Comparative Study of CWO of Phenols in Falling-Film and Back-Mix Reactors

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The present study was classified into two sections. In the first, a kinetic analysis of the oxidation of phenol in aqueous solution over a supported (0.7% Pt)/Al_2O_3 catalyst was investigated at atmospheric pressure in a batch operating system. The kinetic analysis proved that the reaction consists of two mechanisms, and that the initial rate and steady state activity regimes which exhibited first order behavior with respect to phenol concentration. The reaction rates show an unusual dependence on catalyst loading which suggested a heterogeneous–homogenous free radical mechanism. Phenol removal can be increased by increasing the amount of oxygen gas but at higher flow rates of oxygen a retarding effect of oxygen on phenol oxidation was observed. In the second section, a comparative study of the catalytical wet oxidation (CWO) of phenol in two different types of flow reactors (i.e., falling film and back mixing reactors) was carried out and design parameters such as inlet temperatures, residence time of reactants and catalyst loading in the reactors were used to establish similarity reaction conditions in the two reactors. The study supports the following conclusions: The oxidation rate of phenol was low because of the solubility of oxygen under atmospheric conditions. At low flow rates of liquid reactant the falling film reactor showed a better performance as a result of its lower resistance to mass and heat transfer while the result is completely different at higher liquid flow rates. Non-isothermal operation showed that water evaporation has a strong impact on phenol conversion and must be taken into account in scale up and adiabatic CWO reactor design. Neglecting evaporation can lead to erroneous calculation of the exit stream phenol conversion and temperature. The power law technique has been utilized to correlate the phenol conversion with the operating parameters in the two reactors.

Introduction

Oxidation of dilute aqueous solution of organic pollutants with or without catalysts offers an attractive alternative process to biological oxidation as a mean to purify liquid water. This is particularly true when the pollutant is toxic, hazardous or in some cases non-biodegradable (refractory). Furthermore, liquid phase oxidation is definitely more cost effective as compared with incineration (Joshi *et al.*, 1985). The petrochemical, chemical and pharmaceutical industries produce waste water containing organic pollutants; such as phenols; and oxalic acid, which are toxic to aquatic life. These pollutants are difficult to oxidize biologically, since biological processes are very time consuming and operate well only for relatively dilute organic wastes. The use of chemical oxidation may be especially favored when the pollutants concentration are too high for direct biological systems and/or the effluent has higher temperatures (Harmankaya and Gündüz, 1998). The choice of a particular method depends on, among other factors, the nature of the pollutant, the desired removal efficiency, and cost. Many researchers have reported that among the wastewater treatment techniques, catalytic wet oxidation (CWO) of organic wastes in water seems to be effective and promising (Mishra *et al.*, 1995; Matatov-Meytal and Sheintuch, 1998; Luck, 1999). CWO is a reaction involving an organic compound in water and oxygen over a catalyst. Although many studies have shown that metal oxide catalysts of transition metals like Zn, Cu, and Mn are very effective in the removal of organic wastes, the use of noble metal catalysts for liquid phase oxidation is preferred since no leaching or dissolution of the active metal occurs even under hot acidic conditions (Gallezot, 1997; Luck, 1999). Among the noble catalysts reported for liquid phase oxidation, platinum supported catalysts seem to be promising. Platinum catalysts are well known to be effective for aqueous phase oxidation of alcohols (Mallat and Baiker, 1994; Besson and Gallezot, 2000; Kluytmans *et al.*, 2000) and ammonia (Ukropec *et al.*, 1999). However, there is still little information on the application of platinum catalysts for CWO of organic pollutants (Gallezot *et al.*, 1996; Harmsen *et al.*, 1997; Chollier *et al.*, 1999; Gomes *et al.*, 2000). One of the important factors in CWO is the selection of a suitable reactor. Most laboratory studies have been carried out either in slurry or fixed bed reactor. Pintar and Levec (1994); and Stüber *et al.* (2001) have carried out a comparison between these two reactor types. They reported that the latter is more advantageous in terms of process selectivity and stability. They concluded that agitated reactors, such as slurry or spinning basket reactors, exhibited high liquid-to-catalyst ratio which adversely favors parallel homogeneous

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polymerization reactions, thereby leading to catalyst activity loss and lower selectivity towards complete mineralization products.

Levec and Smith (1976) have reported the kinetics of catalytic oxidation of formic acid (CuO–ZnO catalyst) and acetic acid (Fe_2O_5 catalyst), using trickle bed reactors. The activation energy and effectiveness factors for the catalysts have been reported. The rate data have also been explained on the basis of the oxidation–reduction mechanism.

