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# Analysis of the shear stresses in a filling line of parenteral products: The role of sterilizing filters

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## ABSTRACT

This study investigates the impact of shear stress on protein-based parenteral drug products, particularly during the filling process, with a focus on the filtration unit. A 3D computational replica of a real sterilizing filter was developed using a stochastic geometry generation methodology, able to replicate realistic values of permeability, porosity, and pore size distribution. Computational Fluid Dynamics simulations were employed to analyze fluid dynamics within the filtration unit. The evolution of shear stress for massless tracers was examined to assess average shear stress in the sterilizing filtration unit. The results provide insights into shear stress dynamics and suggest potential scale-down strategies for industrial applications. The presented analysis is novel compared to most estimations of shear stress in sterilizing filtration units, which generally rely on mathematical calculations.

## 1. Introduction

Drug manufacturing consists of a series of operations, generally referred to as formulation, fill and finish. First, the purified form of the drug product is formulated with selected excipients, then it is filled into vials or syringes and is ready for packaging, labelling and quality inspection before distribution [1]. Among the other operations, the filling step represents the most critical one, in which the pharmaceutical drug product undergoes different stages, including mixing, pumping, filtration, and final filling into vials. As filling lines continue to operate under more and more accelerated conditions, there is great concern over stability of protein-based products [2,3]. They may be sensitive to temperature changes, oxidation, light, ionic strength and shear stress [4]. Computational Fluid Dynamics (CFD) investigation of some of the typical unit operations (product flowing in tubing and sampling through fittings) has been already published from our research group [5,6]. However, one of the most challenging steps to analyse is the filtration process. Filtration plays a critical role in guaranteeing the absence of microbiological or particle contamination in the product. This pressure-driven process pushes the product across a semi-permeable membrane in an effort to separate the contaminants from the solution [7] and has been broadly studied in the literature for various applications [8–10].

In the realm of bioprocessing steps, more specifically, there are various types of filtration; among them dead-end filtration has wide use in bioprocessing [11]. In particular, sterile filtration is widely used for parenteral drug products to verify they meet the sterility criterion. It specifically allows the complete retention of a microbiological challenge with  $10^7$  colony forming units/cm<sup>2</sup> of filter area [12]. In accordance with the Food and Drug Administration and the United States Pharmacopoeia guidelines, sterilizing filters should be product compatible [13] and made of non-fiber releasing materials with a nominal rating of 0.2  $\mu$ m and 0.22  $\mu$ m respectively [14,15]. Such membranes can be placed in different housings, depending on the need. Typically, studies for initial screening and preliminary filter sizing are performed using flat membrane discs [16]. At manufacturing scale, on the other hand, capsule [11], cartridge [17] filters with membrane pleating are generally employed; the pleating allows a larger area of the filter medium to be packed in such a way to increase performance in terms of fluid flowrate and total throughput, meaning the amount of fluid which can be processed before the filtration is stopped [18,19]. For the sake of clarity, an example of these membranes is given in Fig. 1. Such membranes have a characteristic thickness of hundreds of  $\mu$ m and a length on the order of cm.

Scale-up is necessary to pass from bench-scale to manufacturing-scale and vice versa for scale-down [17]. Linear scale-up (or scale-

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Nomenclature		$\gamma$	shear rate, $s^{-1}$
$A$	streamline area of influence, $m^2$	$\varepsilon$	porous medium porosity, -
$C$	filtration capacity, $m^3$	$\mu$	dynamic fluid viscosity, $kg\ m^{-1}\ s^{-1}$
$CV$	coefficient of variation, -	$\nu$	kinematic fluid viscosity, $m^2/s$
$d_c$	cell diameter, $m$	$\rho$	fluid density, $kg\ m^{-3}$
$d_w$	window diameter, $m$	$\sigma$	shear stress, $Pa$
$D_{pore}$	mean pore diameter, $m$	$\tau$	residence time, $s$
$EFA$	Effective Filtration Area, $cm^2$	$\tau_f$	fluid time scale, $s$
$g$	gravity, $m\ s^{-2}$	$\tau_p$	particles response time, $s$
$k$	porous medium permeability, $m^2$	<b>Subscripts</b>	
$m$	number of streamlines, -	<i>comm</i>	commercial
$p$	fluid pressure, $Pa$	<i>i</i>	index
$\Delta P/L$	pressure drop over porous medium length, $kg\ m^{-2}\ s^{-2}$	<i>j</i>	index
$Q$	volumetric flowrate, $mL\ min^{-1}$	<i>lab</i>	laboratory
$Re$	Reynolds number, -	<i>ms</i>	microscale
$St$	Stokes number, -	<i>ta</i>	time-averaged
$SF$	Safety Factor, -	<i>theo</i>	theoretical
$t$	time, $s$	<i>tot</i>	total
$u$	fluid velocity, $m/s$	<b>Abbreviations</b>	
$u_{pore}$	pore-scale velocity, $m/s$	CFD	Computational Fluid Dynamics
$V$	Darcy velocity, $m/s$	CLSM	Confocal-Laser-Scanning Microscopy
$w$	volumetric flowrate weight, $m^3/s$	CT	Computed Tomography
$x$	spatial coordinate, $m$	SEM	Scanning Electron Microscopy
<b>Greek letters</b>		TEM	Transmission Electron Microscopy
$\alpha$	coefficient of proportionality, -		

down) assumes that the filtration area increases proportionally to the volume of the solution to be processed; however, a more thorough understanding of scale-up from discs to pleated filters has become necessary and is generally referred to as Safety Factor ( $SF$ ) [20]. It considers both the Effective Filtration Area ( $EFA$ ) and the filtration capacity ( $C$ ), i. e., the cumulative filtrate volume, as follows:

$$SF = \frac{\left(\frac{EFA}{C}\right)_{comm}}{\left(\frac{EFA}{C}\right)_{lab}} \quad (1)$$

where *comm* and *lab* stand for commercial and laboratory scale, respectively.

