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Laminar flow and conjugate heat and mass transfer in a hollow fiber membrane bundle used for seawater desalination

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ABSTRACT

A hollow fiber membrane-based bundle or humidifier (HFMH) is used for air humidification in the process of membrane type seawater desalination, where seawater flows inside the fibers and the process air stream flows across the fiber bundle. The HFMH has a staggered arrangement of cylindrical fiber tubes with equal transverse and longitudinal pitches. The forced convection heat and mass transfer in the cross-flow tube bundle under naturally formed boundary conditions are investigated numerically on a periodic computational cell. A membrane-based air humidification-dehumidification type sea-water desalination (MHDD) system is constructed and used to validate the model. The mathematical model proposed here takes into account of the fiber-to-fiber interactions. The typical velocity vector profiles, temperature contours as well as the concentration contours for both the air stream and saline water stream are plotted to disclose the heat and mass transfer mechanisms. The effects of air Reynolds numbers (ranging from 50 to 300) as well as the module packing fractions (ranging from 0.126 to 0.503) on the fluid flow and heat mass transfer properties of the module are investigated. A set of correlations are also proposed for the prediction of the friction factor, mean Nusselt numbers and Sherwood numbers in air side. These fundamental data is of interest for the design and optimization of the HFMH, a promising membrane type seawater desalination system.

1. Introduction

An increasing amount of water is demanded by many developed Countries (USA, EU, Japan, etc.) as well as highly crowded emerging countries (China, India, Brazil, etc.) [1]. Fresh water shortage is becoming a worldwide challenge because of the rapid growth of global population and the accelerated industrialization and urbanization [2]. Membrane distillation (MD), an emerging membrane separation technology used in desalination, is being studied worldwide as an alternative solution to overcome the fresh water shortage. Compared with the traditional desalination technology such as Multi Stage Flash (MSF) and Reverse Osmosis (RO), MD can work well at lower operating temperatures and hydrostatic pressures, which means that it can be powered by low-grade solar energy and/or industrial waste heat. Accordingly, MD claims to be a cost- effective and environmental-friendly alternative to the existing commercial desalination technologies [3,4].

MD is a thermally driven separation process based on the principle of vapor-liquid equilibrium and coupled heat and mass transfer. MD for desalination recovers pure water vapor by passing hot brine on one side of a porous hydrophobic membrane which prevents the solution from passing through the membrane micropores, while the generated vapor can pass through due to a difference in vapor pressure between two sides of the membrane, which can be established by a difference in temperature. Based on the type of the condensation method used in the other side of the membrane in MD process, MD systems can be mainly classified into four different configurations: Direct contact membrane distillation (DCMD), vacuum membrane distillation (VMD), air gap membrane distillation (AGMD) and sweep gas membrane distillation (SGMD) [4,5]. Among them, DCMD is the simplest configuration because the hot feed and the cooler is in direct contact with the membrane without an external condenser and thus be researched widely. However, the main disadvantage of this configuration is that the direct contact with the condensing solution significantly increases the sensible heat losses of the hot feed

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Nomenclature

A	area (m^2)
A_m	fibers membrane area in the air stream side
A_v	packing density (m^2/m^3)
A_{tot}	total membrane areas of the module (m^2)
d	diameter (m)
d_h	hydraulic diameter of the flow channels (m)
D_{ws}	water diffusivity in saline water (m^2/s)
D_{va}	moisture diffusivity in air (m^2/s)
D_{vm}	moisture diffusivity in membrane (m^2/s)
f	friction factor
h	convective heat transfer coefficient ($kW\,m^{-2}\,K^{-1}$)
H	module height (m)
H_{eva}	evaporation heat of water (kJ/kg)
J/m_v	moisture emission rate ($kg\,m^{-2}\,s^{-1}$)
L	fiber tube length (m)
k	convective mass transfer coefficient (m/s)
n_f	number of fibers
Nu	Nusselt number
p	pressure (Pa)
p_{atm}	atmospheric pressure
Pe	Peclet number
P_D	diagonal pitch
P_L	longitudinal pitch
P_T	transverse pitch
Pr	Prandtl number
r	radius (m)
Re	Reynolds number
RH	Relative humidity (%)
Sc	Schmidt number
Sh	Sherwood number
T	temperature (K)
u	velocity (m/s)
V_a	volume flow rate of the air (m^3/h)
V_s	volume flow rate of the saline water (L/h)
W	module wide (m)
$x/y/z$	coordinates for physical domain (m)
X	mass fraction of water in saline water (kg water/kg saline water)