Pal *et al.* (1982) have shown that when the size of catalyst particles is less than that of the liquid film, a substantial increase (by a factor of 10) in the rate of oxidation of sodium sulfide can be obtained even when, in the absence of the catalyst, the reaction occurs in the diffusion film. They used a theoretical analysis to compare the performance of bubble columns, sectionalized bubble columns, horizontal sparged contactor, packed columns and pipeline reactors. The optimum values of height-to-diameter ratio and the total pressure in the case of bubble column were also given.

Pintar and Levec (1994) studied the oxidation of aqueous phenol in a slurry reactor over a catalyst comprising CuO, ZnO and gamma alumina above atmospheric pressure at up to 13 bars. Their rate measurements showed that the reaction progressed autocatalytcally. Pintar and Levec (1994) also studied the liquid-phase oxidation of phenol in a differential liquid filled fixed bed reactor operated over a zinc, copper and cobalt–oxide catalyst. Roy *et al.* (2010) presented a review paper on the use of platinum catalysts in bubble column reactor for catalytical wet air oxidation (CWAO) of oxalic acid. It has been shown that temperature, pressure, relative gas flow rate and pH evolution during oxidation are important parameters to be taken into account when predicting the performance of wet air oxidation unit.

The purpose of the present work was to perform the kinetic analysis of phenol oxidation in aqueous solution over a supported Pt catalyst at atmospheric pressure and to carry out a comparative study of the CWO of aqueous phenol in two different types of continuous flow reactors (i.e., falling film and back mixing reactors) by investigating the effect of operating parameters such as temperature, liquid and oxygen-feed rates and initial phenol concentration on the performance.

1. Mechanism and Reaction Pathways

According to Gallezot (1997) catalytic oxidation of organic molecules can proceed via different mechanisms, namely:

1 enzymatic oxidation,
2 free radical auto-oxidations initiated by transition metal cations,
3 metal ion oxidation of coordinated substrates,
4 oxygen transfer to the substrate mediated by metaloxo or peroxo complexes, and
5 oxidative dehydrogenation on metal surface.

In alcohol oxidations, oxidative dehydrogenation on metal surfaces is commonly reported (Mallat and Baiker, 1994; Gallezot, 1997). The mechanism of alcohol oxidation on a noble metal catalyst involves the dehydrogenation of the organic substrate on the metal surface, while oxygen is needed to scavenge the adsorbed hydrogen from the surface.

Sadana and katzer (1974) found that, phenol oxidation, involves an induction period, in which the generation of radicals is poor, followed by a higher steady state activity period with a free-radical reaction regime. These mechanisms are likely to occur even for noble metal catalysts.

Matatov-Meytal and Sheintuch (1998) reported that in wet oxidation, water with dissolved oxygen is used to oxidize the target compound. The main reactions are described in Eqs. (1)–(8). Hydroxyl radicals are produced from the dissociation and oxidations of water according to Eqs. (1) and (2). Hydroperoxyl radicals are formed from the oxidation of water (Eq. (3)) and the target compound RH (Eq. (6)). Hydroxyl radicals are also produced from hydrogen peroxide (Eq. (4)) and from the reaction of atomic oxygen with the target compound (Eq. (8)). Although the hydroperoxyl radical is less reactive than the hydroxyl radical, it plays an important role because of its relative abundance.

$$H_2O \rightarrow OH + H \quad (1)$$

$$H_2O + O_2 \rightarrow OH + H_2O \quad (2)$$

$$2H_2O \rightarrow H_2O_2 + O_2 \quad (3)$$

$$H_2O_2 \rightarrow 2OH \quad (4)$$

$$O_2 \rightarrow 2O \quad (5)$$

$$RH + O_2 \rightarrow R + HO_2 \quad (6)$$

$$RH + HO_2 \rightarrow R + H_2O_2 \quad (7)$$

$$RH + O \rightarrow R + OH \quad (8)$$

2. Kinetic Analysis of the Catalytical Oxidation of Phenol

Figure 1 illustrates the experimental apparatus used for the kinetic study of the oxidation of phenol. The reactions were performed in a four necked, 500-mL Pyrex flat-bottom flask. A distributor for gaseous oxygen was embedded in the vertical neck. Oxygen was sparged at a metered rate into the liquid phase through a coil containing 20 holes, each 0.2 cm in diameter. Temperature of the reaction mixture was monitored with a calibrated thermometer inserted in the second