This factor is introduced to minimize the risk of underestimating the filter area and compensate the variability usually connected with

upscaling processes, which include differences in flow geometries, extra-membrane flow resistances, and fouling phenomena [21]. There is a wide range of potential  $SF$  and a thorough list is reported in Haindl et al. [22] who found values from 0.75 to 2.5 in their review of the literature, although calculations performed on different filters gave them a much higher  $SF$  (7.17).

Membrane characterization is fundamental to gain insights into the porous structure and get medium properties, such as porosity, mean pore size and pore size distribution. In addition, it can be used to inform computer simulations of digital structure reconstruction in an effort to obtain a realistic representation of the porous media. This characterization can be performed through several techniques, including Scanning Electron Microscopy (SEM), tomography, and Confocal-Laser-Scanning Microscopy (CLSM) [23]. In particular, SEM is a magnification tool which makes use of focused beams of electrons to obtain

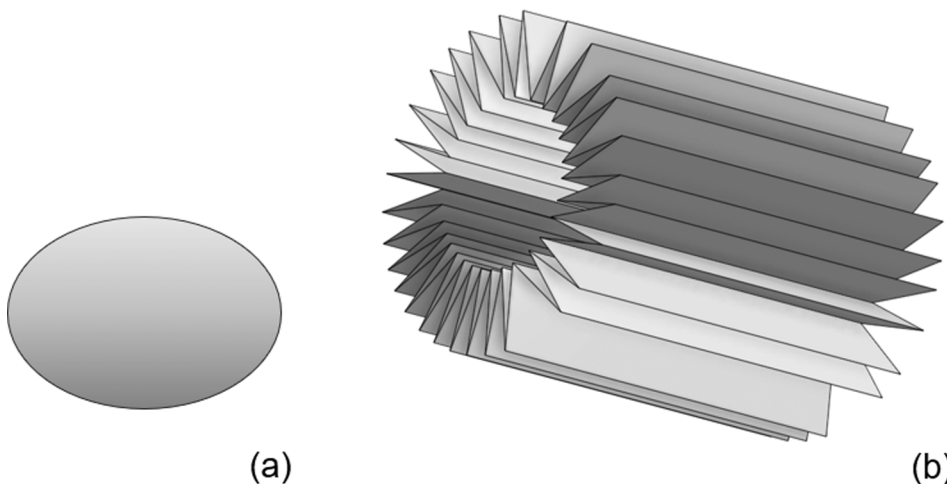


Fig. 1. Example of flat membrane (a) used in the disc configuration and pleated membrane (b) in capsule / cartridge.

information on the porous structure. One of the missing features of SEM is that it only provides morphological information thus making it invaluable in getting 3D information. Similarly, Transmission Electron Microscopy (TEM) can provide high-resolution views of objects but is limited to 2D images and cannot identify 3D morphology of samples [24]. Another valuable technique is the nano/micro-Computed Tomography (CT) which does not need any invasive procedure prior to imaging and allows 3D characterization of the membrane [25]. There are different types of CT scanners that allow for different sample size and resolution, thus providing multiscale opportunities with CT. The typical resolution of micro-CT and nano-CT scanners, for instance, is around 1–100  $\mu\text{m}$  and as low as 0.5  $\mu\text{m}$ , respectively [26]. Similarly, CLSM can be employed for this purpose. After the 2D images of the 3D membrane are obtained, they are used in combination with computer-based image analysis to derive a complete 3D representation of the membrane structure, [23] which can be further used to perform CFD analysis. In a former work, Affandy [27] reports a case of 2D reconstruction of sterile filters passing through TEM images. For 3D reconstruction, on the other hand, the analyses went up to dimensions of 8  $\mu\text{m}$ . As sufficiently high-resolution 3D images of the micro-structure of real porous membranes are difficult to obtain by tomographic methods, detailed studies are more and more based on 3D digitalized models, obtained by various stochastic reconstruction techniques [28–30].

As previously stated, CFD is being utilized more and more in pharmaceutical applications enabling process monitoring and optimization without the requirement for products [31]. In such a realm, two approaches are generally used to deal with a porous medium, named ‘microscale’ and ‘macroscale’ approaches. The first strategy takes into account the actual pore size and morphology inside the membrane or the porous medium in general, which contributes to a complex fluid dynamics behavior of the fluid [32,33]. The macroscale approach, on the other hand, is an averaged representation of the pore-scale process and is a computationally efficient alternative to microscale model [34,35]; however, it is not suitable to detect local fields through the pores, i.e., shear stresses. For this reason, a microscale framework was chosen for the present analysis.

In this manuscript, a 3D reconstruction of a sterilizing membrane used in biopharmaceutical applications is obtained based on experimental data, i.e., pore size distribution, porosity, and permeability. Then, models for the shear stress distribution derived in a former work from our research group [6] are applied in order to assess the average shear stress that the fluid experiences when flowing through the medium in a microscale framework. This led to a relation between the Darcy velocity in the porous medium ( $V_m$ ) and the resulting shear stress, which is applicable to the membrane filter under investigation. Finally, considerations on the applicability of scale-down strategies for this unit operation are provided. To the best of our knowledge, this analysis is brand-new compared to most shear stress estimations in filtration units, which generally rely on mathematical calculations [36,37].

## 2. Governing equations and theoretical background

As suggested elsewhere [38], although the characteristic length of the pores of the membrane is rather small, i.e., in the order of hundreds of nm, the continuum hypothesis still holds true. The transport phenomena inside the pores are governed by the continuity equation and the Navier-Stokes equation. In the case of an incompressible fluid flowing through the void fraction of the geometry with constant density and viscosity, these equations read as follows:

$$\frac{\partial u_i}{\partial x_i} = 0 \quad (2)$$

$$\frac{\partial u_i}{\partial t} + u_j \frac{\partial u_i}{\partial x_j} = -\frac{1}{\rho} \frac{\partial p}{\partial x_i} + \nu \frac{\partial^2 u_i}{\partial x_j^2} \quad (3)$$

where  $u_i$  is the  $i$ th component of the fluid velocity,  $p$  is the fluid pressure and  $\rho$  and  $\nu$  are its density and kinematic viscosity, respectively.

