated in Figure 6 is occupied by a mixture consisting of $2n$ constituents (n constituents for each product, A and B) moving through the duct of variable cross section area along the spatial coordinate s (due to its original form) and time t (as a result of hydrodynamic loads on the inside). Herein are presented the mechanics basic laws of Theory of Mixtures, namely: mass balance, balance of the amount of linear motion and thermodynamics laws (1st and 2nd).

2.1.1 Mass conservation

The parameters $\rho_i^{(j)}(s, t)$ and $v_i^{(j)}(s, t)$, with $i \in \{A, B\}$ and $j \in \{1, \dots, n\}$, are the spatial fields of density of the constituent in the mixture (i.e., the ratio between the mass of the i -th constituent per unit volume of the mixture) and the speed associated with the j -th constituent of the i -th product. It then follows that the principle of mass conservation in the mixture as a whole in the integral formulation can be expressed as:

$$\sum_{j=1}^n \sum_{i=A}^B \left\{ \frac{d}{dt} \int_a^b \rho_i^{(j)} A ds + [\rho_i^{(j)} v_i^{(j)} A]_{s=b} - [\rho_i^{(j)} v_i^{(j)} A]_{s=a} \right\} = 0 \quad (3)$$

with $\forall [a, b] \subset (0, L)$ and $t \in (0, \infty)$.

To take into account the possibility that there may be an exchange of mass between the constituents of the same product or between different products, the integral formulation of the principle of mass conservation for each constituent is postulated as:

$$\frac{d}{dt} \int_a^b \rho_i^{(j)} A ds + [\rho_i^{(j)} v_i^{(j)} A]_{s=b} - [\rho_i^{(j)} v_i^{(j)} A]_{s=a} - \int_a^b \Phi_i^{(j)} A ds = 0 \quad (4)$$

where $\forall i \in \{A, B\}$, $j \in \{1, \dots, n\}$ and $\forall [a, b] \subset (0, L)$, with $\Phi_i^{(j)}$ representing the temporal rate of mass transfer of all other constituents of both products A and B to the j -th constituent of the i -th product. Assuming that there is no mass transfer from one product to another due to chemical reactions and no phase change, then the term $\Phi_i^{(j)}$ can be properly expressed as:

$$\Phi_i^{(j)} = \sum_{\substack{k=1 \\ k \neq j}}^n \hat{\Phi}_i^{(j, k)} \quad (5)$$

and represents the temporal rate of mass transfer of the k -th component for the j -th constituent, both referring to the i -th product. The term $\Phi_i^{(j)}$ appears in the model in question to take into account the mass flow in the radial direction, since the proposed model herein is essentially one-dimensional.

Assuming that the spatial fields appearing in Equation (4) have sufficient regularity, it is possible to show that the differential form of the principle of mass conservation for each constituent is expressed as:

$$\frac{\partial \rho}{\partial t} (\rho_i^{(j)} A) + \frac{\partial}{\partial s} (\rho_i^{(j)} v_i^{(j)} A) - A \Phi_i^{(j)} = 0 \quad (6)$$

with $\forall i \in \{A, B\}$, $j \in \{1, \dots, n\}$ and $\forall (s, t) \in (0, L) \times (0, \infty)$.

Taking the sum in i and j of Equation (4) and combining with Equation (3), and assuming sufficient regularity of the fields, one can write the differential form of the overall mixture as:

$$\sum_{j=1}^n \sum_{i=A}^B \Phi_i^{(j)} = 0 \quad (7)$$

2.1.2 Conservation of linear movement quantity

The local form of the principle to conservation of linear movement quantity for the j -th constituent of the i -th product in an inertial frame is reduced to the following expression:

$$\rho_i^{(j)} \frac{d^{(i, j)} v_i^{(j)}}{dt} = \frac{\partial t_i^{(j)}}{\partial s} + b \rho_i^{(j)} - a_i^{(j)} - m_i^{(j)} - \sum_{\substack{k=1 \\ k \neq j}}^n \hat{\Phi}_i^{(j, k)} (v_i^{(j)} - \hat{\Gamma}_i^{(j, k)}) \quad (8)$$

with $\forall i \in \{A, B\}$, $j \in \{1, \dots, n\}$ and $\forall (s, t) \in (0, L) \times (0, \infty)$

where $\rho_i^{(j)}$ is the density of the j -th constituent of the i -th product, b is the projection of the body force per unit mass in the axial direction of the tube, $t_i^{(j)}$ is the surface force per unit area cross section exerted on the j -th constituent of the i -th product in the vicinity of $s = a$ and $s = b$, $a_i^{(j)}$ is the friction force per unit volume exerted by the