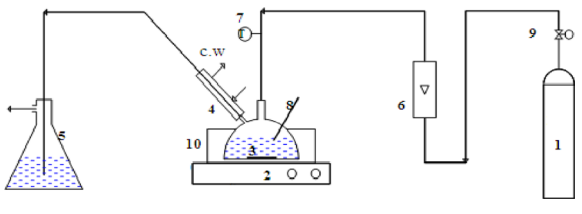

Fig. 1 Laboratory apparatus for the kinetics study. 1, oxygen bottle; 2, controlled electric heater with magnetic stirrer; 3, 0.5-L reaction vessel; 4, condenser; 5, 1 L-beaker with 0.5 M NaOH, 6, gas rotameter; 7, temperature indicator; 8, port for sampling and thermometer; 9, pressure regulator; 10, Jacket for reaction vessel

neck. Feeding and sampling tubes were housed by the third neck of the flask. A vapor condenser was connected to the fourth neck where the effluent vapors and gases were sent to a washing bottle containing 0.5 M NaOH solution where gaseous CO_2 produced by the reaction was absorbed and measured. The flask was heated with a heating mantel and stirred magnetically. A known weight of phenol and catalyst in powder form were loaded into the flask and heated to the desired temperature. The mixture was stirred, dispersing the catalyst uniformly throughout the liquid. Cooling water was started and oxygen was sparged into the flask. Samples were withdrawn periodically using a syringe and the catalyst was separated from the aqueous phase by a centrifuge. The samples were analyzed with gas chromatography (GC)–(ZB-FFAP), capillary column, 30 m long, 0.32 mm inside diameter, nitroterephalic acid modified ploy ethylene). A calibration curve was used to determine the phenol concentration in the reaction products in each run. Each run was repeated three times to reduce human and device errors to increase the precision. The absolute average error between each of three repeated experiments were in the range of (3–5%). In the kinetic analysis of the reaction, the rate equations proposed by Sadan and Katzer (1974) were tested. Equations (9) and (10) represent these equations for the initial rate and the rate in the steady state activity regime, respectively.

$$\frac{V_{\text{liq.}}}{W_{\text{cat.}}}\cdot\frac{dx}{dt}=k_1\frac{1-x}{\sqrt{W_{\text{cat.}}\dfrac{1}{V_{\text{liq.}}}}} \tag{9}$$

$$\frac{V_{\text{liq.}}}{W_{\text{cat.}}}\cdot\frac{dx}{dt}=k_2\frac{1-x}{\left(W_{\text{cat.}}\dfrac{1}{V_{\text{liq.}}}\right)3} \tag{10}$$

If $-\ln(1-x)$ is plotted verses time, the slope of the curve at $x=0$ and $t=0$ gives (k_1), while the slope in the steady state activity regime gives (k_2).

3. Experimental Setup

The primary components of the experimental apparatus were the catalytic falling-film reactor (CFFR), back mix reactor (CCSTR) with suspended catalyst, solution tank, return solution tank, heating system, pumping system and solution sampling system. **Figure 2** shows the compact experimental setup. The (CFFR) contains a concentric tube that defines the space in which the reaction takes place. The stainless steel inner tube (5) has a 2.9 cm OD, a total length of 1.6 m and an effective test section length of approximately 95 cm which was covered by a 3 mm thickness layer consisting of catalyst pellets (0.7% Pt on Al_2O_3) that were stacked on the outer surface of the tube using a special type of adhesive .The outer tube (6) has an 8 cm OD and was made of pyrex glass pipe permitting flow observation. The contaminated liquid solution was introduced to the distributor (4) at the top of the reactor and flows downward under gravity through the wetted catalyst and into the collector attached to the bottom of the reactor. In the reactor, it interacted with the upward flowing gaseous oxygen and reaction occurred. The contaminated solution was prepared with the desired concentration in a stainless steel tank (12) with a volume of 60 L equipped with an agitator and an electric heater. The (CCSTR) was a stainless steel tank with a volume of 10 L equipped with an agitator and a gas sparger. These mixing tools helped to keep the catalyst particles suspended in the solution flowing through the reactor. To prevent the catalyst