Simulation results are then interpreted based on macroscale continuum equations for fluid flows in porous media. The parameters of these spatially smoothed equations (e.g., porosity and permeability) are derived from averaging procedures performed on the microscopic governing equations of both the porous medium and fluid flow, as explained by Whitaker [35]. Thus, Equations (2) and (3) become:

$$\frac{\partial V_i}{\partial x_i} = 0 \quad (4)$$

$$\frac{\partial V_i}{\partial t} + V_j \left( \epsilon \frac{\partial V_i}{\partial x_j} \right) = -\frac{1}{\rho} \frac{\partial p}{\partial x_i} + g_i + \frac{\nu}{k} V_i \quad (5)$$

where  $V_i$  is the Darcy velocity in the porous medium,  $\epsilon$  the medium porosity, and  $k$  the membrane permeability. Assuming the flow is stationary, the regime is laminar – as demonstrated in the next paragraph – and the gravity effects are unimportant, some terms can be neglected; this leads to the well-known Darcy’s Law written as a 1D expression:

$$\frac{\Delta P}{L} = -\frac{\mu |V|}{k} \quad (6)$$

where  $\Delta P/L$  is the pressure drop over the length of the porous medium and  $\mu$  is the dynamic fluid viscosity.

## 3. Experimental set up

Experimental work was conducted to assess the permeability of a typical sterilizing membrane used for filling purposes. The Durapore membrane [39] was adopted, being a membrane with homogeneous average pore size, without upstream / downstream side, like other membrane in Polyether Sulfone (PES) for instance [40]. It is a hydrophilic Polyvinylidene Fluoride (PVDF) membrane used for applications requiring the highest degree of sterility assurance.

Many different housing configurations are available for such membrane, including cartridges, capsules, and disc membrane [41]. This last membrane has a porosity ( $\epsilon$ ) of 70 %, a depth of 125  $\mu\text{m}$  and a diameter of 47 mm. SEM was used to assess the surface and cross-sectional morphologies of the sterilizing membrane. For cross-section images, the samples were cut in liquid nitrogen using a sharp knife. This analysis provided an idea of the morphology of the structure and proved the uniformity of the pore size distribution. An example of cross-section image is shown in Fig. 2, while Fig. 3 shows the experimental set up used in this study for the estimation of the membrane permeability; a

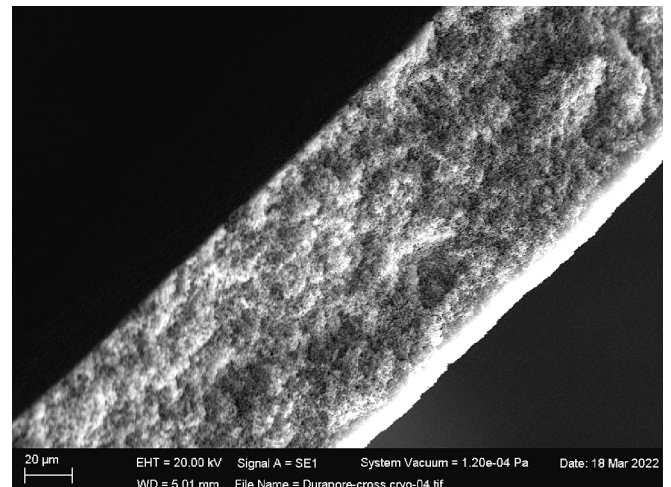
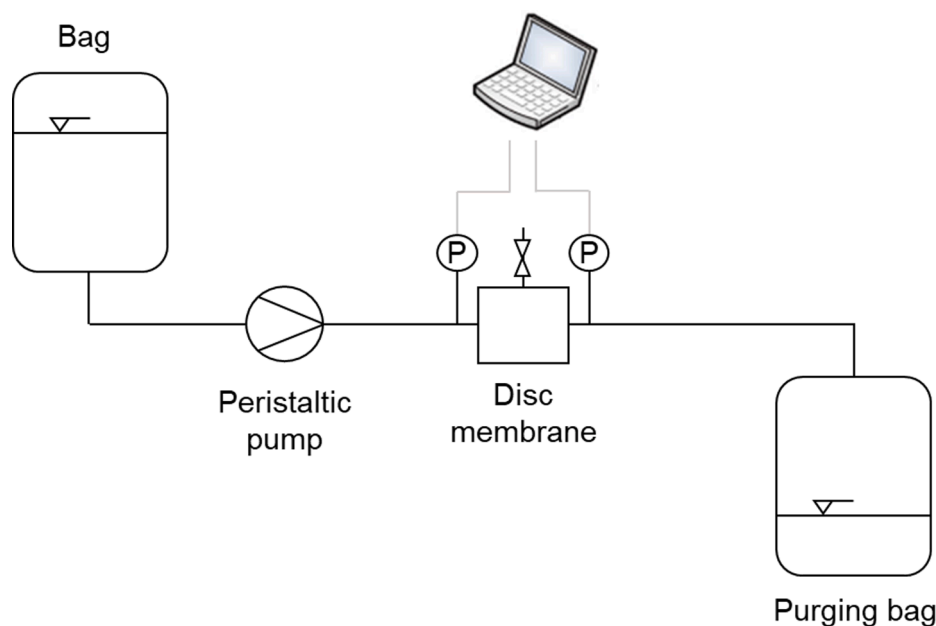


Fig. 2. SEM cross-section of the Durapore membrane.



**Fig. 3.** Experimental set up that has been used for the estimation of the pressure drop across the membrane given a certain flowrate and, thus, its permeability.

Watson Marlow peristaltic pump was used to drive water across the disc (with Durapore membrane) at different flowrates.

Pressure sensors (P) were used to detect the pressure drop across a membrane filter when water was flowing through the circuit at different flowrates. The relationship between the inlet flowrate and the resulting pressure drop was found out to be linear, i.e., laminar condition [42], and the trend is displayed in Fig. 4. Three trials for each flowrate were run on clean membranes to account for experimental variance and the relevant coefficients of variation (CV) are also reported. For clarity, a

higher CV can be observed for the  $20 \text{ mL min}^{-1}$  flowrate, which could be due to an experimental uncertainty in one of the three samples and the limited number of experimental data sets. Nonetheless, although higher than the other flowrates investigated, the CV at  $20 \text{ mL min}^{-1}$  is still small and the overall linear trend in the pressure drop is preserved. Therefore, the pressure drop with the lowest CV was then selected, i.e., that at  $40 \text{ mL min}^{-1}$ , and it was used to calculate the membrane  $k$  using the previously reported Equation (6).