reactive nature on the side wall of the duct, $m_i^{(j)}$ represents the internal interaction force per unit volume exerted by all others constituents of products A and B on the j -th constituent of the i -th product, $\Gamma_i^{(j)}\Phi_i^{(j)}$ is the term that defines the amount of linear motion per unit volume supplied to the j -th constituent of the i -th product due to mass flow $\Phi_i^{(j)}$, and the $\frac{d^{(i,j)}v_i^{(j)}}{dt}$ term represents the material derivative in Eulerian coordinates of the velocity of the j -th constituent of the i -th product.

2.1.3 First law of thermodynamics

The local form for each component product can be expressed as:

$$\rho_i^{(j)} \frac{d^{(i,j)}u_i^{(j)}}{dt} = t_i^{(j)} \frac{\partial v_i^{(j)}}{\partial s} + m_i^{(j)} v_i^{(j)} + a_i^{(j)} v_i^{(j)} + \rho_i^{(j)} r_i^{(j)} - {}_L q_i^{(j)} - \frac{1}{A} \frac{\partial}{\partial s} ({}_s q_i^{(j)} A) + \frac{{}_R r_i^{(j)}}{A} \frac{d^{(i,j)}A}{dt} + \sum_{\substack{k=1 \\ k \neq j}}^n \hat{\Phi}_i^{(j,k)} \left(\hat{G}_i^{(j,k)} + \frac{1}{2} (v_i^{(j)})^2 - u_i^{(j)} - \hat{\Gamma}_i^{(j,k)} v_i^{(j)} \right) + \varepsilon_i^{(j)} \quad (9)$$

with $\forall i \in \{A, B\}$, $j \in \{1, \dots, n\}$ and $\forall (s, t) \in (0, L) \times (0, \infty)$, where $r_i^{(j)}$ is the temporal rate of generation/absorption of heat per unit mass associated with the j -th constituent of the i -th product, ${}_s q_i^{(j)}$ is the heat flux per unit tube cross section area, ${}_L q_i^{(j)}$ is the radial heat flow through the pipe wall, $\hat{G}_i^{(j,k)}$ represents the sum of kinetic and specific internal energy of the k -th constituent of the i -th product immediately before becoming part of the j -th constituent of the i -th product, and $u_i^{(j)}$ represents the specific internal energy of the j -th constituent of the i -th product.

2.1.4 Second law of thermodynamics

Defining $s_i^{(j)}$ as the specific entropy of the j -th constituent of the i -th product, and assuming the basic premise that all constituents of both products share the same temperature, the second law can be postulated for the overall mixture in the integral form as:

$$\sum_{j=1}^n \sum_{i=A}^B \left\{ \frac{d}{dt} \int_a^b \rho_i^{(j)} s_i^{(j)} A ds + [\rho_i^{(j)} s_i^{(j)} A v_i^{(j)}]_{s=b} - [\rho_i^{(j)} s_i^{(j)} A v_i^{(j)}]_{s=a} \right\} \geq$$

$$\geq \sum_{j=1}^n \sum_{i=A}^B \left\{ - \left[\int_a^b {}_L q_i^{(j)} \frac{ds}{\theta} + \int_a^b \frac{\partial}{\partial s} \left({}_s q_i^{(j)} A \right) ds \right] + \int_a^b \frac{\rho_i^{(j)} r_i^{(j)} A}{\theta} ds + \int_a^b \frac{m_{i,R}^{(j)} v_i^{(j)}}{\theta} A ds + \int_a^b \frac{a_i^{(j)} v_i^{(j)}}{\theta} A ds \right\} \quad (10)$$

with $\forall [a, b] \subset (0, L)$ and $t \in [0, \infty]$, where $m_{i,R}^{(j)}$ is the portion of the strength of interaction between adjacent constituents, given from:

$$m_i^{(j)} = m_{i,R}^{(j)} + m_{i,s}^{(j)} \quad (11)$$

where $m_{i,s}^{(j)}$ is the portion of the strength of interaction between axial constituents.

It should be noted that the postulation of the second law for the overall mixture, expressed by Equation (10), implies the classical form adopted for the second law for the mixture as a whole (Atkin and Craine, 1976; Rajagopal and Tao, 1995).