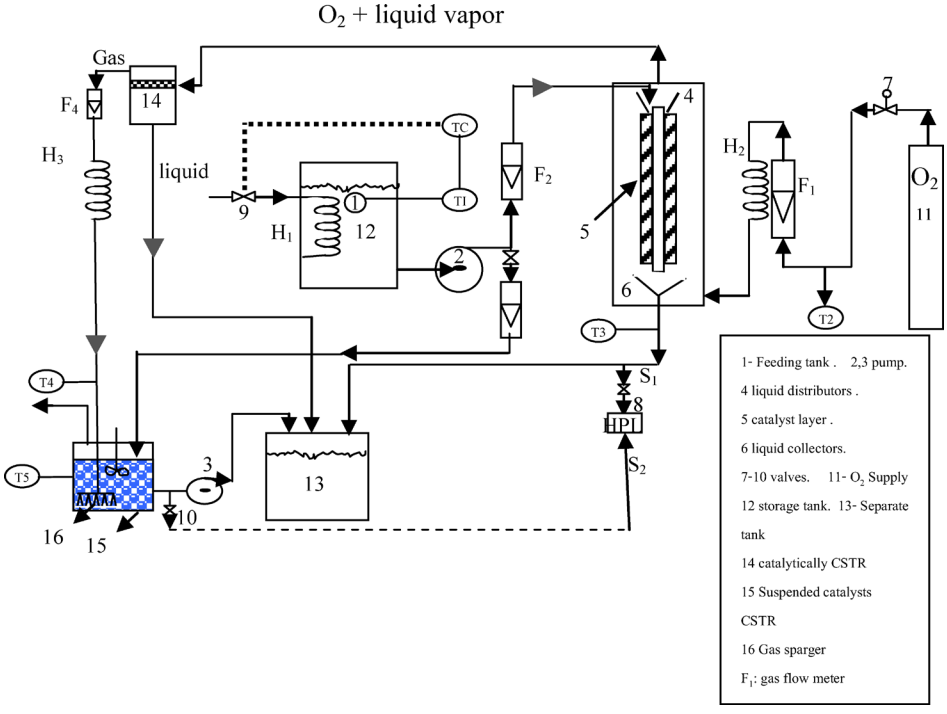

Fig. 2 Experimental set-up for the comparative study

Table 1 Characteristics of phenol, catalyst, and dimensions of reactors

Phenol, purity 99% from Bayer GmbH	
CFFR (OD/H) [m/m]	0.035/0.95
CSTR (OD/H) [m/m]	0.487/0.536
Composition of the catalyst (from the north refinery in Beji)	0.7% Pt/Al_2O_3
Density of catalyst particle [kg/m^3]	1030
Catalyst particle porosity	0.43
Catalyst particle diameter [mm]	2
Catalyst bed porosity	0.36
Packing specific surface area [m^2/m^3]	1.43×10^3

particles from escaping out of the reactor, a perforated stainless steel plate with 0.2 mm holes was installed at the outlet line of the reactor. The heated contaminated solution was discharged from the tank (12) by a pump (2) through a calibrated rotameter (F2) to the (CFFR) and (CCSTR) respectively. A compressed gas bottle (11) was used as a feeding source of pure oxygen which entered the (CFFR) at the bottom and flowed to the top. Drops of solution which carried by the gaseous oxygen out of the reactor were separated in a stainless steel separator (14). The oxygen continued to flow from the separator (14) to the (CCSTR) (15) while the separated liquid droplets were collected at a tank (13). The flow rates of pure oxygen fed to the (CFFR) and (CCSTR) were adjusted using calibrated flow meters (F1) and (F4). Samples were collected at equal time intervals using sample ports (S1 and S2) and analyzed using HPLC. Temperature managements was performed with K-type thermocouples (T1 to T5) calibrated to an of accuracy ±0.1°C. To compare the performance of the two different types of reactors, similar conditions were established by using the same operating parameters (i.e., weight of catalyst, inlet temperature and residence time of all reactants in each reactor). The characteristics of the phenol and catalyst and the dimensions of the reactors are presented in **Table 1**.

4. Experimental Designs

In the present work, factorial design method was used to plan the experiments because of its reliability in finding the effects of and interaction between the controlled variables of the system. The real value of the controlled variables (F) and their corresponding levels (L) are shown in **Table 2**.

The operating pressure and the weight of catalyst inside the two reactors were kept constant at 0.5 bars and 222 g respectively for all experimental runs. In addition, both reactors were operated with the same parameters shown in Table 2.