The estimated permeability was  $2.63 \times 10^{-15} \text{ m}^2$ . It must be noted that  $k$  assessed with such experiments considers both the resistance provided by the filter housing and the membrane pores. Nonetheless, as suggested elsewhere [43], the filter housing resistance was assumed to be negligible because the pores undoubtedly make up the major contribution.

#### 4. Numerical set up

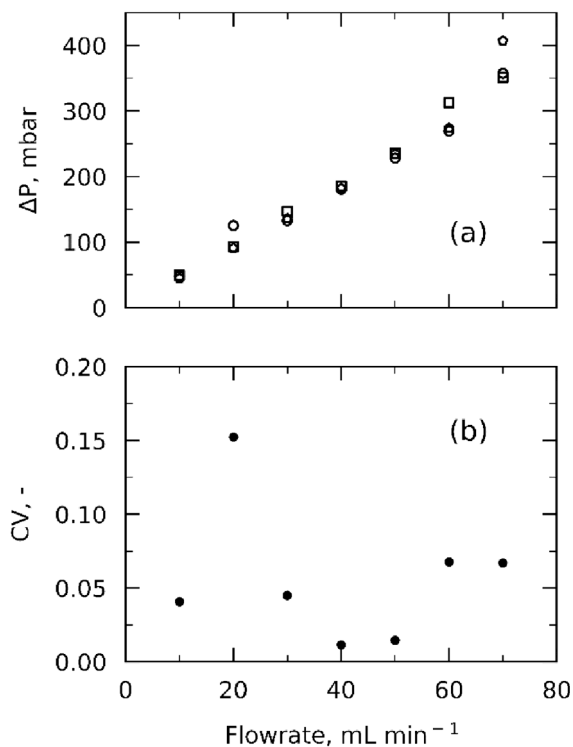
In this work, a computational replica was achieved using the code developed previously by one of the authors and available in [44], which makes use of the software plugim! (<https://www.plugim.fr/>). For the computational analysis of fluid flow, on the other hand, the finite volume method-based open-source code OpenFOAM 9 (<https://github.com/OpenFOAM/OpenFOAM-9>) was used. The simulations were carried out on an HP workstation with eight logical processors and 16 GB of RAM. Python 3.8 was used for the post-processing of the simulations. For the sake of clarity, the details for the membrane and grid generation are reported in two separate sections (Sections 4.1 and 4.2).

##### 4.1. Membrane generation

The workflow used for the development of the computational replica [44] is composed of three distinct processes, including sphere aggregation, Voronoi tessellation, and concluding morphological operations. The number of spheres for the aggregation with the relative cell diameter ( $d_c$ ), the repulsion factor ( $Repul$ ), the size of the spheres and cylinders for the Voronoi tessellation ( $R_{node}$ ,  $R_{strut}$ ), and the resolution of the box were some of the parameters that were tuned.

Some details of the membrane are represented in Fig. 5, which zooms on the representative cell.

As a starting point, it was assumed that the given pore size equals the



**Fig. 4.** Pressure drop vs flowrate for three different data sets through the Durapore membrane in (a). Each data set is represented by a geometric figure (circle, square, pentagon). Coefficients of variation CV are reported in (b).



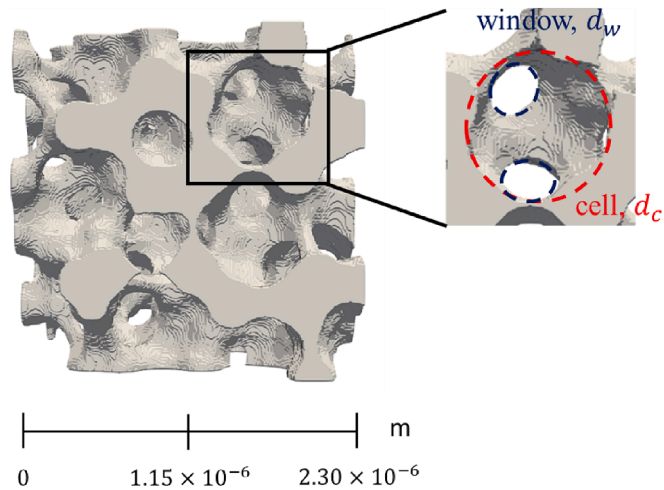


Fig. 5. Cell unit of the membrane showing its components.

window diameter ( $d_w$ ), which delimits the area where the fluid passes through. The relation between cells diameter and window diameter was found in Inayat et al. [45]:

$$d_w = \frac{d_c}{2.3} \tag{7}$$

and it was therefore used to assess  $d_c$  for the initial sphere aggregation.

Based on [44], the dimension of the box was chosen equal to five times the  $d_c$ ; this value was found high enough so as to ensure the chosen computational domain is a representative elementary volume, which is defined as a volume surrounding a point in which all averaged attributes are independent on the size of the volume [46].

The structure was developed to meet the actual  $\varepsilon$  (provided by manufacturer) and experimental  $k$ . An iterative workflow was used until the desired  $\varepsilon$  and  $k$  were both achieved (Fig. 6).

For the sake of clarity, Table 1 reports the dimensionless parameters that yielded the desired values for  $\varepsilon$  and  $k$ .

Further analysis on the membrane were carried out, using the por-eSpy module available in python, which is a toolkit for microstructure investigation of porous media [47]. The resulting pore size distribution is shown in Fig. 7, and it is consistent with experimental data from the literature [48].