2.2 Mixing volume

Assuming that tolerable levels of purity of product A in the mixture (C_{AB}) and product B in the mixture (C_{BA}) are equal, $C_{AB} = C_{BA} = C$, then the mixing length L_m can be calculated and expressed by Equation (12).

$$L_m = 4\sqrt{KLD} \cdot \text{erf}^{-1}(2C - 1) \quad (12)$$

where K is the virtual diffusion coefficient of an axial mass in dimensionless form, L is the pipeline length, D is the pipe diameter, erf^{-1} is the reversed error function and C is the mixture concentration ($C_{AB} = C_{BA} = C$).

The mixing volume (V_m) is then calculated by multiplying the mixing length by the area of the pipe cross section, as in Equation (13).

$$V_m = L_m \cdot A \quad (13)$$

In this paper, we aim to show the influence of the speed with which the batch transfer is conducted and also the influence of the pipeline extension on the mixing volume. The analytical solution presented above could then be used in a situation where the fluid shifter (product B) and displaced fluid (product A) are diesel and gasoline, whose kinematic viscosities are $\nu_B = 8.7$ cSt and $\nu_A = 0.9$ cSt, respectively. It is assumed that the

diameter of the duct is 0.58 m and the viscosity of the mixture (ν) is constant and equal to 3.45 cSt, assuming $C_A = C_B = 0.5$, based on the Gambil correlation (1959), which is given by the following expression:

$$\nu^{1/3} = C_A \nu_A^{1/3} + C_B \nu_B^{1/3} \quad (14)$$

Furthermore, the mixing volumes are calculated for different average velocity values of the fluid (0.01; 0.10; 1.00; 2.00 and 3.00 m/s) and for different pipeline extensions (100, 200, 400 and 800 km). The dispersion coefficient for the calculated Reynolds number (Re) and the estimated Schmidt number ($Sc = 100$) were obtained through the correlation of **Krantz and Wasan (1974)**. Table 1 shows the values of Re and K for the different speeds mentioned above.

Table 1. Values of Re and K for the different speeds.

$\langle \bar{v}_s \rangle$	Re	K
3.00	4.42E+05	0.16
2.00	2.94E+05	0.17
1.00	1.47E+05	0.18
0.10	1.47E+04	0.70
0.01	1.47E+03	10.0

3. RESULTS AND DISCUSSION

The mixing volumes had been calculated using our own software developed in Fortran® for different speeds, considering $C_{AB} = C_{BA}$ as the permissible concentrations ranging from 1 to 10 % after half of the conventional mixture had covered a distance of 100, 200, 400 and 800 km respectively. The results are depicted in Figure 7, and are also summarized in Table 2, for concentrations equal to 1, 5 and 10 %, for average speeds of 3.00; 2.00; 1.00; 0.10 and 0.01 m/s, and pipe lengths of 100, 200, 400 and 800 km. The highlighted cells in Table 2 indicate the different speeds for each permissible concentration.

Figure 7. Changes in mixing volume as a function of flow velocity for each value of permissible concentration.

In Figure 7, one can see that the mixing volume changes as a function of flow velocity for each value of permissible concentration ranging between 1 and 10 %.

The graphs shown in Figure 8 illustrate the mixing volume computed as a function of the Reynolds number after half of the conventional mixture had covered 100 km and 800 km of the pipeline, respectively. The graphs for 200 km and 400 km are not shown because they have the same curve profile of the previous ones.

Figure 8. Mixture volume as a function of Reynolds number for (a) 100 km and (b) 800 km with $C_{AB} = C_{BA} = 1\%$.

Table 2. Calculated mixture volume at different C_{AB} concentrations.