5. Results and Discussion

In **Figure 3** the linearity of the experimental data in the steady-state activity regime indicates first order kinetics with respect to phenol concentration. The rate constants k_1 and k_2 for the two runs shown in Figure 3 were found to be:

Table 2 Selected levels and factors

L \ F	Real variables			
	Phenol concentration [mg/L]	Liquid flow rate [$L\,min^{-1}$]	Gas flow rate [$L\,min^{-1}$]	Temperature [°C]
1	500	0.1	4	80
2	1000	0.4	8	90
3	1500	0.7	12	100
4	2000	1.0	16	110

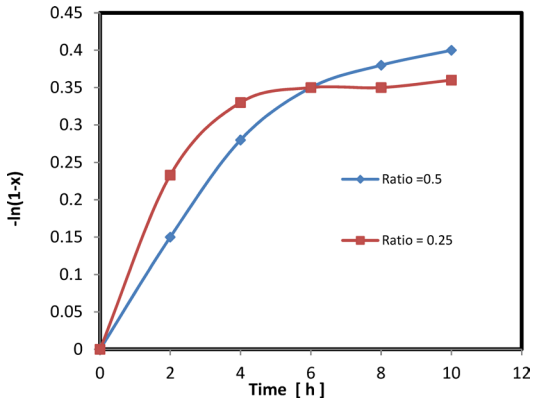

Fig. 3 $-\ln(1-x)$ vs. time for different phenol/catalyst ratio for kinetic analysis study $T=373$ K, $V_g=2\,L\cdot min^{-1}$

For phenol/catalyst ratio = 1/4 = 0.25

$$k_1 = 6.97\times10^{-4} \quad \left[\left(L_{liq.}/kg_{cat.}\right)^{0.5} s^{-1}\right]$$

$$k_2 = 3.84\times10^{-9} \quad \left[\left(kg_{cat.}/L_{liq.}\right)^{2} s^{-1}\right]$$

For phenol/catalyst ratio = 2/4 = 0.5

$$k_1 = 6.21\times10^{-4} \quad \left[\left(L_{liq.}/kg_{cat.}\right)^{0.5} s^{-1}\right]$$

$$k_2 = 7.63\times10^{-9} \quad \left[\left(kg_{cat.}/L_{liq.}\right)^{2} s^{-1}\right]$$

These experiments, exhibit a first order dependence on the phenol concentration in both the initial activity and steady state activity regimes. The effect of variations of the catalyst loading on the rate constants proved that the reaction involves a heterogeneous homogeneous free radical mechanics.

Figure 4 shows a proportional relationship between temperature and the rate of conversion of phenol. It can be seen from Figure 4 that the activity of the catalyst, which is expressed in terms of conversion, increases with temperature (Masende *et al.*, 2003). As can be seen, the final phenol conversion results obtained at temperatures of 353, 363, and 373 K for the studied catalyst were about (0.14, 0.24, and 0.3), respectively for all the samples. However, the time necessary to reach this steady state conversion (induction period) is lower for higher temperatures as shown in **Table 3**. The liquid effluent samples changed from colorless to light

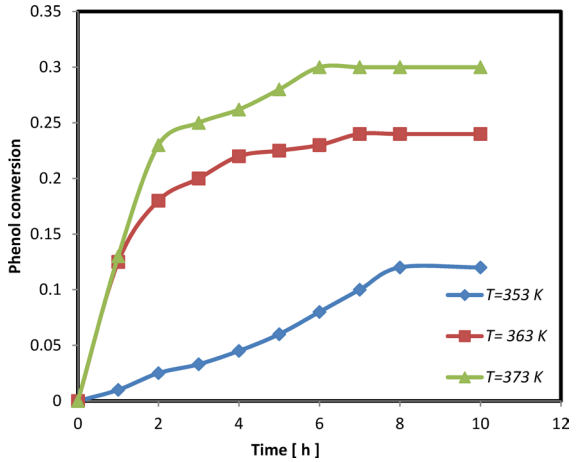

Fig. 4 Effect of temperature on conversion for kinetic analysis study, $V_g = 8$ L/min, ratio = 0.25

Table 3 Catalyst activity results for different temperatures

	353 K	363 K	373 K
Induction period [h]	8	6	4
Final phenol conversion	0.14	0.24	0.3

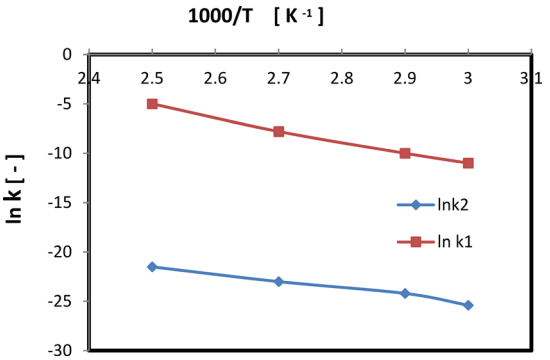

Fig. 5 Arrhenius plots, k_1 for the induction period, k_2 for steady state activity regime

yellow. The yellowish color in the liquid samples is possibly due to the formation of *p*-benzoquinone and polymer precursors.