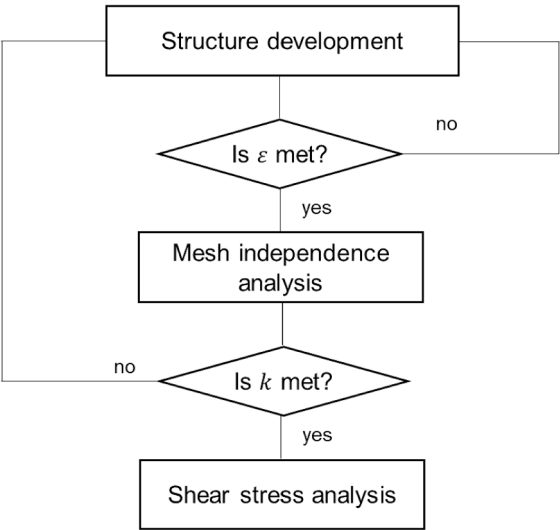


Fig. 6. Numerical set up.

Table 1  
Dimensionless parameters used in structure development.

Parameter	Meaning	Value
<i>NB</i>	Initial number of spheres	380
<i>Alpha</i>	Compactness parameter	0.55
<i>Beta</i>	Compactness parameter	0.75
<i>Repul</i>	Repulsion	20
<i>R</i>	Sphere diameter	25
<i>W</i>	Box size	250
<i>R<sub>node</sub></i>	Sphere diameter for the tessellation	17
<i>R<sub>strut</sub></i>	Cylinder diameter for the tessellation	6
<i>Operation</i>	Morphological operation	Closing
<i>Iter</i>	Number of iterations for the tessellation	1
<i>tv</i>	Size of the chosen morphological operation	19
<i>vX<sub>res</sub></i>	Voxel resolution	$9.2 \times 10^{-9}$

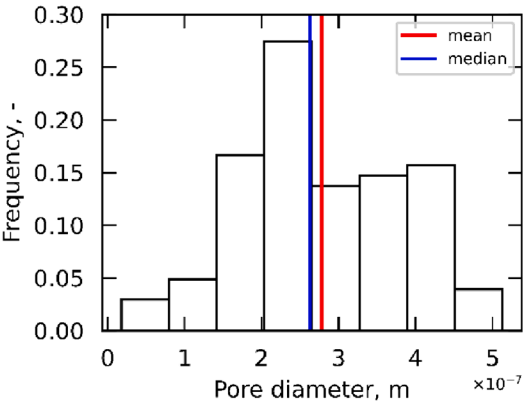


Fig. 7. Pore size distribution. The mean and median values ( $2.8 \times 10^{-7}$  and  $2.6 \times 10^{-7}$  m, respectively are also shown).

4.2. Grid generation

Pre-mixing and post-development regions are added to ensure development of the flow prior to and past the actual membrane, as

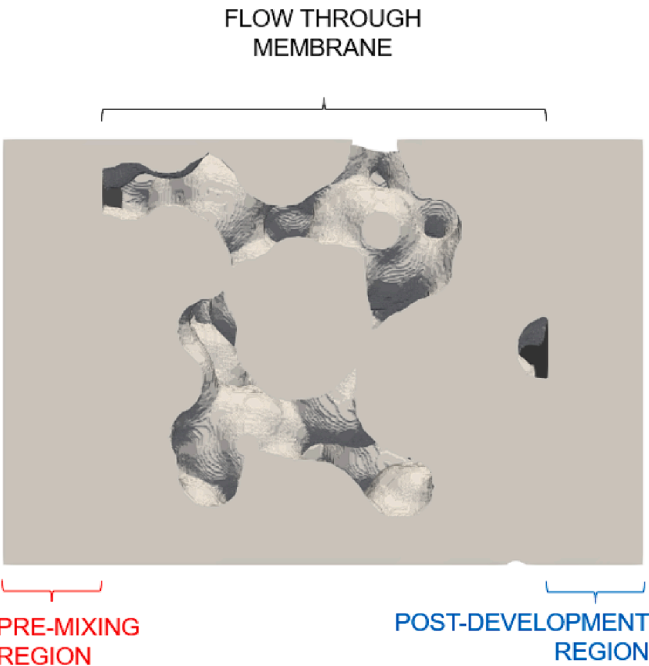


Fig. 8. Computational domain. Pre-mixing and post-development regions are shown.

shown in Fig. 8 where the computational domain is displayed.

Special emphasis should be given to the boundary conditions, summarized in Table 2.

A Newtonian incompressible fluid with water-like properties [49] was considered. Steady-state conditions were assumed and the solver *simpleFoam* was used to perform the fluid flow simulations. Based on the initial experimental investigation reported in Section 3, laminar regime was considered, where Reynolds number ( $Re$ ) was evaluated as follows:

$$Re = \frac{\rho u_{pore} D_{pore}}{\mu} \quad (8)$$

where  $u_{pore}$  is the pore-scale velocity and  $D_{pore}$  is the mean pore diameter of the membrane, defined as the diameter of a sphere having the same volume as the mean pore. Darcy's velocity (or surface velocity) in the porous medium ( $V_{ms}$ ) was chosen from standard commercial filtration values using the following scaling equation:

$$\left( \frac{Q}{EFA} \right)_{comm} = \left( \frac{Q}{EFA} \right)_{ms} \quad (9)$$

where  $EFA$  is the Effective Filtration Area of the filter. For the commercial scale, capsules with  $EFA$  864 and 6000 cm<sup>2</sup> respectively were considered and several operating conditions were tested ( $Q = 2500$  mL min<sup>-1</sup>, 1700 mL min<sup>-1</sup>).

The membrane grid was developed by using two OpenFOAM utilities, named *blockMesh* and *snappyHexMesh*. The former is used to build a background structured hexahedral mesh, which is then used as a starting point to build a body-fitted grid via *snappyHexMesh*. The final grids were mainly comprised of hexahedral cells in such a way to ensure a good numerical performance of the solver. The grids were further refined close to the solid walls of the pores, thus resulting in cells of decreasing dimension moving from the pores walls to the bulk of the fluid.