Greek letters

δ_m	membrane thickness (m)
ε/η	coordinates for computational domain (m)
ζ^* a	dimensionless temperature for periodic boundary condition
λ	heat conductivity ($W\,m^{-1}\,K^{-1}$)
ν	kinematic viscosity (m^2/s)
ξ^* a	dimensionless humidity for periodic boundary condition
ρ	density (kg/m^3)
φ	packing fraction
ψ	variable
ω	humidity (kg moisture/kg dry air)

Superscripts

$*$	dimensionless

Subscripts

a	air
b	bulk
c	cross section
Cal	calculated
e	equilibrium
eva	evaporation
h	heat
H	uniform heat flux condition
i	inlet
m	mass, membrane
max	maximum value
o	outlet
s	saline water
tot	total
T	uniform temperature condition
Test	tested
v	vapor
w	wall, water
x	x axis direction
y	y axis direction

through the membrane. AGMD can diminish the heat losses because a layer of stagnant air is introduced between the permeate side of the membrane and a condensing surface. Obviously, in this configuration the air gap increases the mass transfer resistance, which result in a lower permeate flux. To reduce the mass transfer resistance of AGMD because of the air gap, the stagnant air can be replaced by a cold sweeping gas flowing through the permeate side of the membrane. Then the water vapor is carried outside and cooled in a condenser (SGMD). The stagnant air can also be replaced by a vacuum pressure (VMD). In this case the condensation also takes place outside the module and the heat losses through the membrane as well as the mass transfer resistance are reduced even more. However, the risk of membrane wetting becomes larger due to the higher pressure difference across the membrane. Moreover, besides the external condenser, a vacuum pump is also needed to maintain the vacuum pressure. For the cases of MD driven by atmospheric sweep air, the membrane module (or membrane contactor) is usually used to generate water vapor from the saline water, which is also called the membrane type humidification process.

Recently, membrane contactors with plenty of module geometries including parallel-plate module, spiral module and hollow fiber membrane module, have been demonstrated in a range of liquid/liquid and gas/liquid separation applications in pharmaceuticals, food processing and desalination [6,7]. Though the parallel-plate membrane module is simple in structure and easy to be fabricated, the packing density of the module is about 500 m^2/m^3, which is not large enough [8]. On the contrary, due to the higher packing density and thus a higher heat and mass transfer capability, hollow fiber membrane contactors made of various materials and structures are more attractive and they have been employed as humidifiers for desalination of saline water and/or brackish water in many studies [9–11]. Song et al. [9] proposed a small desalination plan based on hollow fiber membrane module for crossflow direct contact membrane distillation (DCMD) in which cold distillate flows inside the hollow fibers while hot brine flows outside the hollow fibers in crossflow arrangement. As the shell-side velocity of 90 °C brine is increased to 900 cm/min, a water flux of 50 $kg\,m^{-2}\,h^{-1}$ is achieved. Cheng et al. [10] numerically modeled a hollow fiber-based module for air gap membrane distillation (AGMD) for desalination and concluded that an increase in module packing density by reducing the air gap thickness will enhance membrane productivity. When inlet hot brine and inlet cold water temperatures are at 77 °C and 25 °C respectively and the maintained air gap thickness is 0.2 mm, the highest mean permeate flux reported in their calculations is around 6 $kg\,m^{-2}\,h^{-1}$. Sun et al. [11]