In **Figure 5** the activation energies can be estimated from the slopes of the straight lines calculated by means of the least squares techniques around 353 K; k_1 and k_2 were 96 kJ/mol and 50 kJ/mol with R^2 values of 0.95 and 0.99, respectively. The mean relative deviations (MRD) between experimental points and the kinetic model were 6.2% and 3.4%, respectively.

The Arrhenius equation for k_1 and k_2 can be represented by:

$$k_1 = 1.8\times10^{10}\exp\left(-96/RT\right)$$

and

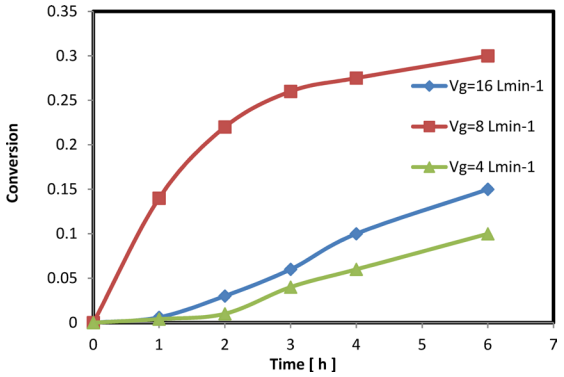

Fig. 6 Effect of gas feed rate on conversion at $T = 372$ K for kinetic analysis study, ratio = 0.25

$$k_2 = 2.4\times10^{-4}\exp\left(-50/RT\right)$$

The reported activation energies may even vary for the same compound and catalyst (Efftaxias, 2002). The literature activation energies of phenol oxidation in aqueous solutions above atmospheric pressure is substantial, but there is little information for atmospheric pressure. For phenol, at pressure above 10 atm most authors agree on values between 85 and 105 kJ/mol (Ohta *et al.*, 1980; Fortuny *et al.*, 1995). Conversely, Sadana and Katzer (1974) reported a much higher value of 176 kJ/mol, for the steady state period. In the present work, the activation energies were found to be 96 kJ/mol and 50 kJ/mol for the induction and steady state periods, respectively.

Figure 6 shows the effect of the oxygen feed rate on conversion of phenol after different amounts of time. The highest phenol conversion was obtained with an oxygen feed rate of 8 L min^{-1}. It is evident that phenol removal increases with increasing amounts of bubbled oxygen gas. A decrease in phenol conversion was observed for oxygen feed rate of 16 L min^{-1}. This can be attributed to the generation of larger gas bubbles as the gas flow rate was increased. These large bubbles have a higher rise velocity which results in a shorter contact time between the gas and liquid reactants. Consequently, a lower conversion rate was obtained.

Figure 7 shows the effect of the initial concentration on the phenol conversion in CWO runs over a 0.7 wt% Pt/Al_2O_4 catalyst at 383 K at a partial pressure of oxygen of 0.5 bars for the CFFR and CCSTR reactors. The removal of phenol was shown to decrease markedly as the liquid hourly space velocity (LHSV) of the influent stream increased, because the retention time during the CWO process was reduced. It can be seen that at lower values of the LHSV, the conversion was higher for the CFFR system than that for the CCSTR. This was attributed to the effect of the ratio of surface area to volume liquid which is higher for the CFFR system. This ratio is proportional to the rate of oxygen diffusion and reaction in the liquid phase. As the LHSV continued to increase, the phenol conversion of the CCSTR became higher than that of the CFFR. This can be attributed to the effect of back mixing which becomes larger in the CCSTR system