A grid independence analysis was conducted to find the optimal mesh, resulting from a trade-off between computational cost and solution's accuracy. This was done by tuning two main parameters: the number of background cells in the initial background mesh and the level of refinement  $Ref$  performed by *snappyHexMesh* on the cells close to the solid pore surface. The values of the refinement level taken into account were  $Ref = \{0, 1, 2\}$ . The grid independence study was conducted by monitoring the pressure drop across the membrane, and the results are presented in Fig. 9, where the relative errors to the most computationally expensive grid are also shown for comparison.

It can be noted that  $Ref = \{0\}$  does not lead to an accurate solution. Therefore, a more computationally expensive mesh should be used. On the other hand, grids corresponding to  $Ref = \{1, 2\}$  and a number of cells higher than  $2 \times 10^6$  seem to result in more accurate and converging solution. For the subsequent analysis, the mesh with  $Ref = \{2\}$  and around  $1.9 \times 10^6$  cells was used. This led to a  $k$  of  $2.55 \times 10^{-15}$  m<sup>2</sup>, with a small percent relative error with respect to the experimental value reported in Section 3, i.e., 3%.

Having as objective the evaluation of viscous stresses, an investigation of the microscale of the membrane became necessary. As mentioned, macroscale models solve equations for macroscopic, and thus averaged, quantities. Therefore, when a closer look at local quantities at the scale of the pores is needed, microscale investigations like

the present one need to be performed.

## 5. Results

### 5.1. Streamlines and shear stress analysis

Drawing on earlier works [5,6], innovative strategies were used to track biological particles of fluid through their trajectories and detect their relevant fields. This made it possible to develop shear stress statistics which accounted for the actual shear stress history of each individual particle. Therefore, the Stokes number ( $St$ ) for these particles was initially estimated; this correlates the particles response time ( $\tau_p$ ) and the fluid time scale ( $\tau_f$ ), as follows:

$$St = \frac{\tau_p}{\tau_f} \quad (10)$$

The biological particles in the drug products have a density similar to water [49], and their size is smaller than 200 nm [50]. This results in  $St$  below the unity for all the operating conditions described. In such conditions, biological particles are expected to follow fluid streamlines closely and, therefore, massless tracers were used for the investigation. More specifically, the utility 'Stream tracer with custom source' available in ParaView [51] was utilized, whose seed points were randomly given as an input in the pre-mixing region. To provide an example, Fig. 10 shows the streamlines for 200 seeds.

Local fields across each streamline were then acquired through post-processing python scripts. A typical pore-scale velocity path is displayed in Fig. 11 (a); local velocity and shear trends for such streamline can be observed in Fig. 11 (b).

It can be noted from Fig. 10 that local fields greatly vary through the geometry because of the complexity (and tortuosity) of the membrane structure; therefore, as cumulative definitions appeared too sensitive to domain size and hardly applicable [6], an average shear stress per streamline is calculated, as reported in Equation (11). The trapezoidal rule was used to estimate the integral and implemented through the 'trapz' function available in the numpy module in python. The averaged shear stress  $\sigma_i$  per streamline was therefore evaluated by time-averaging the shear stress along the streamline and using the relative flowrate  $w_i$  as weight.

$$\bar{\sigma}_i = \sum_i \left( \left( \frac{1}{\tau_i} \int_0^{\tau_i} \sigma_i(t) dt \right) \cdot \frac{w_i}{w_{tot}} \right) \quad (11)$$

$$w_i = A_i \frac{1}{\tau_i} \int_0^{\tau_i} u_i(t) dt \quad (12)$$

with  $u_i(t)$  being the velocity at the relevant time step,  $i$  the streamline index. For the sake of clarity, the time-averaged shear stress introduced in Equation (11) will be referred to as  $\sigma_{ia,i}$ .

The average shear stress  $\bar{\sigma}$  is then simply assessed as follows:

$$\bar{\sigma} = \frac{1}{m} \sum_i \bar{\sigma}_i \quad (13)$$

where  $m$  is the number of streamlines.

This weighing operation helped to consider streamline contributions to shear stress in a manner proportional to the actual flowrate through that specific point in the membrane.

An analysis was first conducted to find the minimum number of particles to release in order to fully describe the whole fluid dynamics within the membrane. A varying number of seeds was then released in the pre-mixing region and  $\bar{\sigma}$  was monitored, as displayed in Fig. 12. Twenty trials with varying seed location were run for each number of seeds, and the related CV is reported.

The higher the number of seeds was, the lower the uncertainty span in  $\bar{\sigma}$  was. From 200 seeds onwards, CV was bounded between 0 and 0.025 and, therefore, the number of seeds can be claimed sufficient to

**Table 2**

Boundary conditions for the main patches, expressed in terms of OpenFOAM classes.

Patch	Boundary Condition	
	$p$	$u$
inlet	zeroGradient	fixedValue ( $\bar{u}$ )
outlet	fixedValue (0)	zeroGradient
wall	zeroGradient	noSlip
sides	symmetry	symmetry

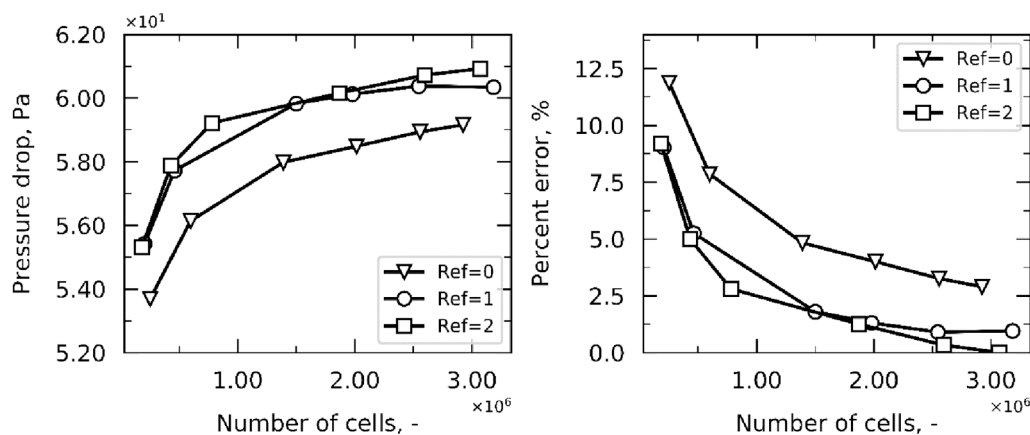


Fig. 9. Grid independence analysis.