C_{AB} (%)	v (m/s)	L (km)	L_m (m)	V_m (m ³)	C_{AB} (%)	v (m/s)	L (km)	L_m (m)	V_m (m ³)
0.01	3.00	100	593.21	120.23	0.05	1.00	400	889.75	180.33
0.01	3.00	200	838.93	170.03	0.05	1.00	800	1258.30	255.03
0.01	3.00	400	1186.43	240.46	0.05	0.10	100	877.31	177.81
0.01	3.00	800	1677.86	340.06	0.05	0.10	200	1240.70	251.46
0.01	2.00	100	611.47	123.93	0.05	0.10	400	1754.62	355.62
0.01	2.00	200	864.75	175.27	0.05	0.10	800	2481.41	502.92
0.01	2.00	400	1222.94	247.86	0.05	0.01	100	3315.92	672.06
0.01	2.00	800	1729.50	350.53	0.05	0.01	200	4689.42	950.44
0.01	1.00	100	629.2	127.52	0.05	0.01	400	6631.84	1344.12
0.01	1.00	200	889.82	180.35	0.05	0.01	800	9378.84	1900.88
0.01	1.00	400	1258.40	255.05	0.10	3.00	100	326.79	66.23
0.01	1.00	800	1779.64	360.69	0.10	3.00	200	462.16	93.67
0.01	0.10	100	1240.80	251.48	0.10	3.00	400	653.59	132.47
0.01	0.10	200	1754.75	355.65	0.10	3.00	800	924.31	187.34
0.01	0.10	400	2481.59	502.96	0.10	2.00	100	336.85	68.27
0.01	0.10	800	3509.50	711.3	0.10	2.00	200	476.38	96.55
0.01	0.01	100	4689.77	950.51	0.10	2.00	400	673.7	136.54
0.01	0.01	200	6632.34	1344.22	0.10	2.00	800	952.76	193.1
0.01	0.01	400	9379.54	1901.02	0.10	1.00	100	346.62	70.25
0.01	0.01	800	13264.67	2688.44	0.10	1.00	200	490.19	99.35
0.05	3.00	100	419.43	85.01	0.10	1.00	400	693.23	140.5
0.05	3.00	200	593.17	120.22	0.10	1.00	800	980.38	198.7
0.05	3.00	400	838.87	170.02	0.10	0.10	100	683.54	138.54
0.05	3.00	800	1186.34	240.44	0.10	0.10	200	966.67	195.92
0.05	2.00	100	432.34	87.63	0.10	0.10	400	1367.07	277.07
0.05	2.00	200	611.43	123.92	0.10	0.10	800	1933.33	391.84
0.05	2.00	400	864.69	175.25	0.10	0.01	100	2583.53	523.62
0.05	2.00	800	1222.85	247.84	0.10	0.01	200	3653.66	740.51
0.05	1.00	100	444.88	90.17	0.10	0.01	400	5167.05	1047.24
0.05	1.00	200	629.15	127.51	0.10	0.01	800	7307.32	1481.03

According to the results presented in the data analysis, we could conclude that:

1. If the mass transfer is conducted with an average speed of 0.01 m/s, very high mixing volumes are obtained, as seen in Figure 8. Although speeds of that magnitude are not observed in transport within pipelines, they were included in these simulations only to characterize the fact that laminar flow regimes or full transitions should always be avoided, because they induce very large dispersion coefficients and, consequently, significantly higher mixing volumes;
2. It was also observed in Figure 8 that the volume of mixture increases with decreasing admissible concentration. Regardless of the permissible concentration, the higher the transfer speed, the larger the Reynolds number, and therefore the lower the mixing volume. This last statement normally surprises the common

sense of those who mistakenly associate directly the turbulence to the mixing phenomenon.

There are similarities in the profile of the curves presented in Figure 8, and the mixing volume is sequentially higher due to greater line extension; therefore, the larger the line, the higher the contact time of fluids. In other words, regardless of the permissible concentration adopted, the greater the distance, the higher the mixing volume. Strictly speaking, Equation (12) shows that the mixing length, L_m , is proportional to the square root of the traveled distance.

4. CONCLUSIONS

As explained above, the mixing volume depends on the permitted concentration, the flow velocity, the pipeline extension and the Reynolds number.

When the dependence of the mixing volume with the Reynolds number is assessed, one confirms that the mixing volume increases with increasing pipeline length. It was demonstrated that the mixing volume is reduced with increasing Reynolds number. Alternatively, as the Reynolds number increases, it was seen that, from a certain value, the mixing volume becomes almost independent of Reynolds. This fact shows that an increase in the flow velocity during the conducted batch transfer will not always result in substantial reduction of the mixture.

This paper presented the formulation of a model to describe sequential transfers of oil and oil derivatives in contemplating the possibility that the velocity profiles of the fluids are reversed due to pumping stops. The model was built based on the Theory of Continuous Structured Mixing, which assigns independent kinematics for products and describes an internal structure for the constituents that make up the products. The existence of spatial velocity fields (one field for each product) can describe the motion of the constituents of individual products, thereby providing products and more realistic velocity profiles when pump stops occur. In a thermodynamically consistent way, the model takes into account the effects of inertia forces, gravity and viscosity, and the phenomenon of diffusion in the axial and radial directions.

Contrary to what occurs in the classical formulation, the proposed model explicitly considers the nature of the velocity profile of both products, allowing for proper description of the mixing process in typical situations where pumping is interrupted.

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