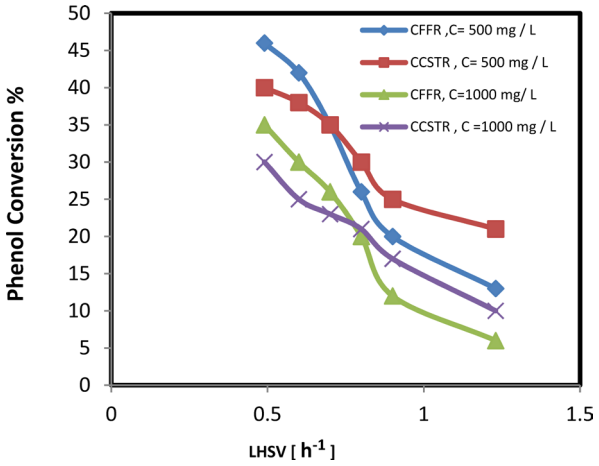

Fig. 7 Effect of initial concentration on phenol removal at $T = 383$ K

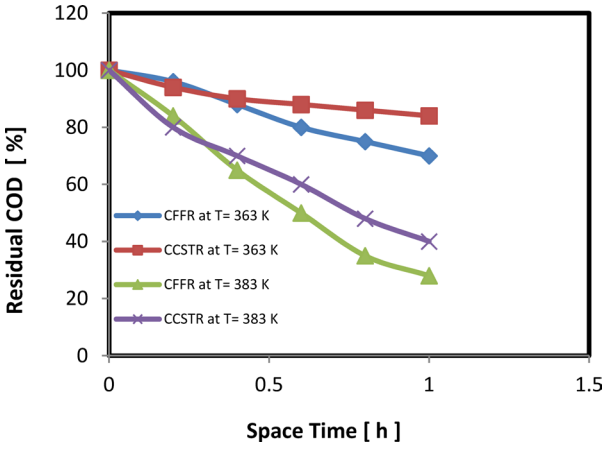

Fig. 9 Effect of space velocity and temperature on residual COD

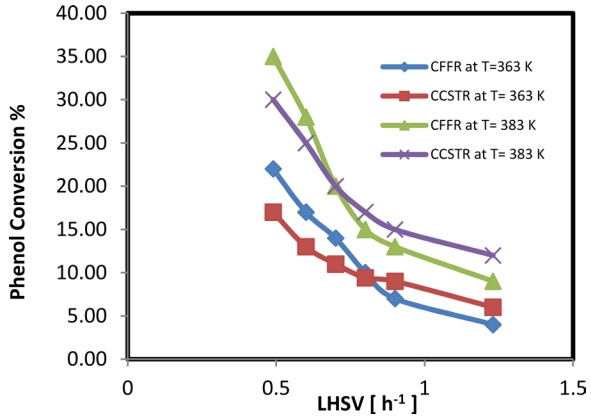

Fig. 8 Effect of reaction temperature and LHSV on phenol removal

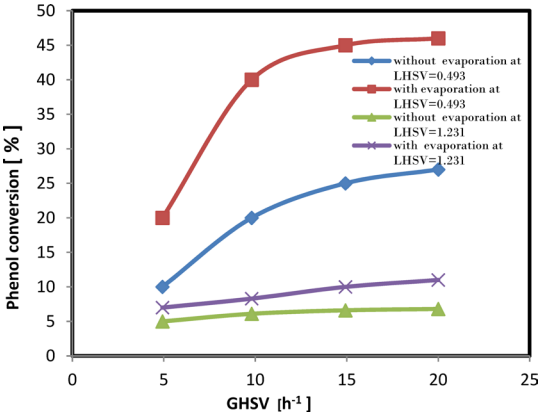

Fig. 10 Phenol conversion as a function of GHSV

and allows it to surpass the CFFRs.

Figure 8 illustrates the effects of reaction temperature and the space velocity of the influent stream on the conversion of phenol. An increase in the reaction temperature was observed to result in increased removal of phenol. The effect of temperature was to increase the specific rate constant which increased the conversion. The CFFR and CCSTR exhibit different levels of performance both with increasing reaction temperature and with increasing space velocity. The latter has an adverse impact on conversion for the same reason mentioned in the discussion of Figure 7.

Figure 9 plots the effect of temperature on the chemical oxygen demand (COD) for the samples taken from the effluents of the CFFR and CCSTR at the same intervals of time. The operating conditions of the two reactors were the same. The figure shows that the residual COD of the effluents from the two reactors decreased as the reaction temperature increased. The experimentally measured reaction rates indicate that the laboratory scale reactors operate in the kinetic control regime when compared with mass transfer rates calculated from available correlations (Iliuta *et al.*, 1999).