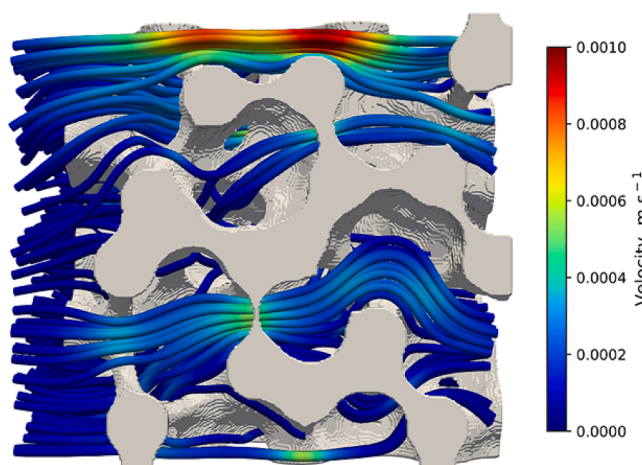


Fig. 10. 200 streamlines are followed through the porous medium and the local velocities are displayed.

describe the fluid dynamics within the membrane. 460 streamlines were finally chosen as the increase in the seeds number did not lead to a significant increase in the computational expense. This can be observed in Fig. 12, where the time to compute each point is displayed; as expected, the computation time varies linearly with increasing number of streamlines but is still way lower than the overall simulation time (which is in the order of hours instead).

Fig. 13 shows the distribution of the time-averaged shear stress in the

case of 460 seeds. It can be noticed that the higher shear stresses were experienced only by a small portion of the solution because the majority was subjected to lower shear stresses.

## 5.2. Industrial considerations

The previously reported analysis was then repeated for other velocities and other filtration unit *EFA*, representative of potential manufacturing units, in order to investigate the relation between the inlet velocity and the resulting shear. To allow for comparison with literature data, the shear rate  $\dot{\gamma}$  is estimated, which is simply  $\sigma/\mu$  being incompressible fluid. The relevant results are reported in Table 3 where the relevant  $\dot{\gamma}$  for each  $Q_{comm}$  is reported for comparison. In addition,  $Re$  is also estimated using Equation (8), where  $u_{pore}$  is estimated based on the membrane porosity as follows:

$$u_{pore} = \frac{V}{\varepsilon} \quad (14)$$

Therefore, being the flow through the pores laminar, a linear relationship is found between the inlet velocity and the resulting shear rate and is represented by  $\alpha$ . For the three runs, the percent difference in  $\alpha$  is below 5%.

A comparison with literature data was finally drawn. Some mention shear rate around  $10,000 \text{ s}^{-1}$  [52,53], which is in line with our estimates for the last operating conditions considered in Table 3. These values are generally calculated assuming that the pore is a cylindrical tube with mean diameter  $D_{pore}$ :

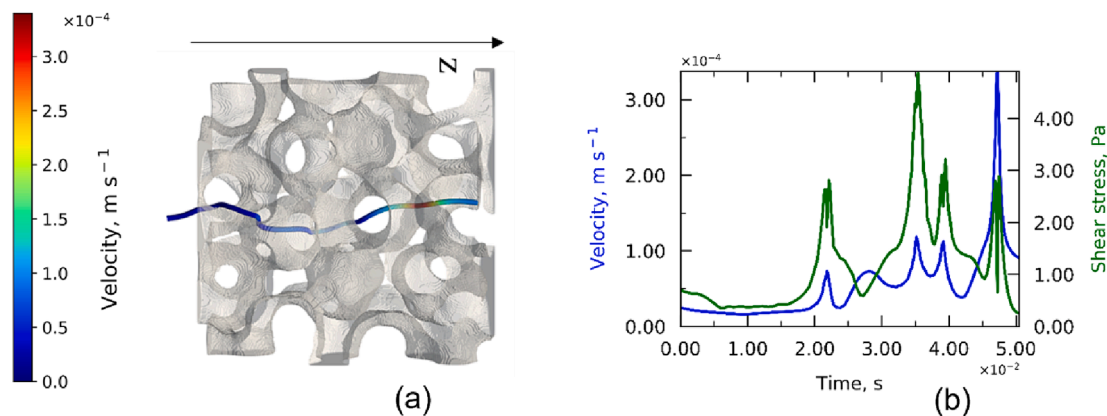
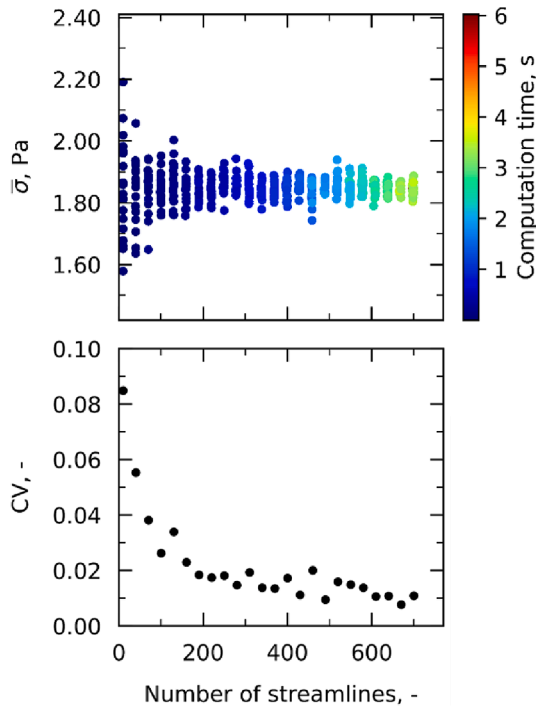
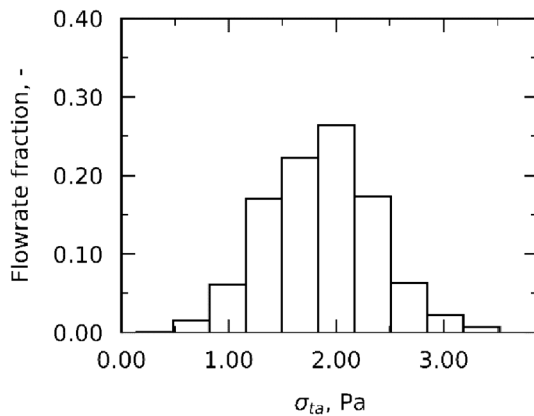


Fig. 11. Streamline for a hypothetical particle released in the pre-mixing region (a), with Z being the flow direction. Velocity (in blue) and shear stress (green) are displayed (b). (For interpretation of the references to color in this figure legend, the reader is referred to the web version of this article.)





**Fig. 12.** Average shear stress  $\bar{\sigma}$  is displayed while varying the number of streamlines and the number of trials to ease the visualization of the analysis in the top plot. Each point in the data set is colored based on its computation time (see color scale). Coefficients of variation CV are reported in the bottom plot.



**Fig. 13.** The time-averaged shear stress ( $\sigma_{ta}$ ) distribution for the analyzed filtration unit is displayed using the flowrate fraction ( $w_i$ ) normalized over  $w_{tot}$  as weight.

**Table 3**  
Results of the analysis for different commercial flowrates.

$Q_{comm}$ mL min <sup>-1</sup>	$EFA_{comm}$ cm <sup>2</sup>	$V_{ms}$ m/s	$Re$	$\bar{\gamma}$ s <sup>-1</sup>	$\bar{\gamma}_{theo}$ s <sup>-1</sup>	$\alpha$
2500	6000	$6.94 \times 10^{-5}$	$2.78 \times 10^{-5}$	$1.85 \times 10^3$	$2.40 \times 10^3$	$26.7 \times 10^6$
1700	864	$3.15 \times 10^{-4}$	$1.26 \times 10^{-4}$	$8.37 \times 10^3$	$1.09 \times 10^4$	$26.6 \times 10^6$
2500	864	$4.63 \times 10^{-4}$	$1.85 \times 10^{-4}$	$1.29 \times 10^4$	$1.60 \times 10^4$	$27.7 \times 10^6$

$$\bar{\gamma} = \frac{16u_{pore}}{3D_{pore}} \quad (15)$$

Additionally, the workflow proposed in this manuscript allows for post-processing analysis to derive temporal and distribution information, which traditional equations (such as Equation (15)) do not provide.

As expected [52], shear rates (and stresses) resulting from sterilizing filtration units are higher than those resulting from other filling line operations, such as flow through tubing and fittings, whose investigation was the focus of other works from our research group [5,6]. To provide some examples from these studies, shear rates of around  $8 \times 10^2 \text{ s}^{-1}$  were found for flow in smooth tubing under turbulent conditions (flowrates around  $2140 \text{ mL min}^{-1}$  in 9.53-mm-tubing); similar values of shear rates were found for Y-fittings under the same operating conditions while higher for T-fittings (i.e.,  $1.2 \times 10^3 \text{ s}^{-1}$ ). Under laminar conditions (flowrates around  $180 \text{ mL min}^{-1}$ ), on the other hand, much lower shear rates were monitored ( $2 \times 10^1 \text{ s}^{-1}$ ).

A final remark can be given regarding scaling approaches, which, as mentioned in the Introduction, are critical to ensure representativeness between laboratory and commercial scales without over- or under-estimating the system. As previously stated, they use appropriate scaling factors that account for differences in flow resistance, fouling phenomena, and flow geometries. Without exploring such macroscopic differences, however, the present work offers a formula for ensuring, at the microscopic level, that the product is exposed to the same shear stress in the laboratory compared to the commercial scale. For this particular filtration unit, being the flow laminar through the pores, the application of Equation (16) can guarantee that the local shear stresses at the microscale level are preserved if the same membrane type is used in the laboratory scale unit. Indeed, the ratio between the two quantities might differ from unity if other factors besides shear stresses are considered.

$$\left(\frac{Q}{EFA}\right)_{comm} = \left(\frac{Q}{EFA}\right)_{lab} \quad (16)$$

## 6. Conclusion

In the present work, CFD simulations were employed to investigate the fluid dynamics within a typical sterile filtration unit. A computational replica was obtained through stochastic membrane reconstruction to meet porosity, pore size distribution, and permeability of the real filter. Shear stress analyses were conducted and the average shear exposures of the product at different operating conditions were estimated. These values were compared with those derived from traditional simplified equations; not only are the values consistent, but the present workflow allows post-processing analyses (on residence time, shear stress distributions) that traditional equations do not provide. Also, as expected, when comparing these shear stresses with those from other components of the filling line, i.e., tubing, fittings, they appeared much higher. Considerations were also drawn on the scale-down operation for the pharmaceutical industry.

The methodology we proposed can be used to explore other areas. For instance, one could evaluate particle impingement in pores during long-term processes, where aggregation phenomena are prominent, or focus on regions of high shear stress. The work could further be extended to membranes with non-uniform pore distributions or pleated membrane deformation could be studied to make the analysis even more meaningful.

This research lays the foundation for future development prospects. The impact of monitored shear stress on product stability and separation efficiency could be investigated.

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## CRediT authorship contribution statement

**Camilla Moino:** Writing – review & editing, Writing – original draft, Visualization, Validation, Software, Methodology, Formal analysis, Conceptualization. **Enrico Agostini:** Writing – review & editing, Visualization, Validation, Software. **Andrea Albano:** Writing – review & editing, Methodology. **Marco Bellini:** Writing – review & editing, Methodology, Conceptualization. **Erwan Bourlès:** Writing – review & editing, Methodology, Conceptualization. **Bernadette Scutellà:** Writing – review & editing, Methodology, Conceptualization. **Gianluca Boccardo:** Writing – review & editing, Software, Methodology, Conceptualization. **Roberto Pisano:** Writing – review & editing, Methodology, Conceptualization.

## Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

## Data availability

Data will be made available on request.

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