Figure 10 shows the effect of water evaporation on Phenol conversion at different values of the gas hourly space velocity (GHSV). As can be seen, the water evaporation can have a substantial influence on the CFFR and CCSTR performance. The plots are parameterized for two LHSVs, corresponding to high and low phenol conversions. For lower values of LHSVs, a higher conversion profile was obtained under evaporation conditions. For higher LHSVs, the lowest conversion profile was observed without evaporation. The figure shows a positive impact of GHSV on phenol conversion, with a steeply increasing rate of conversion at low values of GHSV. Meanwhile the rate of change of conversion is significantly lower at higher values of GHSV. The plot also shows the same performance trend for both the CFFR and CCSTR. As can be seen the phenol conversion of the CFFR is higher than that of the CCSTR at lower values of LHSV (≤ 0.6) with and without evaporation, while the results are completely different at higher values of LHSV (≥ 1.2). This may be suggested that operating the reactors at the saturation temperature and equivalent pressure would be useful for obtaining higher performance.

6. Mathematical Correlations

Based on the experimental results the following power

law based correlations were proposed. The coefficients were estimated using regression analysis technique.

For CCSTR:

$$x = 0.502\left(\Delta T / T\right)^{-0.026}\left(\Delta C / C\right)^{0.267}\left(\text{LHSV} / \text{GHSV}\right)^{-0.018}$$

(Absolute average relative error = 1.23%, and standard of deviation = 1.1%)

For CFFR:

$$x = 0.426\left(\Delta T / T\right)^{-0.152}\left(\Delta C / C\right)^{0.668}\left(\text{LHSV} / \text{GHSV}\right)^{-0.020}$$

(Absolute average relative error = 1.16%, and standard of deviation = 1.18%)

$$\text{for} \quad 0.4 < \text{LHSV} < 1.3 \quad [\text{h}^{-1}]; \quad 5 < \text{GHSV} < 20 \quad [\text{h}^{-1}]$$
$$500 < C < 2000 \quad [\text{mg/L}] \quad \text{and} \quad 80 < T < 110 \quad [^{\circ}\text{C}]$$

Conclusion

The present study was divided into two sections. In the first, a kinetic analysis of the oxidation of phenol in aqueous solution over a supported (0.7% Pt)/Al_2O_3 catalyst was investigated at atmospheric pressure in a batch operating system.

The kinetic analysis proved that the reaction consists of two mechanisms, that the initial rate and steady state activity regimes exhibited first order behavior with respect to phenol concentration. The reaction rates show an unusual dependence on catalyst loading which suggested a heterogeneous – homogenous free radical mechanism. The rate constants k_1 and k_2 for the initial rate and steady state activity regime are given by:

$$k_1 = 1.8\times10^{10}\, e^{(-96/RT)}$$

$$k_2 = 2.4\times10^{-4}\, e^{(-50/RT)}$$

Phenol removal may be increased by increasing the amount of oxygen gas but at higher flow rates of oxygen a retarding effect of oxygen on phenol oxidation was observed.

In the second section, a comparative study of the CWO of phenol in two different types of flow reactors (i.e., falling film and back mixing reactors) was carried out and design parameters such as inlet temperatures, the residence time of reactants and catalyst loading in the reactors were used to establish conditions in the two reactors. The study supports the following conclusions; The oxidation rate of phenol was low because of the solubility of oxygen under atmospheric conditions. At low flow rates of liquid reactant the falling film reactor showed a better performance as a result of its lower resistance to mass and heat transfer while the results are completely different at higher liquid flow rates. Non-isothermal operation showed that water evaporation has a strong impact on phenol conversion and must be taken into account in scale up and adiabatic CWO reactor design. Neglecting evaporation can lead to erroneous calculation of the exit stream phenol conversion and temperature. The power law technique has been utilized to correlate the phenol conversion with the operating parameters in the two reactors respectively. The correlation factor was 0.97 for the CFFR and 0.96 for CCSTR.

Acknowledgment

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Nomenclature

C	=	concentration of phenol	[mol/cm^3]
CCSTR	=	catalytic continuous stirred tank reactor	
CFFR	=	catalytic falling film reactor	
CWO	=	catalytic wet oxidation	
GHSV	=	gas hour space velocity	[h^{-1}]
HPLC	=	high performance liquid chromatography	
k_1	=	rate constant for initial rate	[L of liq./kg of cat.]$^{0.5}$ [s^{-1}]
k_2	=	rate constant for steady state activity	[(kg cat./L liq.)^2 s^{-1}]
LHSV	=	liquid hour space velocity	[h^{-1}]
R	=	universal gas constant	[kJ/mol K]
T	=	temperature	[K]
t	=	time	[s]
V	=	liquid-phase volume	[L]
V_g	=	gas flow rate	[L/min]
$V_{liq.}$	=	volume of the liquid	[L]
x	=	conversion of phenol	[—]
W	=	weight of the catalyst	[kg]

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